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CRANFIELD INSTITUTE OF TECHNOLOGY

**DEPARTMENT OF FLUID ENGINEERING AND
INSTRUMENTATION**

PhD THESIS

Academic Year 1986-7

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**Gas-liquid dispersion in stirred tanks: scale-up and
agitator comparison**

Supervisor:

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August 1987

**This thesis is submitted for the degree of
Doctor of Philosophy**

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ABSTRACT

This thesis is a description of an experimental study of gas-liquid dispersion in mechanically agitated vessels. The objectives of the project were to investigate scale-up and to compare different agitator types. Three vessels up to 2.67m in diameter were available for test work.

The work consisted largely of the measurement of the following parameters:

- (i) Impeller power demand under gassed and ungassed conditions.
- (ii) Overall gas hold-up.
- (iii) Mass transfer coefficient using a dynamic response technique.
- (iv) Flooding point, especially for Rushton turbine impellers.

The influence of scale effects on mass transfer has been discussed in some detail from a theoretical standpoint, especially the influence of liquid depth on the concentration driving force available for mass transfer.

The use of specific power input and superficial gas velocity was adequate to correlate both hold-up and mass transfer coefficient, but some second order scale effects appeared to be present which made the correlations in each vessel statistically significantly different from each other.

The Rushton turbine showed little difference in dispersion efficiency from the concave turbine. Differences were, however, discovered between pitched blade turbines and Rushton turbines, especially when the pitched blade turbine was pumping upwards.

The economics of mixer design has also been considered, and the

potential economic value of academic research into energy efficiency has been clearly demonstrated.

ACKNOWLEDGEMENTS

The author would like to recognise the contribution to this work made by a fairly long list of people.

Firstly, I would like to thank Professor Alvin Nienow of Birmingham University, who has acted as the technical supervisor, and without whose help and advice this work may never have come to fruition. I also acknowledge the contribution of the other members of the support panel who have faithfully met on a regular basis for the last three years to review progress and keep me on the right track:

Dr K Enever	)	
Dr M Sanderson	)	Cranfield Institute of Technology
Dr I Black	)	
Dr R King	)	
Mr M Palmer	)	BHRA
Mr C Armstrong	)	

Of my many co-workers at BHRA, Mike Whitton, Alastair Sinker and Nigel Pendall deserve special mention for their assistance with various aspects of the experimental work. Many workshops and support staff at BHRA have been involved in the work, but particular thanks to the typists who have struggled bravely with my scribbles. There are Karen Sweeting, Shiela Newman and Val Holden. Thanks also to Maureen Battrum and Brenda Pyrah for their assistance with diagrams and production of the final thesis.

It would be remiss of me to overlook the contribution of the FMP Steering Committee who have reviewed this work (and other data in the same field) at six monthly intervals. There have been many useful discussions at the meetings, and I am not going to single out individuals.

More importantly, perhaps, I ought to thank all the member companies of the FMP consortium who have sponsored the research programme of which this thesis forms a part.

Last but by no means least, I must thank my wife, Audrey, for her heroic efforts in proof reading the text. It is to her that this thesis is dedicated.

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NOMENCLATURE

Please note that in view of the large numbers of symbols used in Chapter 6 (Mass Transfer) and Chapter 8 (Mixer Economics), separate Nomenclatures for these chapters can be found at the end of the text for that chapter.

These "local" Nomenclatures take precedence over the "global" Nomenclature listed below where the same symbol has been used for two different quantities. Whilst reading Chapters 6 and 8, the reader's first recourse should therefore be to the "local" Nomenclature.

SYMBOL	MEANING	UNITS
a	Interfacial area per unit volume	m^2/m^3
A	Dimensionless constant in equation (2.16)	-
B	Dimensionless constant in equation (2.16)	-
C	Concentration of ionic solute in bulk liquid	kmol/m^3
C*	Concentration which would be in equilibrium with other phase	kmol/m^3
C _B	Concentration of non gaseous reactant in bulk liquid	kmol/m^3
C _{Gi}	Concentration of gas entering vessel	kmol/m^3
C _{Go}	Concentration of gas leaving vessel	kmol/m^3
C _{Li}	Liquid concentration entering vessel	kmol/m^3
C _{Lo}	Liquid concentration leaving vessel	kmol/m^3
D	Impeller diameter	m
d _{sv}	Sauter mean diameter of bubbles	m
D _A	Diffusion coefficient of gas in liquid	m^2/s
e _g	Hold-up in two-phase region of tank	
f	Activity coefficient	
g	Acceleration due to gravity	m/s^2
h	Characteristic dimension (liquid depth)	m

H	Fill depth of tank	m
H_0	Static head at tank outlet	m
k	Reaction rate constant (first order reaction)	$\text{m}^3/\text{kmol s}$
K	Equilibrium constant (= gas concentration/liquid concentration)	
k_L	Liquid side mass transfer coefficient	m/s
$k_L a$	Mass transfer coefficient	s^{-1}
K_4	Dimensionless constant	
K_6	Dimensionless constant	
L	Liquid flow rate	m^3/s
M	Relative molecular weight of gas	
N	Impeller rotational speed	rev/sec
n_B	Number of impeller blades	
N_L	Impeller speed at transition point	rev/sec
N_0	Minimum speed	rev/sec
p	Hydrostatic pressure	N/m^2
P_g	Gassed power dissipated by agitator	W
P_u	Ungassed power dissipated by agitator	W
Q	Gas flow rate	m^3/s
Q_I	Liquid flow rate in impeller outlet stream	m^3/s
Q_L	Gas flow rate at transition point	m^3/s
Q_{VVM}	Gas flow rate (vessel volumes per minute)	
R	Universal gas constant	J/kmol K
R_V	Mass transfer per unit volume	$\text{kmol/m}^3\text{s}$
T	Tank diameter	m
T_a	Absolute temperature	K
t_G	Gas residence time	s
U_s	Tip speed	m/s
U_t	Bubble rise velocity	m/s

v	Liquid velocity	m/s
V	Volume of dispersion	m^3
V_G	Volume of gas in dispersion	m^3
V_L	Volume of liquid in dispersion	m^3
V_R	Reactor volume	m^3
v_{SG}	Superficial gas velocity	m/s
w	Impeller blade width (absolute not projected)	m
x_1	Disc thickness	m
x_2	Blade thickness	m
α	Regression coefficient (dimensional)	
β	Regression coefficient	
γ	Regression coefficient	
$\dot{\gamma}$	Shear rate	s^{-1}
Γ_T	Ionic strength	
δ	Regression coefficient	
$\bar{\epsilon}_g$	Specific power input by gas	W/kg
ϵ_G	Gas hold-up	
$\Delta\epsilon_G$	Error in estimate of gas hold-up	
$\bar{\epsilon}_m$	Impeller power input per unit mass of liquid	W/kg
$\Delta\bar{\epsilon}_m$	Error in estimate of specific power input	W/kg
θ	Blade angle to horizontal	degrees
μ	Liquid viscosity	Ns/m^2
μ_{APP}	Apparent liquid viscosity	Ns/m^2
ρ	Density	kg/m^3
ρ_a	Air density at inlet conditions	kg/m^3
ρ_g	Gas density at inlet conditions	kg/m^3
ρ_L	Liquid density	kg/m^3
ρ_{LM}	Molar density of liquid	$kmol/m^3$

σ	Surface tension	N/m
τ	Torque on impeller shaft	Nm
τ_L	Response time of oxygen probe	s
τ_R	Liquid residence time	s

Dimensionless Groups

$$Fl = \frac{Q}{ND^3} = \text{flow number}$$

$$Fr = \frac{N^2 D}{g} = \text{Froude number}$$

$$Po = \frac{P}{\rho N^3 D^5} = \text{power number}$$

$$Re = \frac{\rho N D^2}{\mu} = \text{Reynolds number}$$

$$We = \frac{\rho N^2 D^3}{\sigma} = \text{Weber number}$$

Subscripts

CD	Complete dispersion
F	Flooding
g	Gassed
m	Molar quantity
R	Re-circulation

1. INTRODUCTION

The topic of gas-liquid dispersion in stirred vessels has received a great deal of attention in the literature. Industrially, there are many processes or reactions which depend crucially on gas-liquid mass transfer. This industrial significance is reflected in the interest in the gas-liquid mixing area of BHRA's commercially-sponsored Fluid Mixing Processes project, and, perhaps less directly, in the significant activity amongst university-based researchers. Some examples of commercially important processes are:

- 1 Fermentations. Many fermentation reactions are carried out aerobically in stirred vessels. The supply of oxygen to the micro-organisms is frequently a rate-limiting step. Many pharmaceutical processes depend on aerobic fermentation to produce the active ingredient of medical products. Consequently, the added value as a result of the fermentation step is high, providing a strong motivation to reach a fundamental understanding of the mechanisms of gas liquid dispersion with a view to improving production methods.
- 2 Hydrogenations. Many hydrogenations are actually 3-phase reactions, with trace amounts of solid catalyst present. The kinetics of the reactions in the liquid phase may be highly complex, but the mass transfer step is often the overall rate-determining step. Examples of hydrogenation are the conversion of nitro-benzene to aniline, which is an important step in the dye-making process, and the conversion of nitrile to fatty amines.
- 3 Oxidations. These frequently involve the oxidation of organic liquids by air. Jacobi and Baerns (1983) studied the oxidation of p-xylene to terephthalic acid, which under certain conditions is limited by oxygen transfer to the liquid. Terephthalic acid is an important intermediate in the polymer industry. Sridhar and Potter (1986) reported on

the oxidation of cyclohexane in a stirred tank, commenting that mass transfer may affect the selectivity, as well as the overall rate of reaction. Selectivity is very important for complex reaction networks, as shown by Mann, Middleton and Parker (1977) for an example of scale-up of a commercial process. Wachi and Morikawa (1986) considered mass transfer effects on selectivity of chlorination reactions.

- 4 There are many other gas-liquid reactions outside the three generic categories just described, such as sulphonations, chlorinations, etc.

These examples illustrate the commercial importance of the correct design of gas-liquid stirred tank contactors. There is a current trend to seek alternative techniques (Leuteritz, Relmann and Vergeres, 1976), such as the Buss loop reactor which uses a liquid driven jet ejector to contact the gas with the liquid, to supplant the traditional stirred vessel. BHRA has recently completed a study on a similar type of system for a hydrogenation reactor (Muskett and Armstrong, 1987) but these developments are relatively new and unproven: there are many stirred tank reactors in commercial operation, and the prospect of improved performance through retrofitting is an attractive one. Successful modifications depend on a good understanding of the dispersion process and there is still a strong motivation for research into stirred tank contactors. For some systems, especially highly viscous fermentations, it is difficult to envisage a successful loop reactor design. The requirement for gas-liquid contacting in stirred vessels is therefore likely to continue into the foreseeable future, although alternative methods may well take an increasingly important place in the market.

The nature of the processes carried out in gas-liquid contactors reinforces the economic incentive to improve somehow the efficiency of the process. In almost all cases, the final product, in diluted form, or an important intermediate is formed in the gas-liquid reactor. The overall process performance may be crucially dependent

on the performance of the reactor, so although gas-liquid stirred tanks are rather scarce numerically, their economic significance is great. Mixer manufacturers often estimate that as little as 1% of their total sales are for gas-liquid mixing applications: this figure is, of course, very dependent on the nature of the market addressed by each individual manufacturer, specifically the degree of concentration of the business in the fermentation industry.

Having established the industrial significance of the topic discussed in this thesis, it is natural to question whether or not the existing literature provides sufficient information to satisfy the requirements of the designer. The answer to this question for specific applications is almost universally negative. The rate of mass transfer depends not only on hydrodynamics and physical properties but on the presence of trace contaminants, as shown clearly by Warmoeskerken and Smith (1978). It is extremely hazardous to attempt to design a commercial scale reactor for a critical application without small scale experiments as a basis for design unless very large safety margins are allowed which could easily make the design (and possibly the whole process) uneconomic. Even with the benefit of small scale experiments, reliable scale-up procedures are required for a safe design. Scale-up requires the correct selection of impeller speed and size and gas flow rate based on small scale data. It is necessary to estimate the gassed power demand to size the motor (in some cases) or to estimate the mass transfer coefficient and gas hold-up if power per unit volume is used as a correlating parameter.

A common criticism of the literature is that too many papers have concentrated on small scale experiments. Some of the disastrous consequences of extrapolating the correlations derived from small scale data are discussed in a later section, and underline the need for experimental data at scales approximating to production vessels. Some papers, especially in the last five years, have been published to fill this gap, but the need for experimental data at industrial scales, even if only air-water systems could be studied, was apparent at the beginning of this project. In particular, data

was required which would be as reliable and accurate as the small scale data. Too often experiments at large scale have been carried out in production vessels which have, for instance, necessitated the measurement of impeller power using the power drawn by the motor. Such expediency can negate the value of large-scale data. Furthermore, the study of gas-liquid dispersion in stirred tanks is very much an empirical one (as pointed out by Mann in 1983 in highlighting the remarks of Valentin made in 1967!). This is emphasised by the fact that despite many papers on the subject, the prediction of the gassed power demand of a Rushton turbine is still fairly unreliable. If a fairly basic parameter such as power cannot be accurately predicted, it leads one to doubt the prediction of complex phenomena such as mass transfer.

Apart from scale-up, the situation may arise where gas-liquid dispersion may not be carried out using a Rushton turbine, the impeller type on which the overwhelming majority of published studies have concentrated. The chemical engineer in industry may be faced with the necessity to use, for example, a pitched blade turbine for dispersion of gas. Some work on the influence of impeller type has been published relatively recently (for example, Warmoeskerken, Speur and Smith, 1984, and Chapman, Nienow, Cooke and Middleton, 1983) but further data is required, particularly to compare pitched blade turbines with the more traditional Rushton turbine. As a side-effect of this type of impeller comparison, there is the desire to design an agitator which gives real improvements in dispersion efficiency than the Rushton turbine. This might almost be described as the "Holy Grail" of gas-liquid mixing, and a considerable amount of effort has so far failed to produce an impeller with real improvements in terms of energy efficiency. To settle, as it were, for second best, it is still desirable to use impellers with flat gassed power curves for practical reasons, but which produce the same amount of mass transfer per unit power input as the Rushton turbine.

This thesis is a description of an experimental study of gas-liquid dispersion in low viscosity systems using three main types of

agitator in three vessel sizes up to 14m^3 volume, 2.67m diameter. The measurement of gassed power, hold-up and mass transfer is described and discussed in each of three separate chapters. The themes of scale-up and agitator comparison have been explored for each of these three parameters, but other phenomena, such as the differences between coalescing and non-coalescing liquid media and the relationship between hold-up and mass transfer coefficient have also been discussed.

Some measurements of the flooding point of disc turbine agitators at large scale have also been made and compared with the predictions of published correlations.

Particular attention has been paid to the methods of measurement, especially in mass transfer, where scale effects should be considered in deriving a rigorous analysis technique.

There is frequently controversy between research workers in the field of mixing and suppliers of mixing equipment in particular about the practical value of research directed at reducing the energy costs associated with a particular mixing task. There is a trade off between running costs, which are dependent on energy consumption, and capital cost, which depends on the torque produced by the agitator. Torque tends to increase for low energy input processes. An economic comparison has been made for an example based on miscible liquid blending between mixer designs capable of achieving the same process result at different energy inputs. In gas-liquid mixing, such a simple trade off does not exist, but a study has been made of a hypothetical process with gas-liquid mass transfer as the rate-determining step which is the plant bottleneck. This economic analysis considers how the use of knowledge gained from research might be used to improve throughput and profitability.

2. LITERATURE SURVEY

The published literature on gas-liquid dispersion in stirred tanks is extremely extensive. Several reviews of the literature have been published, of which the most recent are those of Charpentier (1982), Joshi, Pandit and Sharma (1982), Mann (1983), and Midoux and Charpentier (1984). This literature survey concentrates on aspects of hydrodynamics and dispersion performance in low-viscosity liquids as a background to the work presented in the rest of the thesis. For reasons of compactness, the survey is not intended to be rigorously exhaustive, but concentrates on recent trends in the literature.

2.1 DEFINITION OF DIMENSIONLESS GROUPS

In the following discussion, several dimensionless groups have been used, which are defined as follows :-

(i) Flow Number

$$F1 = \frac{Q}{ND^3} \quad (2.1)$$

The flow number is intended to represent the ratio of gas flow rate Q , to liquid flow rate out of the impeller. For single phase flow :

$$N_Q = \frac{Q_I}{ND^3} \quad (2.2)$$

If N_Q is constant, as is the case for high Reynolds numbers, the liquid flow rate is proportional to ND^3 , hence $F1$ represents the ratio of gas flow to impeller outlet flow provided that the impeller pumping capacity is unaffected by gassing, which is extremely unlikely. Revill (1982) has given a detailed review of the pumping capacity of Rushton turbine agitators.

(ii) Froude Number

$$Fr = \frac{N^2 D}{g} \quad (2.3)$$

The Froude number represents the ratio of inertial forces to gravitational forces. This form of Froude number is used only for stirred tanks, and is a modification of the original Froude number defined for free-surface flows :-

$$Fr' = \frac{v}{\sqrt{gh}} \quad (2.4)$$

The use of Froude number as defined by equation (2.3) may be criticised when considering the fundamental arguments behind the definition of equation (2.4), but equation (2.3) has been widely used (and, implicitly, accepted) in the literature.

(iii) Power Number

$$Po = \frac{P}{\rho N^3 D^5} \quad (2.5)$$

The power number is the dimensionless form of the impeller power input to the liquid. It is a modified drag coefficient, and has been very widely used in the literature. The gassed power number, Po_g , is also frequently used in the literature. This is defined as:

$$Po_g = \frac{P_g}{\rho N^3 D^5} \quad (2.6)$$

Note that the liquid density is used in the definition rather than a mean density accounting for the reduction in density caused by the presence of gas.

(iv) Reynolds Number

$$Re = \frac{\rho N D^2}{\mu} \quad (2.7)$$

The Reynolds number represents the ratio of inertial to viscous

forces. It has been widely used to correlate power number (Rushton, Costich and Everett (1950), and Bates, Fondy and Corpstein (1963)).

(v) Weber Number

$$We = \frac{\rho N^2 D^3}{\sigma} \quad (2.8)$$

The Weber number represents the ratio of inertial to surface tension forces. Note that an impeller Weber number, as opposed to a bubble Weber number is generally used in correlations.

2.2 GAS CAVITIES : RUSHTON TURBINES

Before considering the process parameters of power consumption, gas hold-up and mass transfer, the fundamentals of the flow field within a gas-liquid stirred tank must be addressed. Much work has been published on the gas cavity structure behind the blades of Rushton turbines and pitched blade turbines.

A full description of the cavity formation sequence behind Rushton turbine blades has been given by Warmoeskerken (1986), following on from earlier work by van 't Riet (1975), Bruijn, van 't Riet and Smith (1974) and Nienow and Wisdom (1974).

A major feature of the flow from a disc turbine stirrer is the pair of roll vortices which trail behind each blade, as described by van't Riet and Smith (1975) and van't Riet, Bruijn and Smith (1976). When gas is introduced to the system, bubbles are drawn into the low pressure regions at the centre-line of each vortex to form clinging cavities (Fig 2.1). The pairs of clinging cavities grow as gassing rate is increased further, and provided that the impeller Froude number exceeds 0.045 (Warmoeskerken and Smith, 1982) three of the six pairs of clinging cavities, almost simultaneously become large cavities (Fig 2.1). The new structure consists of alternate large cavities and clinging cavities provided that the number of

impeller blades is even (Warmoeskerken, 1986).

The reasons for the formation of the alternating large cavity/clinging cavity structure (the "3-3" structure for six-bladed agitators) are unclear, but the structure is very stable. The cavities continue to grow as gassing rate increases further, and the 3 pairs of clinging cavities become large cavities, which remain somewhat smaller than the original large cavities.

If the impeller speed is high enough, 3 very large cavities form which extend from the rear of a blade to the front of the blade behind it. This condition was not encountered for the experiments described in this thesis.

At very high gassing rates, the large cavities change to ragged cavities described by Nienow, Allsford and Kuboi (1983). At this point the agitator is flooded (see Section 2.3).

2.3 FLOW REGIMES : RUSHTON TURBINE

The two-phase flow in a stirred tank can be divided into a series of flow regimes. Not all of the flow regime boundaries can be linked to changes in cavity structure. The flow regime transitions are considered in the approximate order in which they would be expected to occur at constant impeller speed and decreasing gas rate (Fig 2.2).

(i) Flooding

The definition of flooding was disputed for many years until the recent work of Nienow, Konno, Warmosekerken and Smith (1985), following on from the work of Warmoeskerken and Smith (1984 and 1985).

The following correlation has been shown to be valid for disc turbines, provided the sparger is of a smaller diameter than the

impeller :

$$Fl_F = 30 \left(\frac{D}{T} \right)^{3.5} Fr_F \quad (2.9)$$

The form of the equation was derived by considering the relative flow rates of liquid entrained by the gas and liquid in the impeller outlet stream. This is not thought to be a very rigorous theoretical basis, but an equation has been produced which works well in practice. Bujalski (1987) justified the form of equation (2.9) in terms of the balance between radial forces on the bubble as a result of the action of the agitator and bouyancy.

The flooding/loading transition may be detected in several ways.

- (a) Changes in the impeller outlet velocity with increasing stirred speed (Roustan and Bruxelmane (1981), Warmoeskerken and Smith (1984), and Lu and Chen (1985)).
- b) Step changes in the gassed power curve at the flooding point (Nienow et al (1985)).
- c) Changes in the spectra of sound generated in the vessel, as used by Hsi, Tay, Bukur, Tatterson and Morrison (1985).
- d) Vane sensors in the impeller outlet stream (Warmoeskerken, 1986).

The step change in power consumption at flooding is caused by the smaller attached area of ragged cavities compared to large cavities. The flow over the blade becomes less streamlined and the power consumption increases. It is also possible at low Froude number, to have a transition directly from clinging cavities to ragged cavities, producing a step fall in power consumption. This is quite common with large impeller-to-tank diameter ratios.

The sound spectra technique of Hsi et al (1985) requires considerable interpretation of the data, but could be useful in situations where power or velocity measurements were not possible.

The form of equation (2.9) may be seen in the data of Judat (1977), who produced a very complex correlation when a straight line fit would appear to have been adequate, especially in view of the experimental errors involved.

The correlation holds for Rushton turbine agitators and does not appear to be sensitive to geometric factors such as impeller bottom clearance. Sparger type is also of little significance, as shown by Warmoeskerken (1986), for small ring spargers and pipe spargers. Nienow and Allsford (1985) found that for large ring spargers ($D_s > D$) placed close to the agitator, flooding took place at higher gassing rates than for conventional spargers.

(ii) Clinging/"3 - 3" Cavity Transition

The transition from clinging cavities to the "3 - 3" structure is important for the following reasons :

- a) There is a very sharp fall in the power curve.
- b) Authors such as Davies (1986) and Warmoeskerken (1986) have argued that since the flow regime in the impeller region, and thus the dispersion mechanism, is so different for each cavity structure, the gas hold-up and mass transfer coefficient should be treated separately for clinging cavities and the "3 - 3" structure. The cavity transition marks a fundamental change in the gas dispersion mechanism.

The transition has been detected either from the point of maximum slope of the gassed power curve, or by the spectral analysis of the output from a vane sensor placed in the impeller

outlet stream (Smith and Warmoeskerken, 1985). The clinging cavity structure produces a peak at 6 times the impeller rotational speed (for a 6-bladed agitator) whilst the "3 - 3" structure also produces a peak at 3 times the rotational speed. This elegant technique allows the transition point to be detected.

Smith and Warmoeskerken (1985) have given the following correlation for the transition :-

$$Fl = 3.8 \times 10^{-3} \left(\frac{Re^2}{Fr} \right)^{0.067} \left(\frac{T}{D} \right)^{0.5} \quad (2.10)$$

For constant liquid physical properties and geometrical similarity, equation (2.10) reduces to :-

$$Fl = A D^{0.2} \quad (2.11)$$

A = dimensional constant depending on geometry and physical properties.

There is a clear scale effect, with the clinging to "3 - 3" transition taking place at higher flow numbers as scale increases. The liquid physical properties in equation (2.10) were only included to make the equation dimensionless and were not actually varied in the experiments.

Warmoeskerken (1986) tested equation (2.10) for a non-coalescing medium and found it to be valid to a reasonable degree of accuracy.

(iii) Complete Dispersion

The point at which gas bubbles are dispersed below the agitator has been defined as a flow regime transition by Nienow, Wisdom and Middleton (1977). This is the complete dispersion condition, and the following correlation holds for coalescing liquids, as given by Nienow and Wisdom (1978) from measurements

derived from gassed power curves:

$$\left(\frac{N_{CD}^2 D}{g}\right)^{\frac{1}{2}} \left(\frac{D}{T}\right)^{\frac{1}{2}} = 5.06 \left(\frac{Q_{CD}}{N_{CD} D^3}\right) \quad (2.12)$$

(iv) Re-circulation

If the impeller speed becomes sufficiently high, gross re-circulation of gas into the impeller region occurs, which influences the gas phase residence time distribution, tending to produce a well-mixed gas phase.

Re-circulation has been investigated in various ways :-

- a) A peak in the gassed power curve when keeping Q constant and varying N (Nienow et al, 1977).
- b) Comparison of gassed power curves for geometries with and without re-circulation (Van 't Riet et al, 1976).
- c) The response of the impeller power demand to step changes in gas flow rate (Greaves and Economides, 1979) and to oscillations in the gas flow rate (Greaves, Kobbacy and Millington, 1983).
- d) Dynamic gas hold-up after abrupt cessation of gassing (Mann, Mavros and Middleton, 1981).
- e) Sampling and analysis of gas in the cavities when using a conventional sparger feeding nitrogen and a sparger in the impeller outlet stream feeding oxygen (Nienow, Middleton and Chapman, 1979). This produced results in reasonable agreement with van 't Riet et al (1976) but it was not possible to detect the onset of re-circulation because some oxygen was apparently drawn into the impeller region even when the impeller was flooded.

Overall, the information available on re-circulation is rather

sketchy. The only correlation for predicting the onset of re-circulation was given by Nienow et al (1977) :-

$$\left(\frac{Q_R}{N_R D^3}\right)^{-\frac{1}{2}} \left(\frac{N_R^2 D}{g}\right) \left(\frac{D}{T}\right)^{2.5} = 0.28 \quad (2.13)$$

The importance of re-circulation is generally limited to laboratory or pilot-plant scale work since power input per unit volume is often held constant for scale up. From equation (2.13):

$$\frac{N_R^2 D}{g} \left(\frac{D}{T}\right)^{2.5} = 0.28 Fl_R^{0.5} \quad (2.14)$$

But,

$$N_R^3 D^2 = \text{constant (constant power-unit volume)}$$

$$\therefore Fl_R^{0.5} \propto \frac{1}{D^{1/3}} \quad (2.15)$$

Re-circulation occurs at $Fl < Fl_R$, so the maximum flow number at which re-circulation can take place decreases with increasing scale, whilst the flow number of the agitator increases with increasing scale, if the traditional scale-up method of keeping the same gas residence time (expressed as vessel volumes per minute) is applied. The result is that re-circulation becomes less likely to occur as scale is increased.

(v) Flow Maps

Since there are so many flow regimes in which an impeller may be operating, the concept of a two-phase flow map is very useful to illustrate the overall picture. Fig 2.3, from Warmoeskerken (1986), illustrates the flow map for a disc turbine. The flow map may be also represented in dimensionless form (Fig 2.4), relating Fr to Fl .

The key to the figures is as follows, again taken from Warmoeskerken (1986) :-

<u>Area</u>	<u>Cavity type</u>	<u>Flooding/Loading</u>
I	Vortex/clinging	Flooding
II	Ragged	Flooding
III	Vortex/clinging	Loading
IV	"3-3"	Loading
V	Vortex/clinging	Loading/re-circulating
VI	"3-3"	Loading/re-circulating

(vi) Other Definitions of Flow Regime Transitions

Several workers have made inferences about the flow regime in the vessel from mass transfer or hold-up measurements, rather than from observation of cavity transitions.

Westerterp, van Dierendonck and de Kraa (1963) measured interfacial area using chemical methods and obtained results such as those illustrated in Fig 2.5. There appeared to be a minimum impeller speed for complete dispersion, above which interfacial area was a function of impeller speed but not of gassing rate. The correlation for minimum speed was :

$$\frac{N_o D}{(\sigma g / \rho)^{0.25}} = A + B \left(\frac{D}{T} \right) \quad (2.16)$$

A and B are constants, depending on impeller type. Gas flow rate does not influence N_o , which does not agree with the predictions of the flooding correlation of Nienow et al (1985). The same trends were shown by the data of van Dierendonck, Fortuin and Van der Bos (1968) and Mehta and Sharma (1971), who also used chemical methods to measure interfacial area. Nishikawa, Nakamura, Yagi and Hashimoto (1981), using a dynamic response technique, produced a series of plots of $k_L a$ against impeller speed at constant gassing rate which showed that there was a minimum impeller speed below which the level of agitation did not affect $k_L a$ (Fig 2.6). The minimum agitator speed was not a function of gassing rate. This result was supported by the work of Wong and Shiuan (1986), who also used a

dynamic response technique to measure $k_L a$.

Closer scrutiny of the data of Nishikawa et al (1981) reveals that the agitator was not flooded at any time according to equation (2.9), mainly because the gassing rates were extremely low. The experiments of Wong and Shiuan (1986), however, were partly flooded and partly loaded. This makes the agreement between the two sets of data difficult to reconcile. To complicate the picture further, there are the findings of Pollard (1978), who found peaks in "aeration efficiency" obtained from $k_L a$ values which appeared to coincide with the flooding definition of Rushton and Bimbinet (1968). Pollard's data showed the inferred flooding point to be a function of gassing rate. Application of equation (2.9) to Pollard's peaks in efficiency showed that the peaks were very close to the flooding point predicted by Nienow et al (1985).

The use of mass transfer data to define flow regime transitions appears to be very difficult to link to changes in cavity structure. Criticisms might be made of the experimental techniques used in the papers discussed in this section which cast doubt on their conclusions in any case. The topic is rather open-ended, but there is little doubt that the flow regime transitions based on physical data and visual observation are very much better-established than inferences made from mass transfer experiments.

2.4 POWER CONSUMPTION : DISC TURBINES

2.4.1 Ungassed Power

There have been many measurements of the power number of the Rushton turbine in the literature (Rushton et al (1950) and Bates et al (1963), for example. The most definitive and wide-ranging data is that of Bujalski, Nienow, Chatwin and Cooke (1987), who gave the following correlation for ungassed power number:

$$Po = 2.5 \left(\frac{x_1}{D}\right)^{-0.2} T^{0.065} \quad (2.17)$$

The striking feature of the relationship is the scale effect, with power number increasing slightly as scale increases.

2.4.2 Gassed Power

For a given agitator, the impeller power depends on the two independent variables, speed and gassing rate. Since the work of Oyama and Endoh (1955), many workers have presented power data as the ratio of gassed to ungassed power demand against the flow number (see Section 2.1). Authors such as van 't Riet et al (1976), held speed constant and varied gassing rate (Fig 2.7), whilst workers such as Nienow et al (1977) held gassing rate constant and varied impeller speed (Fig 2.8).

Plots such as Fig 2.7 may be explained in terms of cavity structure and re-circulation. The fact that the gassed power curves are different for different impeller speeds shows that flow number alone is an inadequate correlating parameter for gassed power demand. According to Warmoeskerken (1986), the curves differ at low flow number because of re-circulation, and at high flow number because of maximum cavity size, which is a function of impeller speed.

The constant flow rate plots (such as Fig 2.8) are more difficult to interpret, but the sharp maximum corresponds to the onset of re-circulation, whilst the shallow minimum corresponds to the complete dispersion condition (Nienow et al, 1977).

There have been several attempts in the literature to produce correlations for the gassed power demand for design purposes. One of the earliest correlations, which is still widely used, was that of Michel and Miller (1962) :-

$$P_g = A \left(\frac{P_u^{2ND^3}}{Q^{0.56}} \right)^{0.45} \quad (2.18)$$

$$A = 0.72$$

The data of Oyama and Endoh (1955) were also correlated by equation (2.15), but Michel and Miller themselves pointed out the major problem with the correlation : it fails at extreme gassing rates. The studies of Yung, Wong and Chang (1979), and Loiseau, Midoux and Charpentier (1977) verified the form of the correlation, but suggested a value of the constant of approximately 0.83 rather than 0.72. The correlation is versatile, accounting for a variety of non-foaming liquids of viscosities up to 0.1 Pas, and for liquids containing up to 5 wt % of solids.

Nienow, Wisdom and Middleton (1977) pointed out another fundamental weakness of the Michel-Miller type correlation. If the gassed power number is approximately constant, then:

$$P_g \propto N^3 D^5 \quad (2.19)$$

Also,

$$P_u^2 \propto N^6 D^{10} \quad (2.20)$$

At constant gassing rate, the plot reduces, effectively, to a plot of $N^3 D^5$ against $N^7 D^{11}$, and a slope of 0.45 will always result. Hence, the consistency of the correlations produced by various workers may be regarded as inevitable. This still does not explain the exponent on gas flow rate being constant for the results of several different researchers.

The major criticism of the Michel-Miller type correlation is the scale at which the experimental data has been obtained. Michel and Miller used two vessels, the larger of which was 0.29 m in diameter, whilst Yung et al and Loiseau et al used vessels of 0.22 m in diameter. Pharamond, Roustan and Roques (1975) found that the constant in equation (2.18) increased with scale (in tanks up to 1.0 m diameter) and instead produced a correlation

of their own :-

$$1 - \frac{P_g}{P_u} = 1.6 Q_{vvm} D^{0.63} \quad (2.21)$$

$$Q_{vvm} D^{0.63} > 0.3, \quad \frac{P_g}{P_u} = 0.5$$

Equation (2.21) does not predict any dependency of gassed power ratio on impeller speed, and must therefore be regarded with caution. Hassan and Robinson (1977) and Luong and Volesky (1979) used the following correlating equation :

$$\frac{P_g}{P_u} = A We^b Fl^c \quad (2.22)$$

Both studies found $c = -0.38$, but differed in the value of the exponent on Weber number. Hughmark (1980), in attempting to correlate the work of several authors, found that the exponent on Weber number was not statistically significant, which agrees with the findings of Michel and Miller, who used liquids with a wide range of surface tensions to obtain their correlation.

The problem with the correlations discussed so far is that no account of changes in flow regime has been made. To obtain a good fit to the gassed power curve a rather more sophisticated approach than a simple power law relationship would appear to be needed.

Brown (1981), using measurements of the power drawn by the motor of a production scale fermenter (a technique which is notoriously prone to error), produced the following correlation:

$$Po_g = \alpha Poe^{-\beta Q} \quad (2.23)$$

α and β were assigned distinct values at high and low flow rates. Perhaps this reflected the difference between clinging cavities and the "3 - 3" structure. Again, the correlation predicts no dependence of P_g/P_u on impeller speed.

The approach of Greaves and Kobbacy (1981) was to produce a separate correlation for each flow regime. The changes in flow regime, however, were inferred from the power measurements and not corroborated by visual observation.

Bader (1986) made power measurements in a pilot scale fermenter, using the power drawn by the motor and attempted to correlate the data using an exponential decay equation which did not allow for the effects of impeller speed. The fit to the data was rather poor, which appears to have been due to experimental problems as much as the philosophy behind the correlating equation. Experiments were carried out at different agitator clearances from the vessel bottom which showed that there was little influence of clearance on P_g/P_u between one and two impeller diameters from the vessel bottom.

It is clear that an accurate predictive equation for gassed power consumption would be valuable to the industrial designer. All the correlations discussed so far appear to suffer from some fundamental flaw. There are other correlations, such as those of Nagata (1975), Gray, Treybal and Barnett (1982) and Judat (1977), all of which are highly complex. The best approach available appears to be that of Warmoeskerken (1986), who linearised the gassed power curves in different flow regimes to produce a set of simple correlating equations. The experimental data fitted the correlation to $\pm 20\%$, which is not better than the fit obtained by Michel and Miller (1962).

Correlations for gassed power are discussed further in a later section.

Ismail, Nagase and Imon (1984) attempted to use gassed power measurements for Rushton turbines with 6 or 8 blades in conjunction with visual observation to produce correlations for cavity transitions. This involved plotting gassed power ratio against impeller speed at constant gas flow rate. Their

definitions of cavity type were rather obscure, but a flooding point was defined, and the data agree approximately with the form of equation (2.9). The other cavity transitions did not show the relationship between flow number and Froude number predicted by other authors. The interpretation was based on the cavity structures described by van't Riet (1975), which Warmoeskerken (1986) considered to be erroneous as a result of the observation technique.

It is appropriate to include at this stage the power characteristics of the concave turbine (Fig 2.9), first published by van't Riet, Boom and Smith (1976), whose results are shown in Fig 2.10. The curve is much flatter than the corresponding Rushton turbine curve, for reasons which are still poorly understood. Presumably the concave turbine, which gives rise to streamlined flow in single phase operation, forms smaller cavities than the flat bladed Rushton turbine at a given gassing rate, producing a flatter power curve. This is a desirable feature for designs which require the agitator to operate in liquid only at some stage in the process. An example of this was given by Gbenwoyo, DiMasi and Buckland (1986), who pointed out that in certain aerobic fermentations, the nutrient medium is sterilised in the fermenter, requiring agitation. An impeller with a flat power curve will require a smaller motor than a Rushton turbine designed to deliver the same gassed power input. There may also be advantages from reduced surface aeration from the concave turbine in the single phase operation, but information on surface aeration for this impeller is lacking.

2.5 HYDRODYNAMIC ASPECTS : PITCHED BLADE TURBINE

A description of the sequence of cavity formation has been given by Warmoeskerken, Speur and Smith (1984) and information was also presented by Nienow, Konno and Bujalski (1986), Chapman et al (1983) and Bujalski (1987). If one considers an agitator rotating at constant speed, the following events occur as gas

flow rate is increased (Fig 2.11):

- a) At zero gas flow rate, there are strong vortices trailing from the tip of the impeller blade, as described by Tatterson, Yuan and Brodkey (1980).
- b) At low gas flow rates, vortex cavities form behind each impeller blade.
- c) Clinging cavities then appear which gradually grow with increasing gassing rate until large cavities with a characteristic flat rear face are formed. The large cavities form simultaneously and over a very narrow range of gassing rates.

This sequence is valid for downward pumping agitators.

As with the Rushton turbine, there are also distinct two-phase flow regimes within the vessel. This was shown by the work of Chapman et al, as well as Warmoeskerken et al (1984), for downwards-pumping agitators.

At low gassing rates (Fig 2.12) gas bubbles are swept away from the sparger to be partly re-circulated into the impeller by strong liquid flows (indirect loading). As gassing rate is increased the cavities grow and the pumping capacity of the impeller falls until the downwards liquid velocity is smaller than the bubble rise velocity. Beyond this critical gassing rate, bubbles rise directly into the impeller (direct loading). The transition point coincides with the formation of large cavities.

The transition from indirect to direct loading is very sensitive to sparger type and clearance, as shown by Frijlink, Kolijn and Smith (1984) and Warmoeskerken et al (1984). Also, Chapman et al (1983) reported gross instabilities between the two regimes, with accompanying fluctuations in impeller torque. However, these effects were not noted by Warmoeskerken et al (1984). It

seems likely that the ratio D/T is the controlling parameter, since Chapman et al (1983) found gross instabilities at $D = T/4$ but not at $D = T/2$. Warmoeskerken et al (1984) used $D = 0.4T$.

In fact, the two sets of workers produced very similar functional relationships between the impeller speed and the critical flow rate at the flow regime transition :

$$\text{Chapman et al :} \quad N_L = \frac{11.7 Q_L^{0.5}}{D^{1.65}} \quad (2.24)$$

$$\text{Warmoeskerken et al} \quad N_L = A Q_L^{0.5} \quad (2.25)$$

For upwards-pumping agitators, the sequence of cavity formation has been described by Bujalski (1987). At constant gas flow rate, for an impeller with $D/T = 1/2$ and a large ring sparger, the following sequence occurs:

- a) At low agitator speed, no cavities are formed. The agitator is flooded and the gas is poorly dispersed.
- b) With increasing speed, there is a fairly sharp transition to vortex cavities. Liquid circulation loops are formed into which some of the gas is entrained.
- c) With further increases in impeller speed, the vortex cavities gradually form clinging cavities which cover an appreciable area of the trailing surface of the blade. The clinging cavities grow with increasing agitator speed. When clinging cavities are present, the gas is fairly uniformly dispersed in the vessel.

The sequence for agitators with $D = T/3$ is similar except that vortex cavities or large cavities are formed in the flooded condition. These become clinging cavities when the impeller speed is above that required to prevent flooding.

Upwards-pumping agitators do not undergo such dramatic changes

in flow regime as downwards-pumping agitators. The impeller is directly loaded at all times, but this does not lead to poor gas distribution as is the case for a downwards-pumping impeller. Also, instabilities of flow between the impeller inlet stream and liquid entrained into the bubble swarm cannot occur, as the two flows are in the same direction.

2.6 GASED POWER : PITCHED BLADE TURBINE

Downwards pumping agitators produce steep falls in P_g/P_u when plotted against flow number at constant impeller speed until the transition to direct loading takes place. The curves then become flat, with slight increases in P_g/P_u as gassing rate increases (Fig 2.13).

Upwards-pumping agitators (Chapman et al, 1983) give much flatter power curves, presumably because the cavity transitions are gradual. This offers potential practical advantages over downwards-pumping agitators, as discussed earlier for the concave turbine.

2.7 GAS HOLD-UP

The fraction of gas contained within the dispersion is of interest for the following reasons :-

- 1) It is a measure of the efficiency of an agitator in dispersing the gas. Overall hold-up is more straightforward to measure and interpret than mass transfer coefficient. Andrew (1982) showed that hold-up is proportional to mass transfer coefficient over the range of power inputs encountered in industrial stirred tank fermenters. Mass transfer is the desired process result in a gas liquid stirred tank, but for comparative purposes, hold-up is an adequate parameter, at least in vessels of relatively low aspect ratio.

- 2) It is of industrial significance since, for a fixed volume the gas hold-up indirectly determines the liquid volume, in which the productive reactions take place. This is important for high added value process steps such as fermentations.

2.7.1 Measuring Methods

Overall gas hold-up may be measured in the following ways :-

- (i) The level change on gassing is recorded and the gas volume in the dispersion calculated. The method of measuring the dispersion height is critical to the accuracy of the hold-up measurement. The following techniques are available :
- Visual observation. This has been employed by several authors including Rushton and Bimbinet (1968) and, for foaming systems, Davies (1986).
 - Manometric methods. The axial pressure profile in the tank is measured and the position of the dispersion surface inferred by extrapolation to atmospheric pressure. The technique has been extensively used, for example by Calderbank (1958), Robinson and Wilke (1974), and Yung, Wong and Chang (1979).
 - Ultra-sound. The use of commercial ultra-sonic level detectors has been described by Davies (1986) and Bujalski (1987).
 - Conductivity. Vlcek, Steidl and Kudrna (1969) used electrodes to detect the dispersion surface and took a spatial average of several measurements to obtain the dispersion height.
 - Level control. Warmoeskerken (1986) described a

technique using an adjustable funnel and a circulating system with a reservoir. The circulation was set up in the ungassed condition and the liquid level in the reservoir was noted. On gassing the funnel was moved upwards until the liquid in the reservoir returned to its ungassed level. The difference in funnel heights was the change in level due to gassing.

All the techniques listed above have drawbacks, especially when very disturbed surfaces are encountered. The method chosen for the level change measurements in this thesis was the ultra-sonic technique because it is straightforward to use and provides an average value of the dispersion level across a representative proportion of the surface. It is also possible to automate the measurements to a large degree, which speeds up the process of data acquisition. The use of the ultra-sonic technique is described in detail in the section describing experimental methods.

- (ii) The dispersion level is held constant and liquid is displaced from the dispersion by the gas into a collecting tank. The liquid volume is measured, giving the gas volume and hence the hold-up. This method was adopted for a substantial proportion of the hold-up measurements described in this thesis.

2.7.2 Hold-up Correlations

Hold-up correlations fall into two categories :

1. Power per unit volume and superficial gas velocity (Table 2.1):

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (\text{Type A}) \quad (2.26)$$

2. Impeller speed and gas flowrate (Table 2.2) :

$$\epsilon_G = aN^b Q^c \quad (\text{Type B}) \quad (2.27)$$

This form of correlation may be expressed in various ways, as inspection of Table 2.2 will show.

The regression coefficients α , β and γ in the type A correlation are subject to considerable variation. Rushton and Bimbinet (1968) found γ to be reasonably constant, but found α and β vary with scale and D/T. One cause of this may be that different researchers have been operating in different flow regimes. The data of Davies (1986) show big differences in behaviour between the clinging cavity and "3-3" regimes. Another reason for the variations may be the different measuring methods used, or the statistical analysis technique used to obtain the regression coefficients. What is certain is that there is a large variation in the prediction of the correlations in Table 1. For example, at typical operating conditions of 1 W/kg power input and gas superficial velocity of 0.01 m/s, the predictions for hold-up range from 4% to 7%.

Mann (1983) made the comment that gas hold-up is very sensitive to traces of contaminants, citing the example of Nienow and Machon (1979), who found 30% increases in hold-up for the addition of 0.1 wt % of formaldehyde to a CMC solution. Unperceived contamination of test liquids may well be a problem for hold-up measurements, but the effect is probably not dramatic. Certainly, very powerful surfactants exist which affect hold-up markedly in very small amounts but there are visual signs of contamination, such as foaming and the formation of very small bubbles which render the liquid opaque on gassing. The most likely contaminants in a laboratory environment are hydrocarbons from gear-box oil and air-borne dust. Neither of these substances are surface active. It seems unlikely that the differences in hold-up results between different sets of researchers are due to contamination,

especially on inspection of the data of Davies (1986), who observed increases in hold-up of approximately 30% comparing 0.11M potassium sulphate solution with water. The data of Das, Bandyopadhyay, Parthasarathy and Kumar (1985) support the statement that the degree of sensitivity of gas hold-up to contamination is relatively small. They studied the effect of iso-hexanol and a ketone on interfacial area and hold-up and the data imply that the expected change in hold-up for water when saturated with either of the substances studied is insignificant.

The form of the type A correlation is of some interest. The equations marked '*' in Table 2.1 include the power due to the isothermal expansion of the gas :-

$$\epsilon_G = \alpha (\bar{\epsilon}_M + \bar{\epsilon}_g)^B v_{SG}^Y \quad (2.28)$$

The expansion power of the gas is given by :

$$\bar{\epsilon}_g = \frac{\rho_g}{\rho_L} \frac{Q}{V} \frac{RT}{M} a \ln \left(\frac{p_i}{p_o} \right) \quad (2.29)$$

Davies (1986) did not use the isothermal expansion power of the gas, but used the power required to compress the gas to the sparger pressure :

$$\bar{\epsilon}_g = \frac{\rho_L g Q H}{\rho_L V_L} = g v_{SG} \quad (2.30)$$

The two expressions are not equivalent, but give approximately the same values for the gas expansion power provided that the gas pressure at the sparger is relatively small.

The advantage of using the gas isothermal expansion power is that, as the impeller power input tends to zero, there is still a finite hold-up due to the vessel acting as a bubble column. This makes the correlation more conceptually acceptable than the original form.

Miller (1974) pointed out that the two forms of power input may not be simply additive, but that some fraction of the gas isothermal expansion power should be used. A complex method of devising the correct fraction to use was devised based on bubble size measurements in an unagitated vessel. This point is discussed in a later section.

Sridhar and Potter (1980) investigated the predictions of Miller's modifications to Calderbank's hold-up and interfacial area correlations for a set of experiments in a vessel pressurised up to 10 bar. The kinetic energy of the gas increases with increasing pressure at constant volumetric gas flow rate due to increasing gas density, and can become significant. Simply adding the gas isothermal expansion power to the mechanical power input by the stirrer was not sufficient to account for the changes in hold-up and interfacial area caused by increasing the vessel pressure. The discrepancies were attributed to kinetic energy effects. The proposed modified form of Calderbank's correlation was :

$$\epsilon_G = \left(\frac{\epsilon_{GSG}}{U_t} \right)^{0.5} + 2.16 \times 10^{-4} \left[\frac{(P_g/V)^{0.4} \rho^{0.2}}{\sigma^{0.6}} \right] \left(\frac{U_t}{v_{SG}} \right)^{0.5} \frac{E_t}{P_g} \left(\frac{\rho_g}{\rho_a} \right)^{0.16} \quad (2.31)$$

$$E_T = \text{total power input} = P_g + \rho_L g H_L Q$$

The term ρ_g/ρ_a is, for air sparging, equivalent to the operating pressure divided by atmospheric pressure. It accounts explicitly for the increase in gas kinetic energy due to increasing pressure.

One of the major drawbacks of the Type A correlation is that the designer must obtain a reasonably accurate estimate of the gassed power demand before estimating hold-up from the correlation. The accuracy of existing power correlations is not good (about $\pm 20\%$), and the problem is compounded by the fact

that the fit of the hold-up correlation to the data is at best $\pm 20\%$.

The type B correlations (Table 2.2) at least relate hold-up to the independent variables of impeller speed and gas flow rate. There are inherent problems involved with using dimensionless groups in the way proposed by Yung et al (1979) and Warmoeskerken (1986). All the parameters within each dimensionless group should be varied, and there must be a variation of both vessel size and D/T if useful correlations are to be obtained.

The correlation of Hassan and Robinson is very restrictive since the exponent on speed must always be twice the exponent on flow rate. Only two regression coefficients are obtained, which probably explains the relatively poor fit to the authors' own data. The correlation of Yung et al (1979) was only based on data from one scale, but covered a range of D/T. There is a very large difference in dependency of hold-up on impeller diameter between the correlations of Yung et al and Warmoeskerken (1986). Warmoeskerken used data from two vessels and one D/T and found the exponent to be 0.16 (at most). Yung et al found the exponent to be 1.85. This suggests some fundamental flaw in the correlation method.

2.7.3 Effect of coalescence inhibitors

Several authors, such as Hassan and Robinson (1979) and Davies (1986), have produced separate correlations for electrolyte solutions and pure liquids or aqueous solutions of organics. Several papers have been written on the influence of electrolytes and surface active agents on the coalescence behaviour of gas bubbles, such as the work of Keitel and Onken (1982), Lessard and Zieminski (1971), and Marrucci (1967 and 1969). The presence of metal ions in the thin liquid film between two bubbles about to coalesce increases the short range order in the liquid and hinders drainage of the film, thus

repressing coalescence. Surfactants are thought to act in a similar manner.

The significance of these phenomena is that the mean bubble size in the dispersion is reduced. The stable population of bubbles reflects a balance between coalescence and break-up, as shown in the work of Otake, Tone, Nakao and Mitsuhashi (1976). If coalescence is inhibited, but not the processes leading to break-up, the mean bubble size reduces. The rise velocity of the bubble diminishes and the gas hold-up in the vessel increases. The effect of electrolyte on bubble sizes in stirred tanks was shown by Koetsier and Thoenes (1972) (Fig 2.14). Clearly, the interfacial area is also sensitive to coalescence inhibition since :-

$$a = \frac{6\epsilon_G}{d_{SV}} \quad (2.32)$$

The presence of coalescence inhibitors is also critical for mass transfer, as discussed in a later section.

Lee and Meyrick (1970), argued on thermodynamic grounds that a parameter may be defined to characterise the coalescence behaviour of bubbles in electrolytes :

$$\psi = C \left(\frac{d\sigma}{dC} \right)^2 \left(1 + \frac{d \ln f}{d \ln C} \right)^{-1} \quad (2.33)$$

For two solutions with identical values of ψ the changes in surface tension caused by the stretching of the dividing film characteristic of coalescence will be the same.

Hence ψ characterises coalescence behaviour. The hold-up data from two electrolytes were successfully explained using this approach.

Machon, Vlcek and Kudrna (1977), attempted to incorporate this idea into a quantitative correlation :-

$$\epsilon_G = A We^b Fl^c (2 - e^{-\psi+})^d \quad (2.34)$$

ψ^+ was a critical value of ψ , beyond which ψ had no further influence on hold-up. The correlation was designed to give the correct behaviour as $\psi \rightarrow 0$, as is the case for water. Behind this approach is the implicit assumption that the coefficients a , b , c do not depend on the coalescing nature of the liquid. There were some problems in applying the correlation to the data of Lee and Meyrick which were ascribed to differences in geometry between the two sets of experiments and to differences in measuring technique. In view of the problems of compatability of data in water between different researchers, it is not surprising that the data of Lee and Meyrick and Machon et al are difficult to reconcile.

The related problem of foaming liquids was addressed by Loiseau, Midoux and Charpentier (1977). They realised that surface tension alone is not sufficient to characterise foaming, nor is ionic strength. The following correlation was proposed for foaming liquids :-

$$\epsilon_G = 0.26 v_{SG}^{0.24} (\bar{\epsilon}_m + \bar{\epsilon}_g)^{0.57} \quad (2.35)$$

The coefficients on v_{SG} and power input differ somewhat from the exponents recommended by the same workers for non-foaming liquids. It should be noted that the non-foaming exponent on v_{SG} was markedly lower than the results of other workers (Table 2.1).

2.7.4 General Comments

Overall, the picture is rather confused. Although the measurement of gas hold-up is in principle fairly simple, there are large differences between correlations produced by different authors using similar geometries. The reasons for this are not clear. The problem of establishing general trends is made worse by the presence of coalescence inhibitors such as electrolytes.

There are still questions about the correlating techniques. The power input/superficial gas velocity correlations are dimensional and poorly grounded in theory. The dimensionless correlations based on impeller speed and gas flow rate appear to have many underlying problems which have not been addressed.

2.8 MASS TRANSFER

The prime objective of gas-liquid contacting in stirred tanks is to achieve transfer of a component from the gas phase into the liquid. The transfer process is described as follows :-

$$R_V = k_L a \Delta C \quad (2.36)$$

ΔC = appropriate concentration driving force .

The resistance to mass transfer in the gas phase is regarded as negligible for sparingly soluble or pure gases, so the overall mass transfer coefficient is equal to the liquid side mass transfer coefficient.

Work on mass transfer has focussed on the measurement of $k_L a$, or in some early work, simply of the specific interfacial area, a . Much controversy has been aroused by mass transfer measurements because of the assumptions made in analysing experimental data, and as a result of the measuring methods employed. It is appropriate, therefore, to make a brief survey of the measuring methods available before considering the published correlations.

2.8.1 Chemical Methods

Much of the early work on mass transfer involved the absorption of a gaseous reactant into a solution. An example is the conversion of sulphite ions to sulphate by oxygen in the presence of Co^{2+} or Cu^{2+} catalyst (Westerterp et al, 1963, Greenhalgh, McManamey and Porter, 1975). By choosing a reaction

regime in which rate is independent of k_L , and by ensuring that no reaction takes place in the liquid film, interfacial area can be measured by monitoring the rate of reaction by chemical analysis.

A more subtle approach using a technique suggested by Danckwerts and Sharma (1970) has been used by Mehta and Sharma (1971), Robinson and Wilke (1974) and Hassan and Robinson (1980). These authors used the absorption of CO_2 by various bases. The reaction regime must be such that some reaction takes place in the bulk liquid and some in the liquid film. The reaction rate is then described by :

$$R_V = C^* (D_A k C_B + k_L^2)^{\frac{1}{2}} \quad (2.37)$$

By varying the bulk concentration of the liquid phase reactant, C_B , both k_L and a may be obtained simultaneously.

There has been much criticism of the chemical method in the literature. The main objections are :

- (i) It is largely restricted to electrolyte (non-coalescing) systems. The physical properties of the liquid may change with time as the reaction proceeds. A few chemical method studies in coalescing systems have been carried out : Sridharan and Sharma (1976) used organic solutions of amines and Neelakantan and Gehlawat (1982) used butadiene absorption into molten maleic anhydride. These systems are rather exotic, and begin to present experimental problems of their own.
- (ii) The chemical kinetics of the system must be precisely quantified. This may be a problem if the reaction is catalysed by trace contaminants. A classical example of this difficulty was the work of Westerterp et al (1963). The oxidation of sulphite by oxygen was assumed to be first order in oxygen under the experimental conditions used.

Mehta and Sharma (1971) stated that the reaction would actually have been second order in oxygen.

- (iii) The reaction regime may be dependent on position within the tank. The regime is principally determined by the value of the Hatta number:

$$Ha = \frac{D_A k C_B}{k_L^2} \quad (\text{first order reaction}) \quad (2.38)$$

Sridhar and Potter (1978) used both the chemical method and a light transmission technique to measure interfacial area. A discrepancy was found between the two techniques which was attributed to the problem of spatial variation of reaction regime. The local rate of energy dissipation varies over the vessel volume (Okamoto, Nishikawa and Hashimoto, 1981), and k_L is apparently a function of energy dissipation per unit volume (Figueiredo and Calderbank, 1979) although the exact nature of the relationship is unclear. Hence, k_L probably varies with position in the vessel. Sridhar and Potter actually used a weaker version of this argument, but there was a big discrepancy between the interfacial area measured by light transmission and the area measured by sulphite oxidation. Reith (1968) found that the chemical method gave interfacial areas 1.5-3 times larger than those obtained from photographs.

- (iv) Midoux and Charpentier (1979) and Midoux, Laurent and Charpentier (1980) considered the limits on the use of the chemical method imposed by the assumed gas residence time distribution. The assumed degree of back-mixing in the gas phase can significantly affect the values of interfacial area calculated from experimental data when depletion of the gas phase becomes significant.

2.8.2 Steady State Methods

These techniques use a method of continuously removing key component from the liquid phase at the same rate as material is being transferred from the gas phase (Fig 2.15). Assuming perfect mixing in both phases, a simple mass balance yields :

$$k_L a = \frac{L_m (C_{Lo} - C_{Li})}{V \left(KC_{Gi} + \frac{KL}{G_m} (C_{Li} - C_{Lo}) - C_{Lo} \right)} \quad (2.39)$$

If the re-circulating liquid is completely stripped of key component ($C_{Li} = 0$), it may be shown that :

$$\frac{k_L a \tau_R}{\rho_{LM}} = \frac{x}{1 - x \left(1 + \frac{KL}{G_m} \right)} \quad (2.40)$$

$$x = \frac{C_{Lo}}{KC_{Gi}}, \text{ the fractional degree of saturation of liquid.}$$

Since the concentration in the liquid outlet stream must be measured, there are experimental limits on the permissible degree of saturation of the outlet stream. This leads to limits on the liquid residence time which require unacceptably high liquid throughputs at large scale. Also the means of removing the key component must be carefully considered, especially for large scale physical desorption.

Zlokarnik (1978) introduced hydrazine continuously into the test vessel in the presence of a catalyst to remove oxygen by chemical reaction in situ. The technique has been improved by Onken, Sick and Weiland (1985) who used more reproducible catalysts than Zlokarnik. Hydrazine reacts with oxygen to give water and nitrogen, so the physico-chemical properties of the liquid are not affected by the reaction. The experimental conditions, however, require rigorous control to achieve reproducible results.

Another variation on the steady state technique is the use of a micro-organism, such as yeast, to provide an oxygen sink within

the vessel (Hickman and Nienow, 1986). The use of such media tends to produce characteristic liquid properties, but offers great promise for $k_L a$ measurement in viscous liquids in particular.

The method described by Kreysa and Wobeken (1985) utilises an electrochemical reaction to remove oxygen, producing a steady state. This technique appears to be difficult to apply to large vessels, and would appear to be too complex to justify its advantages over dynamic methods. It has the benefit of not requiring a liquid throughput and can also be used for dynamic response experiments. The whole vessel is effectively used as an oxygen electrode.

2.8.3 Dynamic Response Methods

The dynamic response technique uses step changes in the inlet gas composition to produce transient dissolved gas concentrations in the liquid phase. Oxygen is generally used as the key component.

Amongst the first workers to use the method were Bandyopadhyay, Humphrey and Taguchi (1967), and it has been used by many workers since. Nitrogen is used to reduce the oxygen concentration in the vessel to approximately zero, when the feed is switched to air at the same flow rate as the nitrogen. The oxygen concentration time history is recorded for analysis.

The basic equation describing the time history is as follows :-

$$\frac{dC_L}{dt} = k_L a (C_G^+ - C_L) \quad (2.41)$$

C_G^+ = representative gas phase concentration.

The liquid phase is assumed to be well-mixed, an assumption which is discussed in detail in a later section, but which is of less significance than the assumed value of C_G^+ .

Four models are available :-

(i) No depletion

The gas phase concentration is assumed to be constant throughout the tank, and equal to the inlet gas concentration (Bandyopadhyay et al, 1967 and van't Riet, 1975, are examples of workers using this assumption). The gas phase mass balance is violated and the gas phase oxygen concentration is over-estimated.

Consequently the value of $k_L a$ is under-estimated. The error becomes quite serious at high levels of depletion.

(ii) Well-mixed gas phase

The gas phase concentration is assumed to be uniform throughout the vessel and equal to the outlet gas concentration. The full gas phase mass balance is thus written as :-

$$V_G \frac{dC_{Go}}{dt} = QC_{Gi} - QC_{Go} - k_L a (KC_{Go} - C_L) \quad (2.42)$$

The various simplifications of the mass balance were discussed in full by Dunn and Einsele (1975), and Figueiredo and Calderbank (1979) compared the calculated values of $k_L a$ from each well-mixed gas phase model. The variations were considerable with the full mass balance giving the highest $k_L a$ value and the no depletion model (ignoring the accumulation and transfer terms) giving the lowest $k_L a$ value.

In fact, the well-mixed gas phase model gives the lowest possible estimate for C_G^+ , and hence the highest possible value for $k_L a$. The well-mixed model tends to be closest to the true state of gas phase back-mixing when there is a significant amount of gross gas-phase re-circulation, which tends to be the case for small scale apparatus rather than production scale tanks.

(iii) Plug Flow

The gas phase is assumed to be in plug flow, which represents the opposite extreme of assumed residence time distribution to the well-mixed model. Davies (1986) used a full, unsteady state plug flow model described by the following equations :-

$$\text{Liquid Phase : } \frac{dC_L}{dt} = \frac{k_L a}{H} \int_0^H KC_G - C_L dx \quad (2.43)$$

$$\text{Gas Phase : } \frac{dC_G}{dt} = v \frac{dC_G}{dx} - \frac{Kk_L a V_L}{V_G} (KC_G - C_L) \quad (2.44)$$

$$\frac{dv}{dx} + \frac{1 - \epsilon_G}{\epsilon_G} \frac{k_L a}{\rho_{LM}} (KC_G - C_L) = 0 \quad (2.45)$$

$$v = \frac{QH}{V}$$

Equation (2.45) is included to account for the reduction in gas volume flow as a result of mass transfer. By assuming that

$\frac{dC_G}{dt} = 0$ and $\frac{dv}{dx} = 0$, the equations can be reduced to the familiar log-mean concentration driving force expression. This method was used by Nishikawa et al (1981).

(iv) Double response

By the use of overall mass balances on the key component, Chapman, Gibilaro and Nienow (1982), produced the following equation to describe the system response :

$$\frac{dC_L}{dt} = k_L a \left[\frac{Q}{V_G} \int_0^t C_{Gi} - C_{Go} dt - C_L \left(\frac{KV_L}{V_G} + 1 \right) \right] \quad (2.46)$$

No assumption is necessary about the gas phase RTD. In deriving equation (2.46), Chapman et al assumed a uniform bubble size.

Davies (1986) has shown this assumption to be unnecessary.

As well as the liquid phase transient, the gas phase key component concentration must be monitored, and the overall gas hold-up must be measured. This introduces further experimental complexity to the measurement of $k_L a$, but does produce an unbiased estimate of $k_L a$. This is clearly a more satisfactory situation than the assumption of a gas phase RTD. As has already been discussed, the simple gas phase models are limited in their application, and it is not possible, a priori, to determine whether or not a particular model is applicable under a given set of circumstances. Information on gas phase RTD's is very scarce. Gal-Or and Resnick (1966) and Popovic, Papalexou and Reuss (1983) presented experimental data which showed the RTD to be intermediate between plug flow and perfect mixing. No information is available on scale or other effects, and no useful predictions about the gas RTD at given operating conditions can be made.

The error incurred by assuming a simple gas phase RTD is dependent on the amount of depletion in the gas phase. Fig 2.16 is reproduced from Davies (1986), and shows $k_L a$ values calculated using each of the four models described in this section from Davies' experimental data. The double response and plug flow data are in reasonable agreement, especially for $k_L a$ values below 0.1 s^{-1} , and do not show the tendency for $k_L a$ to reach a maximum with respect to power input which was a feature of the data of van't Riet (1975). The graph also illustrates the large differences in $k_L a$ between the various assumed models.

The problem of the influence of assumed gas phase RTD on $k_L a$ complicates the comparison of data from different sources. Another difficulty with early data was the response time of the oxygen electrode. Dunn and Einsele (1975) pointed out the importance of considering the dynamics of the oxygen probe. Fortunately, the response may be regarded as first order (Davies, 1986) for a fast-response probe. Other authors, such as Chandrasekharan and Calderbank (1981), have used second order models for the probe response. Van 't Riet (1975) showed that

the probe response may be neglected for the no depletion model provided that the characteristic time of the probe is less than 3 seconds and $k_L a$ does not exceed approximately 0.1 s^{-1} .

Andre, Moo-Young and Robinson (1985) have described an ingenious variation of the dynamic response technique using analysis of the off-gas concentration only. This is advantageous for viscous liquids, when probe boundary layers greatly complicate the measurement of dissolved oxygen concentration. The drawback is that the off-gas transient is much faster than the liquid transient which leads to problems of measurement and de-convolution of the probe response. There are also potential problems of analysing such a rapid response curve. The technique is not suitable for use in low viscosity media.

Davies (1986) described a method based on the initial response of the liquid oxygen probe to a change in gas inlet concentration. The results agreed fairly well with the double response technique in tap water but not in electrolyte. The advantage of the initial response method is that no assumption about gas phase mixing need be made, but there would appear to be experimental problems in non-coalescing media. The value of $k_L a$ is given by:

$$k_L a = 6b_3 \frac{\tau_L V_G}{Q} \quad (2.47)$$

b_3 = coefficient in quartic equation fitting liquid oxygen probe transient.

One of the problems with this technique is that sensitivity of $k_L a$ to errors in hold-up, flow rate and liquid probe response time is greater than for other techniques.

2.8.4 Mass transfer correlations : coalescing media

Mass transfer correlations take the same form as those generally used for gas hold-up. The most common correlating equation used

in the literature is :

$$k_L a = \alpha \left(\frac{P_g}{V} \right)^\beta v_{SG}^\gamma$$

Table 2.3 is a summary of some of the literature correlations available. The experimental difficulties already discussed mean that the data of different researchers are often incompatible. The range of values for β is 0.15 - 0.8, and the range of values for γ is 0 - 0.75. Van 't Riet (1979) produced a correlation from a wide range of sources, which predicted the experimental values to $\pm 40\%$. The data of Davies (1986) are very poorly correlated by van't Riet's equation, probably because the data used by van't Riet were all, in some degree, inaccurate as a result of assumptions about gas phase mixing. Not only are the absolute values different, but Davies' correlations show very different trends to the rest of the literature data.

The dependency of $k_L a$ on power input is particularly sensitive to the assumed gas residence time distribution (Fig 2.16). This probably accounts for the higher exponent on (P_g/V) found by Davies in comparison to workers using the no depletion assumption.

A further complication is the sensitivity of $k_L a$ to trace amounts of inert contaminant. Warmoeskerken and Smith (1978) reported falls in $k_L a$ of 50% or more on the addition of 20 ppm silicone oil to water. The sensitivity of $k_L a$ to contamination is, of course, highly dependent on the nature of the contaminant, and is impossible to generalise. There may be effects from the material added to tap water during treatment by water authorities, which make comparison of data from different laboratories a potentially hazardous exercise. Wong and Shiuan (1986) measured $k_L a$ values in water in the presence of various additives. Their data show n-dodecane, which is more representative of potential contaminants than silicone oil, to have a less pronounced effect on $k_L a$ than the phenomena described by Warmoeskerken and Smith.

The ranges of operating conditions used by the various workers were approximately the same, but there may have been variations in flow regime which have not been taken into account.

Warmoeskerken (1986) and Davies (1986) produced different correlations for clinging cavities and "3-3" cavities which may explain the variations in β and γ found by earlier workers.

There is also the effect noted by Nishikawa et al (1981) who found distinct differences in behaviour between flow regimes termed agitator-controlled and gas-controlled, after the definition of Oldshue and Connelly (1977).

As a development of the power input, superficial gas velocity approach, Zlokarnik (1978) proposed the use of the following dimensionless function :

$$k_L a \left(\frac{V}{Q} \right) = f(P^*) \quad (2.48)$$

$$P^* = \left(\frac{P/Q}{\rho g^{2/3} v^{2/3}} \right) \quad (2.49)$$

The empirical equation based on data from 0.4 m and 0.6 m tanks was :

$$k_L a \left(\frac{V}{Q} \right) = 0.015 (P^*)^{0.5} \quad (2.50)$$

This can be re-written as for constant physical properties :-

$$k_L a = C \left(\frac{P}{V} \right)^{0.5} \left(\frac{Q}{V} \right)^{0.5} \quad (2.51)$$

This is equivalent to a correlation in terms of power per unit volume and flow rate in VVM. Henzler (1982) was probably right to question the validity of this equation on scale-up. Although the equation can be written in terms of dimensionless quantities, there are only two regression coefficients, and the fit to Zlokarnik's own data was rather poor.

2.8.5 Mass transfer correlations : non-coalescing media

As with gas hold-up, the presence of a coalescence inhibitor, such as an electrolyte, has a profound influence on $k_L a$. This is a direct result of the decrease in bubble size, which increases the specific interfacial area, a . The liquid side mass transfer coefficient is only slightly reduced by the presence of electrolyte (Zieminski and Whittemore, 1971).

Table 2.4 summarises the results of several workers who used electrolyte solutions. Generally, $k_L a$ values were much higher than the corresponding $k_L a$ value in tap water. Smith et al (1977) and Davies (1986), did not observe any marked changes in β and γ with changes in ionic strength. van't Riet (1979) produced a correlation with β higher and γ lower than for coalescing media. Robinson and Wilke (1973) found no dependence of $k_L a$ on v_{SG} which agreed with the interfacial area measurements of Westerterp et al (1963). Robinson and Wilke found that

α increased with increasing ionic strength, whilst β also increased with increasing ionic strength :

$$\beta = 0.4 + \frac{0.862^{r_T}}{0.274 + r_T} \quad (2.52)$$

$$\text{If } r_T \geq 0.4, \beta = 0.9$$

The effect of non-electrolytic coalescence inhibitors can be very concentration-dependent. Van der Donk, Van der Lans and Smith (1979) showed that for a plunging jet n-octanol could either enhance or repress mass transfer depending on concentration. This effect was also found by Aggarwal (1977). There are surface blockage effects caused by strong surfactants which reduce k_L , although a is very much larger than in coalescing systems. This behaviour was also observed by Raymond and Zieminski (1971) in studies of single bubbles.

2.8.6. Scale-Up

Since mixing in stirred tanks is an empirical science, many processes are scaled up from small-scale experiments to production plant. Incorrect scale-up can be very costly, as shown by Coble (1985), and there can be serious losses of yield on scale-up, as shown theoretically by Mann, Middleton and Parker (1977). Many authors have addressed the problem; some, such as Miura (1976), have produced reviews of literature correlations for $k_L a$ without making clear recommendations on the best approach.

The work of Smith et al (1977) and Davies (1986) shows that, for a reasonable range of scales :

$$k_L a = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (2.53)$$

This clearly implies that $k_L a$ will be constant if gassed power input per unit volume and superficial gas velocity remain constant on scale-up. However, this approach leads to longer gas residence times with increasing scale, and increasing depletion of the gas phase if an inert gas is present. The overall objective is to achieve a constant transfer rate per unit volume :

$$R_v = k_L a \Delta C \quad (2.36)$$

Depletion of the gas phase reduces the volumetric transfer rate if $k_L a$ is constant.

Variations on the approach of power per unit volume and superficial gas velocity have been produced by Figueiredo and Calderbank (1979) and Chandrasekharan and Calderbank (1981).

Figueiredo and Calderbank correlated $k_L a$ as

$$\frac{k_L a V}{T} = 0.001 P_g^{0.58} v_{SG}^{0.75} \quad (2.54)$$

On re-arrangement, this equation shows that at constant $k_L a$ and v_{SG} :

$$\frac{P_g}{V} \propto V^{0.15} \quad (2.55)$$

On scaling up from 0.61 m diameter to 2.67 m diameter, the power input per unit volume would rise by 95%.

Chandrasekharan and Calderbank used four, fairly complex, correlating equations for $k_L a$ based on the observation that $k_L a$ was not kept constant in their experiments by holding P_g/V and v_{SG} (or VVM) constant. There were no significant differences between the models, but the preferred equation was :-

$$k_L a = \frac{0.0333}{T^4} \left(\frac{P_g}{V} \right)^{0.54} Q^{0.54/T} \quad (2.56)$$

This leads to a complex relationship between P_g/V and scale for constant $k_L a$ and v_{SG} . Again, higher power inputs per unit volume are predicted with increasing scale.

Nienow and Wisdom (1978) used an analogy with the scale-up criterion for single phase continuous reactors put forward by Nienow (1974). This involves maintaining the impeller speed constant to maintain the same ratio of residence time to circulation time. The gassing rate was scaled up on the basis of constant gas residence time (vvm) based on the vessel volume, not the gas volume. This leads to very large increases in power per unit volume with scale, but Nienow and Wisdom produced experimental data to support the conclusion. Miller (1974) modified the correlations of Calderbank (1958) for bubble size, gas hold-up and interfacial area. By keeping power per unit volume and superficial gas velocity constant, all three of these parameters are constant on scale-up. Miller argued that since hold-up and interfacial area are related to bubble size, the designer should concentrate on constant bubble size as a scale-up criterion.

There are clear contradictions in the literature data on scale-up. The case of constant $k_L a$ and v_{SG} has been considered here to illustrate the different conclusions. Even a weak relationship between power per unit volume and vessel volume leads to big differences in the predicted power demand at large scale since volume ratios are typically very large.

2.9 Agitator comparison : Mass transfer and hold-up

Much of the published work on gas-liquid mixing has concentrated on Rushton turbines. Other agitator types are widely used in industry for dispersing gas and their performance must be evaluated.

The basis of the comparison is important. The workers who have used power input by the agitator have almost unanimously concluded that the agitator type is of little significance. Table 2.5 is a summary of the references and the types of agitator used. Chapman et al (1983) indicated that the downwards-pumping agitators may be less energy efficient than upwards or radial pumping impellers, but the conclusion was rather tentative. Nienow, Konno and Bujalski (1986) state in a review article that upwards-pumping agitators are more energy efficient than downwards-pumping agitators.

Hassan and Robinson (1977) did not compare performance on the basis of constant power input, but presented correlations for gas hold-up in terms of impeller speed. Table 2.6 summarises their results. The two 6 bladed agitators gave similar hold-up at constant speed whilst the 4 bladed flat paddle, which was twice the diameter of the 6 bladed paddle, produced much more hold-up at constant impeller speed. This may be evidence in favour of constant hold-up for constant power input. Hassan and Robinson did not give enough detail in the paper to make proper comparison possible. Similarly, no conclusion can be drawn from the work of Yung, Wong and Chang (1979), whose correlation

quoted in Section 2.7 covered both Rushton turbines and flat paddles.

The only authors to produce data showing significant differences in performance between agitator types have been Figueiredo and Calderbank (1978), who compared a conventional Rushton turbine with a comb turbine. The comb turbine gave higher values of $k_L a$ and hold-up than the Rushton turbine, especially at high gas hold-ups. This was attributed to the smaller bubble size caused by the shearing action of the comb. Rather surprisingly, no further work on this agitator type has been published.

In the field of fermentation, the situation is rather different. Fermenters tend to be high aspect ratio ($H/T > 1$) tanks with several agitators. Roustan, Charles and Martinet (1978), Ramanarayanan and Sharma (1982), Oosterhuis and Kossen (1983) and Jurevic, Berovic, Steriner and Koloin (1984) have all investigated mass transfer in vessels of this geometry. One of the problems is poor axial mixing of the liquid phase when multiple Rushton turbines are used. Another difficulty is the relatively high shear produced by Rushton turbines which may damage some cultures. Tyler, Henry and Park (1985) used hydrofoil impellers to replace all but the bottom Rushton turbine in both pilot plant and production fermenters. The modified fermenters both showed significant increases in $k_L a$ as a result of the modification. The same result was obtained for a different fermentation by Gbenwoyo, Di Masi and Buckland (1986).

3. EXPERIMENTAL FACILITIES

In this section are described the test facilities and other experimental details common to the experiments described in subsequent chapters.

3.1 VESSELS

A geometrically similar series of three vessels was used for the tests, with diameters ranging from 0.61 m to 2.67 m. The vessels had dished bottoms (Fig 3.1), with the radius of curvature equal to the tank diameter. The 0.61 m vessel was constructed from perspex to allow visual observation and the 1.83 m vessel was equipped with four viewing windows, running from the liquid surface to just above the lower tangent line. The 2.67 m vessel was set below ground level, only permitting observation of the dispersion surface.

3.2 BAFFLES

Each vessel was equipped with four flat baffles of width $T/12$, with a space of $T/60$ between the vessel wall and the inner edge of the baffle. The baffles extended to the lower tangent line. This baffling arrangement was used as being in line with standard industrial practice. It should be noted that many published studies on mixing have been carried out with $T/10$ baffles with no gap at the wall.

The influence of the degree of baffling on gas liquid dispersion processes is poorly-defined. In a study on power numbers in single phase mixing, Bates, Fondy and Corpstein (1963) found a complex relationship between the degree of baffling and power number with the ratio impeller diameter/tank diameter (D/T) being a significant parameter. In another study on single phase power numbers, Muskett (1985) showed that the power number of a $T/3$ pitched blade turbine increased by approximately 10% when the gap at the wall was blocked, giving a $T/10$ baffle. These

observations give few clues to the influence of baffling on dispersion processes. Hemrajani (1985) found significant improvements in the power required to just disperse two immiscible liquids by using partial baffling, but this is still insufficient evidence for any influence of baffle width on gas-liquid dispersion processes.

3.3 IMPELLERS

Four different types of impeller have been used in this study :

- (i) Rushton turbine ($D = T/3$ and $D = 0.47T$) (Fig 3.2).
- (ii) Concave turbine ($D = T/3$) (Fig 3.3)
- (iii) 6 bladed, 45° pitched blade turbine ($D = T/2$) (Fig 3.4)
- (iv) 4 bladed, 45° pitched blade turbine ($D = T/2$) (Fig 3.4).

The actual dimensions of the impellers are given in tables accompanying the relevant figure. In scaling up impellers, every effort has been made to maintain geometric similarity, including minor dimensions such as blade thickness.

The disc turbine impellers were designed according to the following nominal dimensions :-

Impeller diameter	= $D = T/3$ (as standard)
Blade width	= $D/5$
Blade length	= $D/4$
Disc diameter	= $2D/3$
Blade thickness	= $D/60$
Disc thickness	= blade thickness

*Clearance from tank bottom = $T/3$ (in all cases).

These dimensions are in line with the standard adopted for research purposes, first used by Rushton, Costich and Everett (1950), except for the relative disc diameter. Many other workers (such as van 't Riet, Boom and Smith (1976)), have used turbines with discs of diameter $3D/4$. The smaller disc diameter

is frequently used in industrial applications, and has been adopted for that reason. The reasons for the use of different disc diameters are rather obscure.

The orientation of the blade of the concave turbine should be noted. The concave face must be leading. If it is trailing (a convex turbine), the gassed power curves are far less favourable than those of the concave turbine (van't Riet et al, 1976).

The 4 bladed pitched blade turbines were designed using the following nominal dimensions :

Impeller diameter	= $D = T/2$
Absolute blade width	= $2D/7$
Blade thickness	= $D/91$
* Clearance from tank bottom	= $T/3$
Projected blade width	= $D/5$

For practical reasons, the impeller used in the 2.67 m tank consisted of blades bolted onto a hub which was attached to the shaft by a keyway. There was consequently a gap between the inner edge of the blade and the hub. It was reasoned that since the gap occurred at the region of lowest linear velocity of the blade there would be no significant effect on the gas dispersion performance of the agitator. If a single dip pipe placed directly beneath the impeller on the vessel centre-line had been used for gas sparging, the gap may have become important, since gas would be able to travel upwards in the immediate vicinity of the shaft without impedance. For all the tests described in this thesis, a ring sparger was used.

The 6 bladed pitched blade turbine was designed using the following nominal dimensions :-

Impeller diameter	= $D = T/2$
Absolute blade width	= $0.23D$
Blade thickness	= $D/191$
* Clearance from tank bottom	= $T/4$
Projected blade width	= $0.165D$

This agitator was a scaled-down version of an impeller used in a commercial three phase (gas-liquid-solid) reactor (Muskett, 1986).

The pitched blade turbines were not machined to give horizontal top and bottom edges in the fashion of Bates et al (1963). This practice is not in line with conventional industrial impellers.

*Note that the impeller bottom clearance is defined as the vertical distance from the horizontal centre line of the impeller to the lowest point on the dished vessel bottom.

3.4 SPARGERS

Ring spargers were used for all the tests. The PCD of the sparger holes was set at $0.6D$ for the disc turbines and the 4-bladed pitched blade turbine. The centre-line of the sparger was $0.2D$ from the lower edge of the impeller blade. The holes pointed downwards unless stated otherwise. The sparge rings were designed using the design rule recommended by Miller (1978) for manifolds :-

$$\sum A_H = \frac{1}{2} A_m \quad (3.1)$$

$$\sum A_H = \text{Total hole area}$$

$$A_m = \text{Cross-sectional area of manifold}$$

This is an upper limit on hole area to give even distribution at a wide range of operating conditions. In fact, for ring spargers, this is conservative since the flow splits into two on

entering the sparger. Largely for structural reasons, it was decided to use two dip legs fixed to baffles, joining the sparger on opposite sides. This split the flow further, giving another safety margin to the design equation. The flow from the sparge rings can be assumed to have been equally distributed at all times. The actual dimensions of the spargers are given in Table 3.1. The maximum gas exit velocity was roughly constant for each vessel, with velocities not normally exceeding 60 m/s.

A sparge ring with holes set on a PCD of $0.85D$ was used with the 6 blade pitched blade turbine, on the recommendations of Nienow, Konno and Bujalski (1986) for upwards-pumping agitators. The same sparger arrangement was used when testing this impeller in downwards-pumping mode. On the advice of researchers from Birmingham University (Nienow, 1986), the holes were pointing upwards, and were located 0.015 m ($0.05D$) below the lower edge of the impeller blade.

3.5 GAS SUPPLIES

The gas used in all tests was air supplied as needed by oil-free screw compressors. The air flow rate was measured using rotameters for the 0.61 m vessel, and orifice plates for the 1.83 m and 2.67 m vessels. The same compressor was used to supply the two larger tanks, and was purchased specifically for the project. The supply system was designed to give steady gas flow rates by means of a control valve on a venting line to the atmosphere. The settling lengths of pipe upstream and downstream of the orifice plates was in accordance with BS1042 on flow measurement.

At high gas flow rates, and thus relatively high supply pressures, the gas temperature rose significantly above ambient (up to 60°C). This did not cause any large changes in the water temperature because of the low thermal capacity of air in relation to that of water. This point has been discussed in some detail in Appendix 3A.

3.6 TEST LIQUIDS

It was considered to be important to use a test liquid which would give non-coalescing behaviour. The influence of coalescence inhibitors of various types on gas hold-up and mass transfer has already been discussed in the literature survey. Many liquids of industrial interest may be classified as non-coalescing, so the main direction of this research has been to use low viscosity, non-coalescing liquids.

The initial tests were carried out using dilute solutions of octan-1-ol (n-octyl alcohol). This is a surfactant which is effective in reducing coalescence at very low concentrations. However, several problems were encountered whilst using octanol solutions :

- (i) Octanol is biologically degradable when in dilute solution. It was impossible to be absolutely sure of the concentration at any given instance since the only monitoring technique available was surface tension measurements using a du Noi balance. This was not sufficiently accurate.
- (ii) The octanol solutions produced significant quantities of foam, which made hold-up measurement using an ultra-sonic level detector impossible.
- (iii) Some early mass transfer experiments showed that octanol affects the liquid side mass transfer coefficient, k_L , as well as the interfacial area, which is greatly increased by octanol addition (Fig 3.5). The rate of change of $k_L a$ with octanol concentration was greatest in the range of concentrations used for the initial hold-up tests. In view of the difficulty of monitoring octanol concentration, this made octanol solutions unsuitable for $k_L a$ measurement.

- (iv) The local water authority refused permission to discharge octanol solutions into the river system without treatment. Since large quantities of effluent were generated by the testing, this was a significant consideration. The main advantage of octanol solutions over electrolytes was lost.

The mass transfer tests in a non-coalescing liquid were carried out using an aqueous solution of sodium chloride, containing 15g/litre of NaCl. The electrolyte was a commercial grade, supplied in bulk by ICI Mond division, and contained traces of sodium hexacyanoferrate anti-caking agent.

The following terminology has been used in the sections describing the experimental procedures and data :

- a) Octanol solution a solution of octan-1-ol in tap water, containing 15 ppm octanol by weight.
- b) Electrolyte a solution of sodium chloride in tap water, described above.

Tap water was used for the large scale mass transfer tests and for a set of reference tests in the 0.61 m vessel.

All tests were carried out at ambient temperature and pressure.

4. POWER: METHODS AND RESULTS

This section describes the experimental methods used to measure the power drawn by the impeller and presents the experimental data.

4.1 MEASURING METHOD

The shaft power was measured by the simultaneous determination of the shaft rotational speed and the impeller torque. Power was then calculated as :

$$P = 2\pi TN \quad (4.1)$$

The shaft speed was measured using an optical tachometer and a toothed wheel fixed either to the impeller shaft (0.61 m and 1.83 m vessels) or to the input shaft to the gearbox (2.67 m vessel). The precision of the measurement was ± 1 rpm, ± 0.5 rpm and ± 0.2 rpm for the 0.61 m, 1.83 m and 2.67 m vessels respectively.

The impeller torque was measured using strain gauges located above the impeller. Semi-conductor strain gauges were used instead of the more traditional foil gauges because of their greater sensitivity to strain. It was not possible to turn down the shaft at the measuring section because the critical speed of the shaft would have become so low that the operating range of the vessel would have been severely restricted. The torque measurements accompanying the mass transfer tests in the 2.67 m vessel were made using foil gauges after experience had shown that the additional sensitivity theoretically gained from using semi-conductor gauges was severely restricted by the maximum current through the measuring bridge. The additional time and effort required to install semi-conductor gauges was such that, on balance, it was best to revert to the use of foil gauges.

The transmission system is illustrated in Fig 4.1. For the 0.61 m and early 2.67 m tests, telemetry was used to transmit the signal to stationary pick-ups located on the vessel frame. The FM signal was converted back to voltage modulation before being amplified and filtered prior to sampling on a PDP 11/23 computer. The PDP 11 was fitted with a 12 bit analogue to digital converter giving a resolution of 1 in 4096 over a range of $\pm 5V$. The digitised signal was sampled using a proprietary DEC subroutine and stored as unformatted binary data files on disk for subsequent analysis. Essentially, the system was the same as that described by Pollard (1983). The transmitter and de-modulation units were manufactured by Astech Electronics Ltd, UK and the filter circuits were manufactured by Kemo Ltd, UK.

The laboratory in which the experiments were carried out contained several mixing vessels, and there were occasions when more than one vessel was in use. It appeared that the thyristor drive units, used for speed control, on neighbouring vessels interfered with the telemetry system. Fortunately, the thyristor drive of the vessel in use for torque measurement did not interfere with the telemetry system on the same vessel. This was tested by running the shaft in air with no impeller. No fluctuation of the signal was observed.

For this reason, low noise slip rings were installed on the larger vessels. The instrument chain is illustrated in Fig 4.2. The slip ring systems were immune to external interference.

The interference problem with the Astech equipment did not affect the quality of the data reported here. The problem was quickly diagnosed, and torque measurements were made only when all other vessels were inoperative.

The strain gauges and instrument chain were calibrated using known loads on a lever arm. For the large vessels, this was done with the shaft in place. A force was applied to a lever

arm and measured using a load cell, which was checked against standard weights before the calibration took place. In all cases the voltage output was linear with applied torque. The units in the instrument chain were individually checked at regular intervals, particularly the filter units. As part of the calibration procedure, a pure bending load was applied to the shaft and the output from the torque gauges monitored. For correctly aligned strain gauges, all of which are properly attached to the shaft, the signal from the torque strain gauge bridge should be zero for all bending loads. This is a useful check on the integrity of the system, since cross sensitivity to bending stresses can cause erroneous torque measurement if significant steady bending loads are present, which is often the case as few mixer shafts, even in laboratories, are dynamically balanced. To supplement the calibration procedure, standard tests were run at regular intervals using a particular impeller type which would give a known result.

The analysis consisted simply of subtracting the zero level from the data and averaging the readings using the computer. On the small scale, 6 windows of 2048 data points sampled at 100 Hz were used. On the large scale, 6 windows sampled at 50 Hz were used. This was considered to be sufficient to produce an accurate average value. The filter unit was set at one quarter of the sampling frequency (on low pass) to remove high frequency noise. The sampling frequency/filter frequency ratio was set by experience with the measurement of shaft bending moments (Pollard, 1983). To prevent signal aliasing, the filter frequency must be less than half the sampling frequency.

The additional factor of two was a safety margin to allow for the filter 'roll-off', the filter does not have an absolute cut-off frequency. In fact, aliasing is only a problem when considering the frequency domain but it seemed prudent to follow the established procedure and the removal of high frequency noise was necessary in any case.

4.2 ACCURACY OF METHOD

The direct measurement of the torque produced as a result of the impeller rotating in the fluid is the most accurate measuring method available. It is superior to dynamometer techniques (discussed in some detail by Nagata, 1975) or measurements of power drawn by the electric motor (used by Brown, 1981).

However, great care is needed in applying the method, especially during calibration. The only check available is repeatability of tests. Some repeat results are discussed in the following section. As a rough guide the accuracy of the torque data is estimated to be in the region of $\pm 2\%$. Allowing for some error in speed measurement, the power is estimated to be accurate to $\pm 3\%$.

4.3 RESULTS

4.3.1 Ungassed Power

For turbulent flow ($Re > 10^4$), the general conclusion in the literature is that power number is independent of Reynolds number. Power number is defined as :-

$$Po = \frac{P}{\rho N^3 D^5} = \frac{2\pi\tau}{\rho N^2 D^5} \quad (4.2)$$

The power number is a function of impeller type and geometric parameters such as blade width, blade number and blade angle (Bates, Fondy and Corpstein, 1963 and Rushton, Costich and Everett, 1950). There are complex interactions between the parameters which mean that no simple relationship between power number and dimensionless geometric parameters may be obtained, as found by O'Kane (1974).

The ungassed power numbers obtained in this study are shown in Tables 4.1 to 4.3. Each power number is the average of at least four impeller speeds. A typical plot of power number against Reynolds number is given in Fig 4.3. The plots do not show

maxima, as reported by Bujalski, Nienow, Chatwin and Cooke (1986) and by Nienow, Wisdom and Middleton (1977). All tests were carried at speeds below the minimum speed for surface aeration. This was checked visually in the smaller vessels. Severe surface aeration causes the power number to fall as gas cavities begin to form behind the stirrer blades.

(a) Rushton Turbine

The most striking feature of the Rushton turbine power number data (Table 4.1) is the increase of power number with increasing scale. The effect cannot be attributed to geometric changes since the ratios $w : D$ and $L : D$ (where L = blade length) were maintained constant to within $\frac{1}{2}\%$ for the three impellers. The largest diameter Rushton turbine had a lower blade thickness/impeller diameter ratio ($x_2 : D$) than the smaller impellers (0.0155 instead of 0.0168). From the data assembled by Palmer (1986), largely on the basis of the work of Bujalski (1987):

$$Po \propto \left(\frac{x_2}{D}\right)^{-0.2} \quad (4.3)$$

Correcting the large scale power number using this equation gives the correlation for $\frac{x_2}{D} = \frac{1}{60}$

$$Po = 5.22 T^{0.06} \quad (4.4)$$

This confirms the result of Bujalski et al, who reported that :

$$Po \propto T^{0.065} \quad (4.5)$$

The larger Rushton turbine ($D = 0.47T$) gave a similar power number to the T/3 impeller, in line with other workers (Bujalski et al) who have found power number to be independent of D/T for Rushton turbines.

The large scale data show two power numbers, measured using completely different strain gauges and transmission techniques. The agreement between the two results is striking and is an indication of the accuracy of the data.

It is useful to compare the absolute values of power number measured in this study with data published by other workers. This must be done with considerable care, since the geometry used by other researchers has often been inadequately specified. This causes problems because of the large number of independent variables which can have some effect on power number. Bujalski et al produced a correlation for the turbulent, fully baffled power number based on very careful, accurate measurements using closely-specified agitators:

$$Po = 2.5 \left(\frac{x_1}{D}\right)^{-0.2} T^{0.065}$$

where x_1 is the thickness of the disc

x_1 = thickness of disc

The following conditions applied for this correlation:

- i) Impeller bottom clearance, $C = T/4$
- ii) Baffle width = $T/10$ (4 flat baffles, no wall clearance)
- iii) Flat bottomed vessels.

Using the 0.61m vessel, T/3 Rushton turbine geometry from this study, a power number of 5.49 is predicted by the correlation. The measured power number was 5.1. For this study:

- i) Impeller bottom clearance, $C = T/3$
- ii) Baffle width = $T/12$ (T/60 wall clearance)
- iii) Dished bottom.

Bates, Fondy and Corpstein (1963) reported that the power number was lowered by approximately 4% for T/12 baffles compared to T/10 baffles, and also that power number tended to increase with increasing clearance from the vessel bottom. Presumably the

effect of a dished bottom is to reduce the effective clearance, a conclusion drawn by Palmer (1986), after an extensive review of the literature on power numbers. The interplay between variables is obviously complex, especially when proximity effects of the vessel bottom are considered. If one reduces the predicted power number from Bujalski et al for the effect of baffle width the predicted value of power number is within $3\frac{1}{2}\%$ of the measured value.

In view of the agreement both in scale effect and in absolute values with the data of Bujalski et al, further detailed comparison with literature data seems to be of little value. The comparison serves as a reference point to check the overall values of power number.

(b) Concave Turbine

The concave turbine power numbers are given in Table 4.2. The 1.83 m and 2.67 m vessels gave power numbers in reasonable agreement with each other. The 0.61 m result was rather lower and there was a discrepancy between two repeated tests. The same repeat test on the Rushton turbine in this vessel (used to check the measuring system) had showed no change in the overall calibration factor. The difference between small scale and large scale results were probably due to changes in the sharpness of the horizontal blade edges or other minor geometrical changes, which have not been characterised in this study. The geometric ratios $x_1 : D$, $w : D$ and $L : D$ were all within a narrow range, and variations in these parameters do not explain the difference between small-scale and large-scale data.

The only published data on the concave turbine is that of van't Riet, Boom and Smith (1976). The concave power number was given as $2/3$ of the power number of an otherwise identical Rushton turbine. This fits the current data at small scale, but underpredicts the concave turbine power number at large scale.

(c) Pitched Blade Turbines

The power number data for both the 4-bladed and 6-bladed pitched blade turbines are given in Table 4.3. The larger scale result for the 4 bladed agitator was rather higher than the corresponding power number at small scale. There were several experimental uncertainties involved in the smaller scale measurement, and it would be wisest to conclude that there is no significant influence of scale on power number for pitched blade turbines. Much of the data produced by BHRA's Fluid Mixing Processes (FMP) service, of which this project has been an integral part, is conflicting in this respect.

Recent data which was corroborated by the repeated power number results for the Rushton and concave turbines in Tables 4.1 and 4.2 has shown that there is no scale effect, and the same conclusion was drawn by Bujalski et al.

The six bladed agitator was tested at one scale ($T = 0.61\text{m}$) for both upwards and downwards pumping. The upwards-pumping power number was approximately 11% lower than the downwards pumping power number. This can be related to the different proximity effects of the vessel bottom on the two pumping directions. Pitched blade agitators increase power number as the bottom clearance decreases, as shown in FMP research by Muskett and Ruszkowski (1985), for a $T/3$ agitator. Simplistically, this is explained as an increase in the discharge pressure required to circulate the liquid as the impeller outlet becomes constricted by the approaching vessel bottom. There is also an increase in the radial component of the flow, signifying less streamlining of the flow over the impeller blades. The impeller begins to

act as an agitator with a steeper blade angle to the horizontal, producing an increase in power number.

Reversing the direction of rotation has two effects :

- (i) The bottom proximity effect disappears.
- (ii) The inlet stream to the impeller is influenced by the vessel bottom.

The upwards pumping flow pattern is not simply the reverse of the downwards-pumping case, when the outlet stream is influenced by the vessel bottom. Fig 4.4 gives a fairly crude impression of the flow patterns produced by the two pumping modes. The upwards-pumping flow pattern was rather unexpected, with upflow at the tank wall and down-flow at the vessel centre-line in the upper regions of the vessel.

Largely because of the proximity effect, one would intuitively expect the upwards-pumping power number to be lower than the downwards-pumping power number, in line with the experimental data. Chapman et al (1983) published some data on upwards-pumping agitators. From their graphs of power number against Reynolds number, it is apparent that the upwards-pumping power was some 6% less than the downwards-pumping power number. This conflicts with the assertion of Nienow, Konno and Bujalski (1986) based on the experimental data of Bujalski (1987) that the upwards and downwards-pumping agitators have the same power number. The degree of the reduction in power demand is probably dependent on bottom clearance and bottom shape for low clearances, especially for large agitators.

As for disc turbines, there is extensive information available in the literature on ungassed power numbers, but little real understanding of the effects of some independent parameters on power number. The FMP programme included the measurement of ungassed power number of pitched blade turbines of several geometries. This work has been reported by Muskett and Ruszkowski (1985), and has been incorporated into the design

guide produced by Palmer (1986), which includes substantial quantities of data from the open literature. If anything the exact prediction of power number for pitched blade turbines is even more problematic than for disc turbines, largely because of the variety of geometries found for pitched blade turbines.

For the 4PBT, there are two impellers from the FMP study (which were measured in a different 0.61m vessel) which may be used for comparison. For $D = T/2$, $w = 0.187D$, Muskett and Ruszkowski reported a power number of 1.12 with $C = T/3$. O'Kane (1974) found that for 4-bladed 45° agitators:

$$Po = K_4 \left(\frac{w}{D}\right)^{0.65} \quad (4.6)$$

Using this correction factor gives a predicted Po of 1.46 for the 4PBT, $T = 0.61m$.

For $D = T/3$, $w = 0.28D$, the FMP data give a power number of 1.56. This is the same w/D and blade thickness as the 4PBT used in this study. Bujalski (1987) gave the following correlations for Po of 6-bladed PBT agitators:

$$Po = 1.54 \left(\frac{D}{T}\right)^{-0.127} T^{0.078} \quad (4.7)$$

$$Po = 0.775 \left(\frac{D}{T}\right)^{-0.171} \left(\frac{x_2}{D^2}\right)^{-0.143} \quad (4.8)$$

x_2 = blade thickness

The first equation links the two FMP data points much better than the second, since for constant x_2 , equation (4.8) reduces to, for a single tank size:

$$Po \propto D^{-0.03}$$

Other FMP data show that there is an influence of (D/T) at constant scale and blade thickness, which supports the alternative form of the correlation. Bujalski recommended

equation (4.8). The situation is far too complex to be resolved on the basis of a few data points.

The 6PBT pumping downwards can be compared with a 6 bladed, $D = T/2$ impeller with $w = 0.187D$ used in the FMP study. From the data of Bates et al (1963), O'Connell and Mack (1950) and O'Kane (1974), the following relationship hold for 6-bladed impellers:

$$Po = K_6 \left(\frac{w}{D} \right) \quad (4.10)$$

Also, the effect of impeller clearance is given by:

$$Po = K'_6 \left(\frac{C}{D} \right)^{-0.2} \quad (4.11)$$

This result has been obtained by Bates et al, Muskett and Ruszkowski, and Frijlink, Koliijn and Smith (1984). The exponent on clearance does not appear to depend on blade number or other geometric parameters. The power number from the FMP study was 1.41 : correcting for blade width and clearance effects gives a predicted power number of 1.86 (compared to a measured value of 1.96). Equations (4.7) and (4.8) predict power numbers of 1.34 and 1.53 respectively (a correction for blade width has been made in these estimates).

The large differences between the data of this study and of Bujalski for the 6PBT are probably due to:

1. Bottom clearance effects. As for disc turbines the dished bottom reduces the effective bottom clearance, giving an increased power number compared with a flat-bottomed tank.
2. The ratio (x_2/D) for the 6PBT was considerably less than the minimum value used by Bujalski (0.0052 compared to 0.0096). Thus, this agitator was outside the range of the correlation. It is a moot point whether (x_2/D) should be used to characterise blade thickness. (x_2/W) might be a more appropriate quantity.

3. Bujalski used impellers with chamfered horizontal edges, giving a vertical section through the blade which was trapezoidal in shape. This probably influences flow over the leading edge of the blade and, hence, could influence the power number.

In fact, Bujalski's correlations also underpredict the power number for the 4PBT if the correction for blade number recommended by Fort, Medek, Placek and Hulamek (1975) is used:

$$Po \propto n_B^{0.64} \quad (4.12)$$

This relationship was in approximate agreement with the FMP data reported by Muskett and Ruszkowski.

4.3.2 GASED POWER

(a) Rushton turbine

A set of gassed power measurements has been obtained in octanol solution at 0.61 m, 1.83 m and 2.67 m scales (Figs 4.5 to 4.8) for the T/3 Rushton turbine. The data were all accumulated by keeping impeller speed constant and varying the gas flow rate to vary the flow number.

Qualitatively, the power curves in each vessel show the characteristics described by Warmoeskerken and Smith (1982). There is a steep fall in power at low flow numbers, with the curve flattening as large cavities are formed and the flooding point is approached. The curves for different impeller speeds are distinct from each other, showing that flow number alone is not an adequate correlating parameter for gassed power. Higher impeller speeds tend to produce lower values of gassed power ratio at constant flow number and constant scale. This is due to two factors :

- (i) Re-circulation of gas at low flow numbers causes the actual gas throughput to exceed the gas throughput from the sparger. This reduces gassed power demand and since the extent of re-circulation increases with increasing impeller speed, the fall in gassed power with increasing flow number became sharper with increasing impeller speed.
- (ii) At high flow numbers the size of large cavities is a function of speed. At high impeller speeds, the cavities are larger than at low speeds, giving lower values of gassed power. There is also probably an effect of gas hold-up. From the data of Hassan and Robinson (1977) and Warmoeskerken (1986) as a rough approximation :

$$\epsilon_G \propto Q^{0.5} N \quad (4.13)$$

$$Fl = \frac{Q}{ND^3} = \text{constant}$$

$$\therefore \epsilon_G \propto (FlND^3)^{0.5} N \quad (4.14)$$

$$\epsilon_G \propto N^{1.5} \quad (4.15)$$

Thus, at constant flow number, gas hold-up is quite a strong function of impeller speed, with hold-up increasing with increasing speed. The effective density of the liquid is reduced as hold-up increases, producing a fall in the relative gassed power demand. The argument can be applied across the whole range of flow numbers. The effect of gas hold-up is most probably secondary to the effects of cavity transitions and cavity sizes.

The influence of scale on gassed power curves is of particular concern to the designer. Since all the correlations for flow regime transitions except the clinging - "3-3" transition relate flow number to Froude number, it may be argued that a plot of gassed power ratio against flow number at constant Froude number at each scale would be a valid way of showing any changes in

hydrodynamics with scale. The resulting curves are shown in Fig 4.9 for a Froude number of 0.2. The clinging to "3-3" transition is predicted to take place at higher flow numbers as scale increases. This tendency is shown in Fig 4.9. The correlation of Warmoeskerken (1986), has been validated, at least qualitatively. At high flow numbers, the power curves tend to converge, a trend which is shown by the equivalent plot at higher Froude number (Fig 4.10). It was not possible to achieve Froude numbers of 0.42 in the 2.67 m vessel because of equipment limitations.

In practice, scale-up at constant Froude number is not used because it leads to unreasonably high power demands in large vessels. Scale-up at constant power input per unit volume is widely used and Fig 4.11 is a comparison of gassed power curves at an ungassed power input of 1.0 W/kg at 0.61 m and 2.67 m scales. The larger scale curve gives higher values of P_g/P_u at all flow numbers, which is consistent with the observations made above about the possible relationship between Froude number and P_g/P_u at high flow numbers, since Fr is lower for the larger scale vessel.

The influence of liquid type on the gassed power curve is shown in Fig 4.12. The differences between the water and electrolyte curves were less than the estimated experimental error. It is probable that differences would have been evident had lower flow numbers been investigated, but the measurement of gassed power was no longer a priority when the measurements in electrolyte were made. The complete set of gassed power measurements in electrolyte are given in Fig 4.13. The data effectively fall on the same curve, with the higher speeds giving slightly higher values of gassed power ratio at high flow numbers, in contrast to the trends discussed above.

Finally, the data for the Rushton turbine of 0.47T diameter in electrolyte at 0.61 m scale are shown in Fig 4.14. The data show the expected trends, and are plotted on the same axis as

the T/3 data in Fig 4.15. The experiments at approximately constant Froude number (albeit one in octanol and one in electrolyte) show approximately the same rate of fall in P_g/P_u at low flow numbers supporting the hypothesis that Froude number is a controlling parameter on gassed power demand, at least at one scale.

(b) Concave Turbine

The power curves obtained for the concave turbine in octanol solution at 0.61 m, 1.83 m and 2.67 m scales are shown in Figs 4.16 to 4.18. All the data show the fall in gassed power ratio with increasing gassing rate becoming more rapid with increasing speed. This effect was noted by van't Riet et al (1976) and is attributed to re-circulation of gas. Although knowledge of cavity formation behind the blades of concave turbines is lacking, it might reasonably be supposed that cavities form and that their size is related to agitator speed.

If the Rushton and concave turbines are compared at a representative speed for each vessel, the concave turbine always shows higher values of P_g/P_u than the Rushton at a given flow number (Figs 4.19 to 4.21) but the comparison does not generally look as favourable as that of van't Riet et al. This is due to the Rushton turbine curves being flatter than expected, especially at low flow numbers, and the concave turbine power curves falling to lower values of P_g/P_u than expected.

The plots of concave turbine gassed power ratios at constant Froude number are shown in Figs 4.22 and 4.23. The relationships between the curves from the various scales is confusing, implying possible effects of minor geometric parameters such as blade sharpness along the horizontal edges. The similarity between the two plots shows that the scale effects are at least consistent, even if no explanation is available. The possible influence of minor geometric parameters is open to dispute, since other workers now use concave turbines

with blades which are semi-circular in section as opposed to a 120° arc. However, there were also inconsistencies in the ungassed power number at different scales which are only explicable in terms of minor differences in blade geometry.

The concave turbine also showed some slightly surprising effects of liquid type. Fig 4.24, from the 0.61 m vessel, shows the power data obtained using tap water, octanol and electrolyte. The tap water data more closely resembles the data of van't Riet et al (1976), and at the lowest flow number, the gassed power did slightly exceed the ungassed power. The electrolyte data are in good agreement with the octanol results, indicating there is relatively little influence of fluid type on the power demand of a gassed concave turbine. A complete set of power curves was obtained in the electrolyte solution. The data are presented in Fig 4.25.

(c) Pitched Blade Turbines

The 4 blade pitched blade turbine (4PBT) was tested in both the 0.61 m and 2.67 m vessels (Figs 4.26 and 4.27) in octanol solutions. The curves both show the characteristic shapes described by Warmoeskerken, Speur and Smith (1984) and Chapman et al (1983). There was a rapid fall in power until the transition from indirect loading was reached, after which the power increased slightly. The rise in power at high flow numbers is assumed to be caused by liquid swept upwards by the gas from the sparger into the impeller, forcing the impeller to pump against an upflow of liquid. This outweighs the effect of increasing cavity size, which tends to reduce the power consumption. It must be emphasised that visual observation of cavity formation was not possible in these experiments. All comments about the indirect to direct loading transition are made on the basis that this takes place at the minimum in the power curve.

As with the disc turbines, flow number is not a sufficient parameter to correlate the gassed power demand. The effect of speed on relative power demand is complex, but in general, at high flow numbers, the higher the speed the lower the relative power demand. The maximum large cavity size is again apparently related to impeller speed. Attempts to obtain a correlation between Froude number and gassed power ratio in the direct loading regime have not been successful, largely because of the scatter in the data and the slope of the lines. Linear regression produced lines with very different slopes, so the dependency of P_g/P_u on Fr became very sensitive to the value of flow number at which the values of P_g/P_u were taken for correlation. There is certainly an inverse relationship of some description between Froude number and gassed power ratio.

The point at which the transition to direct loading takes place is of interest because the gas distribution in the vessel becomes much poorer when the impeller is directly loaded. Warmoeskerken et al (1984) and Chapman et al (1983), both produced equations to relate the speed at which the transition took place to the gas flow rate :

$$\text{Warmoeskerken et al : } N_L (\sin\theta) = A Q_L^{\frac{1}{2}} \quad (4.16)$$

$$\text{Chapman et al : } N_L = \frac{11.7 Q_L^{\frac{1}{2}}}{D^{1.63}} \quad (4.17)$$

The constant A depends on scale and sparger geometry, including spacing between the sparger and the impeller. Chapman et al (1983) verified their equation for 45° impellers whilst the correlation of Warmoeskerken et al includes the blade angle to the horizontal, θ .

By taking the flow number at the minimum in the gassed power curve the points (N_L, Q_L) were estimated for all but the lower agitator speed in the 0.61 m tank. This is rather prone to error, but the objective was to obtain a qualitative assessment of scale effects.

Using this procedure the following equation was obtained (based on only 4 data points) :

$$N_L = \frac{8.8Q^{0.53}}{D^{2.1}} \quad (4.18)$$

The agreement with the literature data is good considering the crudeness of the methodology.

The 6-bladed agitator was tested in both downwards and upwards mode (Figs 4.28 and 4.29). The downwards-pumping results show a very steep fall in P_g/P_u which is probably due to two factors:

- (i) The blade width of the 6 bladed PBT was rather less than the 4PBT. This outweighed the effect of increasing the number of blades.
- (ii) The sparger clearance was very small. The data of Frijlink, Kolijn and Smith (1984) and Warmoeskerken et al (1984) show that the flow regime transition takes place at lower flow numbers as sparger clearance from the impeller decreases.

The upwards-pumping power curves do not show any of the very steep falls in gassed power ratio exhibited by downwards-pumping agitators. The influence of speed is again clear. The curves are consistent with the gradual growth of cavities. A notable feature of the curve at the lowest agitator speed is that $P_g > P_u$ at low flow numbers. This is counter-intuitive, and no ready explanation is available.

The dispersion showed no tendency towards violent oscillations in level which can occur with downwards-pumping agitators in the direct loading regime when large pockets of gas appear due to poor dispersion by the agitator. No phenomenon analagous to flooding was observed during the power measurements, even though flow rates up to 3.8 VVM were used.

Both the flat power curves and the apparent lack of a flooding point are important operational features which are discussed in a later section.

5. OVERALL HOLD-UP

Extensive measurements of overall hold-up have been made to give an indication of agitator dispersion performance from experiments which are not subject to the assumptions and experimental problems associated with mass transfer tests.

Hold-up is not as sensitive to external parameters like fluid contamination as $k_L a$ and is from the experimental point of view, much easier to handle. This section describes the measuring techniques used and also presents the raw data along with discussion of the correlating methods available.

5.1 MEASURING METHODS

(a) Spill-Over Method

The tests in octanol in all three vessels and the tests in tap water in the 1.83 m tank were carried out using the spill-over method illustrated in Fig 5.1 for the 0.61 m and 1.83 m tanks, and Fig 5.2 for the 2.67 m tank. The 0.61 m and 1.83 m tanks were equipped with a re-circulation line to return liquid from the measuring tank. A liquid flow rate of not more than 1.0 litre/sec used in the re-circulating line, to ensure that liquid re-circulation effects did not influence the measurement of hold-up. The effect of the re-circulation was to damp out the effects of surface waves, if no re-circulation were present, the apparent hold-up would be determined by the largest wave to appear in the vessel. In some cases the mean dispersion level can be well below the level of the funnel, giving a false value of hold-up.

The diameters of the funnels were 0.04 m and 0.12 m for the 0.61 m and 1.83 m tanks respectively.

The measuring vessels were of smaller cross-sectional area than the mixing vessels to maximise the level change caused by displacement of liquid from the mixing vessel. The changes in

level were measured using a point gauge in the 0.61 m experiments, and using measuring tapes in the larger vessels.

The procedure adopted for the 0.61 m and 1.83 m tanks was as follows :-

1. Measure zero level in measuring tank with liquid re-circulation and the impeller running. Check that the funnel is acting as a weir and is not completely submerged.
2. Start gas flow, set gas flow and impeller speed to correct conditions.
3. Wait for dispersion level to fall to point where funnel is acting as a weir again. Measure new level in measuring tank, check after one or two minutes to make sure that system is at equilibrium. Calculate volume of liquid displaced into measuring tank.
4. Proceed to next gas flow rate.
5. At the end of the series of tests at different flow rates, constant impeller speed, stop gassing and check that liquid level returns to original zero value.

For the 2.67 m tank, it was not possible to use a re-circulating system because the tank was below ground level. Fig 5.2 shows the manifold which was positioned at the desired dispersion level. The experimental procedure was as follows :-

1. The mixing vessel was filled with liquid above the height of the manifold.
2. The liquid was pumped into the measuring tank using a rotating vane pump until the level reached the height of the manifold. At this point pumping ceased abruptly. This gave the zero level for the volume measurement.

3. Gas was then introduced into the vessel, causing the dispersion level to rise above the level of the manifold. The pump was switched on again and the excess liquid volume removed to the measuring tank. Provided the dispersion surface was reasonably level, a clear end point was obtained; pumping stopped when the dispersion level again reached the manifold. The displaced liquid volume was determined from the level change in the measuring tank.

The measurement of hold-up was limited to conditions for which the dispersion level was relatively stable. Large surface waves cause errors in the measurement, but do not appear for non-flooded conditions, except under conditions of very intense agitation. This limitation holds for virtually all methods of measuring overall gas hold-up.

(b) Level Change Method

For the tests using water in the 0.61 m and 2.67 m vessels and using electrolyte in the 0.61 m vessel, an ultra-sonic level detector was used to determine the change in dispersion level caused by gassing. The device used was a Nivosonic FMU 2780 gauge with a DU213 ultra-sonic head manufactured by Endress and Hauser Ltd, UK. This was the same type of unit as that used by Davies (1986) and Bujalski (1987).

The level detector was calibrated by filling the vessel with water to a height exceeding that reached by the dispersion level at the highest hold-up anticipated. The level was then reduced incrementally until the still liquid level for testing was reached. The gauge output was thus related to distance. In all cases, excellent linearity was achieved.

It was noticed that whilst conducting some exploratory experiments the ultra-sonic level detector appeared to be under-estimating the gas hold-up, especially at low hold-ups. Further investigation showed that the unit produced a change in output

when the agitator was rotating, but with no gas entering the vessel. This corresponded to a level change of approximately 5 mm, and the error was virtually independent of impeller speed. This was attributed to surface waves influencing the ultra-sonic pulses and was significantly greater than the inherent inaccuracy of the instrument (± 2 mm). Clearly some correction was necessary, but it was not immediately clear how to proceed. The error for an ungassed, agitated surface was not necessarily that for a gassed, agitated surface.

To resolve the problem, a series of measurements was carried out involving the simultaneous estimation of hold-up by visual observation, and by using the ultra-sonic level detector zeroed on a moving, ungassed surface. The zero level was taken at the same impeller speed as that at which the test was carried out unless this was so high that significant surface aeration took place. For high agitator speeds, the highest speed possible without producing surface aeration was used to zero the instrument. Visual observation was carried out by marking the estimated liquid level in each quadrant of the tank using clear adhesive tape in areas away from the baffles. The four estimates generally agreed very well with each other, and were regarded as providing a reliable indication of the true position of the liquid surface.

The results of this experiment are given in Fig 5.4. The agreement between the two sets of data was remarkably good, and was taken as a validation of the ultra-sonic level detector as a device for measuring hold-up.

Davies (1986) found that the unit estimated hold-up in foaming systems quite poorly. The hold-up obtained from ultra-sound was intermediate between that obtained from visual observation of the lower surface of the foam layer and that obtained from the top surface of the foam layer. The magnitude of the error is dependent on the nature of the foam, which is itself a notoriously difficult parameter to quantify. A few preliminary

experiments with octanol solution, which can produce several centimetres of foam at high gassing rates, quickly confirmed the trends observed by Davies. It was decided to use the ultra-sonic level detector for the tests in electrolyte since the electrolyte produced only a light foam. This was supported by periodic visual observation to check that false readings were not being obtained.

The position of the ultra-sonic head is given in Fig 5.3. The detector was placed mid-way between two baffles at a radius from the shaft of half a tank diameter. A crown tube was fitted over the unit to reduce errors from stray echoes, such as from the tank wall. This was a section of uPVC pipe 76 mm internal diameter, with the end furthest from the ultra-sonic head cut off at 45° , in accordance with the manufacturers recommendations. The length of the tube was 0.45 m, equal to the 'blocking distance' of the device, which is the minimum distance from the ultra-sonic head that the unit is capable of measuring.

Data acquisition was carried out using a PDP 11/23 laboratory computer. The logic of the program, which also sets out the experimental procedure, is given in Fig 5.5.

Data was collected in four windows of 1024 samples, each window lasting approximately 60 seconds. The signal was first filtered using a low pass filter set at 1.0 Hz to remove instrument noise. Low frequency oscillations of the signal were observed, which were due to oscillations of the dispersion surface itself rather than instrument noise. The zero level was periodically checked to ensure that no drift was taking place.

Although the ultra-sonic level detector was extensively evaluated and used carefully, problems were still encountered with the measurement of hold-up in electrolyte solutions. This is discussed in detail in Appendix 5A.

(c) Comparison of Measuring Methods

In order to use data collected using different measuring techniques for scale-up and agitator comparison, it was essential to make a set of measurements to compare the spill-over and level-change methods directly. This was done in the 1.83 m tank, using a T/3 Rushton turbine in tap water. A series of hold-up measurements using the spill-over method had already been carried out, and the collection system was simply disconnected and blocked off to allow the dispersion level to change. The upper limit of hold-up measureable by the level change method was restricted by the tank geometry, but a fairly representative range of hold-up was covered.

The results of this experiment are given in Table 5.1. The agreement between the two sets of data is within experimental error, with the ultra-sonic level detector estimate being the higher of the two at higher hold-ups. This was attributed to the change in nature of the dispersion surface as the hold-up increases. At high hold-up, gas which is disengaging from the surface tends to coalesce into large pockets which eventually burst, releasing gas into the head space above the dispersion. The local gas concentration at the surface becomes relatively high, and the layer of bubbles becomes analagous to a very light foam. As mentioned previously, foam is not transparent to the ultra-sonic level detector, so one would expect the ultra-sonic estimate of hold-up to exceed that of the spill-over method.

The agreement between the two methods also implies that there is little influence on hold-up of small changes in still liquid height. The spill-over technique effectively conducts the experiment with the still liquid level at some value less than the nominal fill level. The level change method utilises a still liquid height equal to the nominal fill level (in this case, one tank diameter from the vessel bottom).

In fact the trend shown by the data can be rationalised quantitatively when one considers the effect of the gas being distributed over only part of the vessel. For coalescing liquids, there exists a speed, defined by Nienow et al (1977) as the speed for complete dispersion, below which gas is not dispersed into the region below the agitator. Referring to Fig 5.6, consider the vessel to be split, in the case of the spill-over method into a volume V_1 , which contains no gas, and a volume V_2 in which gas is uniformly distributed with hold-up e_g .

By definition :

$$e_g = \frac{V_G}{V_2} \quad (5.1)$$

The gas volume, V_G , is measured directly, and a hold-up calculated according to the relation :

$$\epsilon_G = \frac{V_G}{V_D} = \frac{V_G}{V_1 + V_2} \quad (5.2)$$

From equations (5.1) and (5.2) :

$$\epsilon_G = \frac{V_2}{V_1 + V_2} e_g \quad (5.3)$$

For the case of the level change method, a third volume, V_3 , must be considered, which is just the change in volume due to gassing. If it assumed that within the volume $(V_2 + V_3)$ gas is uniformly distributed with hold-up e_g , the true gas hold-up in the vessel above the agitator is given by :

$$e_g = \frac{V_3}{V_2 + V_3} \quad (5.4)$$

But the measured gas hold-up is calculated from :

$$\epsilon_G = \frac{V_3}{V_1 + V_2 + V_3} \quad (5.5)$$

Eliminating V_3 from equations (5.4) and (5.5) :

$$\frac{\epsilon_G}{1 - \epsilon_G} = \frac{V_2}{V_1 + V_2} \frac{e_g}{1 - e_g} \quad (5.6)$$

Equations (5.3) and (5.6) relate the measured hold-up, ϵ_G , to the supposed true hold-up, e_g in the part of vessel in which gas is dispersed. Note that for the case of $V_1 = 0$ (uniform dispersion) the measured hold-up becomes equal to the true hold-up in each case. However, when the dispersion of gas is confined to the region above the agitator, the value of

$\frac{V_2}{V_1 + V_2}$ is 0.713 for a vessel with a dished bottom and

$C = T/3$. If, in this situation, the spill-over method measured a hold-up of 0.1, the true hold-up, from equation (5.3), is 0.14. Substituting these values into equation (5.6) shows that the ultra-sonic level detector would measure a hold-up of 0.104, 4% higher than the spill-over method. In fact, level change methods will always measure higher hold-ups than spill-over methods if the gas distribution is not uniform, provided the assumption that the true gas hold-up in the two-phase region of the tank is constant holds true. The experimental data support this tentative assumption although other effects, such as gas disengagement and energy dissipation rate effects, are also present and prevent definite conclusions from being drawn.

The correlation of Nienow et al (1978) for the minimum speed for complete dispersion predicts that no gas would be distributed into the lower region of the vessel for the test at approximately 10% hold-up in Table 5.1, so distribution effects should certainly have been present in this set of data.

For the purposes of scale-up and agitator comparison, the agreement between the two methods is satisfactory provided that the hold-up is not too great, although some systematic effects of inhomogeneity of the dispersion have been identified from simple geometric arguments.

5.2 HOLD-UP DATA

The following is a summary of the tests carried out to measure gas hold-up, along with the location of the raw data :

Tank diameter (m)	Impeller	Liquid (raw data)	Table Number
0.61	T/3 Rushton	Water	5.2
		Octanol	5.3
		Electrolyte	5.4
	0.47T Rushton	Electrolyte	5.5
	T/3 Concave	Octanol	5.6
		Electrolyte	5.7
	T/2 4 PBT (down)	Octanol	5.8
	T/2 6 PBT (down)	Electrolyte	5.9
	T/2 6 PBT (up)	Electrolyte	5.10
1.83	T/3 Rushton	Water	5.11
		Octanol	5.12
	T/3 Concave	Octanol	5.13
2.67	T/3 Rushton	Water	5.14
		Octanol	5.15
	T/3 Concave	Water	5.16
		Octanol	5.17
	T/2 4 PBT (down)	Octanol	5.18

5.3 CORRELATION TECHNIQUES

If one were to take the conventional approach to dimensionless analysis of the overall gas hold-up (disregarding for the moment correlating techniques that are already established), the list of independent variables would be :

- Impeller type
- Impeller diameter
- Tank diameter
- Gassing rate
- Impeller speed
- Liquid type
- Impeller bottom clearance
- Baffle geometry
- Liquid height

This list is probably not exhaustive, but illustrates the problem facing both the experimenter and the industrial designer. In this series of tests, geometric similarity between vessels has been maintained and both coalescing and non-coalescing liquids used to reduce the number of independent variables as much as possible.

A common approach to the correlation of hold-up is :

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (5.7)$$

$\bar{\epsilon}_m$ = energy dissipated by impeller per unit mass [W/kg]

v_{SG} = superficial gas velocity [m/s]

The current data include a wide range of scales, and the Rushton turbine data using octanol from the 0.61 m and 2.67 m tanks were selected to make an initial evaluation of this approach for scale-up purposes. To test the validity of using superficial gas velocity to represent gassing rate, the following alternative correlating equations were tested :-

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta (VVM)^\gamma \quad (5.8)$$

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta Q^\gamma \quad (5.9)$$

VVM = Gassing rate in vessel volumes per minute

Q = Gassing rate [m^3/s]

The results of this exercise are shown in Table 5.19. The correlation coefficient obtained using v_{SG} was far superior to the alternative approaches. This was anticipated from the correlations of other workers, but the result is a useful confirmation. An explanation of the use of correlation coefficients and the statistical analysis techniques used to correlate the data can be found in Appendix 5B.

As a refinement of this correlating approach, Miller (1974) and Davies (1986), amongst others, have included the expansion power of the gas :

$$\epsilon_G = \alpha (\bar{\epsilon}_m + C \bar{\epsilon}_g)^\beta v_{SG}^\gamma \quad (5.10)$$

$\bar{\epsilon}_g$ = Gas isothermal expansion power [W/kg]

C = Coefficient (dimensionless)

Miller (1974) produced a rather complex set of arguments to show that the gas expansion power and the power dissipated by the agitator were not simply additive, whilst Davies (1986) and Rushton and Bimbinet (1968), for example, took C as unity. The Rushton turbine data for octanol in the 1.83 m diameter tank were used to test the effect of adding various proportions of the gas expansion power to the power dissipated by the agitator (Table 5.20). The effect on the correlation coefficient was minimal, as was the effect on the maximum error between measured and predicted values. Even for the extremely unrealistic case of $C = 2$ (adding twice the gas expansion power to the impeller power), virtually no change in correlation coefficient or maximum error was obtained. Statistically, there are no grounds for including the expansion power for data from a single scale.

The combined data set for the Rushton turbine at all three scales using octanol was also checked using equation (5.10).

Again, no improvement in correlation coefficient was obtained by adding the isothermal expansion power of the gas to the power dissipated by the agitator.

One feature which emerges from the correlations is that the sum $(\beta + \gamma)$ is independent of the value of C . This means that for the case of $\bar{\epsilon}_m = 0$, when the vessel operates as a bubble column, the equation reduces to a correlation giving an exponent on superficial gas velocity which is independent of C .

For $\bar{\epsilon}_m = 0$, equation (5.10) reduces to :

$$\epsilon_G = \alpha (C \bar{\epsilon}_g)^\beta v_{SG}^\gamma \quad (5.11)$$

The isothermal expansion power of the gas is given by :

$$\bar{\epsilon}_g = \frac{\rho_g Q}{\rho_L V_L} \frac{RT}{M} a \ln \left(\frac{p_i}{p_o} \right) \quad (5.12)$$

At constant scale and geometry, equation (5.12) reduces to (for the 1.83m vessel) :

$$\bar{\epsilon}_g = 7.35 v_{SG} \quad (5.13)$$

Hence., from equation (5.11) :

$$\epsilon_G = \alpha (7.35C)^\beta v_{SG}^{\beta + \gamma} \quad (5.14)$$

Note that if equation (5.14) is to be independent of C (which it should be since the equation is now to correlate a bubble column), the product $\alpha (7.35C)^\beta$ should be constant, which from Table 5.20, it clearly is not. This is a rather unsatisfactory situation and underlines the point that the theoretical basis of the power input, superficial gas velocity correlation approach is rather unclear, although the equation does appear to fit the data.

At least the trend shown by the equation for zero impeller power input is correct in the dependence of hold-up on superficial gas velocity. The exponent on v_{SG} in equation (5.14) is independent of the value of C and shows a higher value than the stirred vessel correlation. Several authors who have studied hold-up in bubble columns, such as Hughmark (1967), Akita and Yoshida (1973), Mersmann (1978) and Botton and Cosserat (1978), have found the exponent on superficial gas velocity to be approximately 1.0, which is rather higher than the range of exponents found by workers using stirred vessels (see Literature Survey).

Unfortunately, it is not possible to verify equation (5.14) experimentally because of the problems involved in measuring hold-up for a bubble column with gas confined almost to the vessel axis, as is the case for an unagitated mixing vessel. The liquid surface shows pronounced changes in height with radius, which prevents the use of level change and spill-over methods to measure hold-up. Furthermore, it is rather unsound to attempt to apply literature correlations for bubble column hold-up to the current situation. It might be argued that equation (5.14) should reduce to the same equation as might be found in the bubble column literature, but which literature correlation?. The effects of sparger design are important for bubble columns in the flow regimes of interest here. Should the reference situation of zero impeller power be taken as a stationary agitator, or as the agitator being completely absent?. For a sparger beneath the agitator, the bubble sizes produced could be very different for the two cases, especially for a disc turbine.

Whilst adding the gas expansion power to the impeller power input has conceptual advantages over the use of impeller power input only, there are considerable problems in the interpretation of the result particularly when one considers the case of zero impeller power input.

5.4 SCALE-UP

5.4.1 Correlation using power input and superficial gas velocity

Table 5.21 to 5.23 shows the results of correlating the data using:

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (5.7)$$

The complete sets of data (discounting any effects due to flow regime changes, which are discussed in a later section), all show that power input and superficial gas velocity can be used to correlate hold-up over the complete range of scales used in this project. This was true for :

- (a) Rushton turbine, T/3, octanol solution (Table 5.21 and Fig 5.7).
- (b) Rushton turbine, T/3, water (Table 5.21 and Fig 5.8).
- (c) Concave turbine, T/3, octanol solution (Table 5.22 and Fig 5.9).
- (d) 4 blade, pitched blade turbine, T/2, octanol solution (Table 5.23 and Fig 5.10).

The values of the exponents α , β and γ varied with scale in a non-linear fashion for all four cases, although combining the three data sets did not result in a significant fall in the correlation coefficient. Furthermore, for all cases, the separate correlations for each scale were statistically significantly different from each other.

Table 5.24 shows the range of the variables $\bar{\epsilon}_m$ and v_{SG} covered in each vessel. Ideally, the range of operating conditions in each tank would have been the same, but there were limitations of equipment, generally of the mixer drives, which placed considerable restrictions on the achievable performance. This was especially true of the 0.61 m vessel, which was limited in

speed for various mechanical reasons, and of the 2.67 m tank, where the speed was limited by the natural frequency of the mixer. This in turn placed restrictions on the maximum gas flow rate, since experiments were not carried out on flooded agitators.

To test the effect of the range of operating conditions on the correlations, a set of data was formed by combining the Rushton turbine octanol data from all three vessels, and then eliminating all data points for which $\bar{\epsilon}_m > 0.8 \text{ W/kg}$. The coefficients for the overall correlation were not significantly changed compared to the original data set, and the correlation for the 1.83 m data with $\bar{\epsilon}_m < 0.8 \text{ W/kg}$ was similar to the correlation for the data set as a whole. This implies that the range of independent variables covered in the correlations was not significant for this data, and that the apparent scale effects cannot be explained in terms of changes in the range of operating conditions.

The results of Davies (1986) and Warmoeskerken (1986) suggest that separate correlations should be made for the clinging and "3 - 3" cavity regimes. Applying Warmoeskerken's correlation to the data used to produce the correlations shown in Table 5.21 for the Rushton turbine predicts that only 4 data points were in the clinging cavity regime (out of 92 points in total). It was assumed that the octanol solutions changed from clinging cavities to "3 - 3" cavities at the point predicted by the correlation. This was in line with Warmoeskerken's own data for an electrolyte solution compared to water, for which the correlation was developed. If there is an error arising from this assumption, it is to overestimate the number of clinging cavity data points, which supports the conclusion that the apparent scale effects on the regression coefficients are not caused by changes in flow regime.

The arguments developed to explain the differences in apparent hold-up between the level-change and spill-over methods may be

of value in considering scale effects. If the gas is not distributed over the entire vessel, then the overall hold-up is given by (for constant dispersion volume):

$$\epsilon_G = \frac{V_1}{V_1 + V_2} \epsilon_g \quad (5.3)$$

V_1 = volume of two-phase region

V_2 = volume free of gas

Although there is no quantitative evidence from the literature, it might reasonably be assumed that the volume free of gas is related to the minimum speed for complete dispersion, N_{CD} , which is predicted by the correlation of Nienow et al (1978) :

$$Fr_{CD}^{0.25} \left(\frac{D}{T}\right)^{0.25} = 2.25 Fl_{CD}^{0.5} \quad (5.16)$$

This assumption at least has the merit of predicting similar gas distributions at different scales for $N = N_{CD}$.

Re-arranging this equation gives an expression for N_{CD} as a function of gassing rate :

$$N_{CD} = \frac{3.98Q^{0.5}T^{0.25}}{D^2} \quad (5.17)$$

Since,

$$v_{SG} = \frac{4Q}{\pi T^2} \quad (5.18)$$

$$N_{CD} = 3.53 \left(\frac{T}{D}\right)^{1.25} \frac{v_{SG}^{0.5}}{D^{0.75}} \quad (5.19)$$

Now consider geometrically similar scale-up at constant power input per unit volume and superficial gas velocity. Gassed power input by an agitator is extremely difficult to predict, so the case of constant ungassed power per unit volume will be taken (which is the same as assuming that P_g/P_u is constant):

$$\frac{P}{V} = \frac{\rho P_o N^3 D^5}{V} \quad (5.20)$$

$$V \propto D^3$$

Therefore a relationship for impeller speed as a function of scale at constant D/T may be obtained:

$$N = kD^{-2/3} \quad (5.21)$$

k = constant

By combining equations (5.19) and (5.21) :

$$\frac{N}{N_{CD}} = 0.284 k \left(\frac{D}{T}\right)^{1.25} \frac{D^{0.06}}{v_{SG}^{0.5}} \quad (5.22)$$

The scale effect on power number has been accounted for in arriving at equation (5.22). The implication of equation (5.22) is that $\frac{N}{N_{CD}}$ increases slightly with increasing scale, giving

better dispersion (and therefore higher hold-up) at larger scale.

The trend predicted from the argument so far is the opposite of that shown by the experimental data. The correlation is in terms of gassed power input; for scale-up at constant gassed power input, the following expression holds:

$$N_1^3 D_1^2 \left(\frac{P_g}{P_u}\right)_1 = N_2^3 D_2^2 \left(\frac{P_g}{P_u}\right)_2 \quad (5.23)$$

1 = small scale

2 = large scale

If $N_1^3 D_1^2 = \text{constant} = x$, then for the larger scale:

$$N = D^{-2/3} x^{1/3} \left[\frac{(P_g/P_u)_1}{(P_g/P_u)_2} \right]^{1/3} \quad (5.24)$$

Combining this with equation (5.19) gives a modified equation for N/N_{CD} :

$$\frac{N}{N_{CD}} = X' D^{0.06} \left[\left(\frac{P_g/P_u}{P_g/P_u} \right)_1 \right]^{1/3} \quad (5.25)$$

$X' = \text{constant}$, independent of scale

It now only remains to estimate the effect of scale on the gassed power ratio P_g/P_u . Since superficial gas velocity and (approximately) $N^3 D^2$ are constant, an expression can be found for the flow number as a function of scale:

$$Fl = \frac{Q}{ND^3} \quad (5.26)$$

But $Q = \frac{\pi T^2}{4} v_{SG}$ and D/T is constant, so:

$$Fl = \frac{k' v_{SG}}{ND} \quad (5.27)$$

$k' = \text{constant}$

But $N \propto D^{-2/3}$ (approximately), hence:

$$Fl = \frac{k'' v_{SG}}{D^{1/3}} \quad (5.28)$$

$k'' = \text{constant}$

The flow number decreases somewhat with increasing scale for constant superficial gas velocity. Using Fig 4.11, the increase in P_g/P_u with scale can be estimated as approximately 25% in scaling up from $T = 0.61m$ to $T = 2.67m$. This tends to remove the scale effect on N/N_{CD} .

The overall conclusion from this exercise is that keeping gassed power input per unit mass and superficial gas velocity constant on scale-up is equivalent to keeping N/N_{CD} constant. If one accepts that the distribution of gas in the vessel is important in determining the overall hold-up, and that N/N_{CD} is a

parameter which could be held to characterise the gas distribution in the tank, this result may be taken as a partial justification for the success of the power input/superficial gas velocity correlation method over such a wide range of scales.

It is useful to consider individual data points taken at approximately the same values of $\bar{\epsilon}_m$ and v_{SG} in the three vessels. This could be done most conveniently for the tap water data with the Rushton turbine (Fig 5.11). A limited number of data points could be considered in this way as a result of the different operational constraints in each vessel. However, there is a clear trend for the larger scale (1.83 m and 2.67 m) vessels to give lower values of hold-up than the equivalent experiment in the 0.61 m tank. This trend is also shown by plots of the predictions of the correlations at constant superficial gas velocity (Fig 5.12).

The results for the concave turbine in octanol (Fig 5.13), show a similar trend to the Rushton turbine tap water data, whilst the pattern for the Rushton turbine is difficult to discern from individual data points (Fig 5.14), but the general trend predicted by the correlations is for hold-up to decrease slightly with scale.

The pitched blade turbine with 4 blades, pumping downwards scaled in the same manner as the disc turbines, with the data adequately described by a single correlation, but with differences in hold-up appearing between scales at the same operating conditions. Once again, there was a general trend for hold-up to be lower at large scale (Fig 5.15) than at small scale.

To test whether the scale effect on hold-up could be quantified in a systematic way, the following correlation was attempted :

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^B (v_{SG})^\gamma D^\delta \quad (5.29)$$

This is a simplified version of the ideas of Chandrasekharan and Calderbank (1981) developed to produce scale-up laws for $k_L a$ (see Literature Survey). The addition of the term for scale produced virtually no increase in correlation coefficient. The fit to the Rushton turbine octanol data was not significantly better than the original correlating equation. The same conclusion was drawn from correlating the tap water data in a similar way.

Overall, although a trend for hold-up to decrease slightly with scale has been observed, this cannot be adequately quantified and there are no grounds to account for scale explicitly in the form of equation (5.29). The implications of this result for scale-up will be considered in a later section. It should be remembered that maintaining constant superficial gas velocity on scale-up is not standard industrial practice. In practical terms, the degree of reduction of hold-up with scale observed in this work is probably insignificant.

5.4.2 Correlations based on speed and gas flow rate

Table 5.25 and 5.26 show the results of correlating gas hold-up using:

$$\epsilon_G = \alpha N^{\beta} Q^{\gamma} \quad (5.30)$$

The motivation for using this equation is that it is potentially of more practical value to the designer than the power input, superficial gas velocity method, since it avoids the prediction of gassed power, which is likely to be highly unreliable.

Correlation using equation (5.30) was markedly less successful than the superficial gas velocity, power input approach in accounting for the effects of scale, but for individual sets of data (Table 5.25 and Table 5.26), it was at least as successful as power input/superficial gas velocity. Warmoeskerken (1986) used an additional term to account for scale, giving four

regression coefficients. This was applied to the current data for the Rushton turbine at all 3 scales, giving the following equations :

$$\text{OCTANOL : } \epsilon_G = 0.21N^{0.64} Q^{0.53} D^{-0.53} \quad (5.31)$$

(CORRELATION COEFFICIENT = 0.946)

$$\text{WATER : } \epsilon_G = 0.14N^{0.53} Q^{0.45} D^{-0.50} \quad (5.32)$$

(CORRELATION COEFFICIENT = 0.958)

The exponent on scale reported by Warmoeskerken (1986) was +0.1, very different from the current result.

A check for consistency with the alternative correlation method can be made by assuming that power input per unit volume is constant. Hence, for constant D/T :

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma$$

$$\bar{\epsilon}_m \propto N^3 D^2, v_{SG} = \frac{4Q}{\pi T} \quad (5.33)$$

$$\therefore \epsilon_G = a N^{3\beta} Q^\gamma D^{2(\beta-\gamma)} \quad (5.34)$$

Expressing the original equation for ϵ_G in the following terms :

$$\epsilon_G = \alpha' N^{\beta'} Q^{\gamma'} D^{\delta'} \quad (5.35)$$

It follows from (5.34) and (5.35) that :

$$\beta' = 3\beta \quad (5.36)$$

$$\gamma' = \gamma \quad (5.37)$$

$$\delta' = 2(\beta - \gamma) \quad (5.38)$$

Inspection of the various exponents obtained from the combined correlations shows that the relationships between β' and β and γ' and γ are indeed approximately true. The exponent δ' on D is interesting because as long as $\gamma > \beta$, which is the case for virtually all the correlations published in the literature, δ' is negative. Inspection of the correlations in Table 5.21 in fact shows that the expression derived for δ' is relatively inaccurate, for instance the octanol results predict a value of -0.84 for δ' . This is probably a result of the invalidity of the assumption that $N^3 D^2$ is constant. However, the overall trend is consistent with the predictions made from a comparison of the two correlation techniques.

5.5 COMPARISON OF AGITATOR TYPES

The most extensive data available in this work for agitator comparison is the set of octanol results for the concave and Rushton turbines. The data were compared on a power input basis by using the analysis of variance (F-test) on the correlations using power input and superficial gas velocity. This was done for each vessel and for the data sets as a whole (Table 5.27). The differences between the best fit to the data for one impeller and the fit using the regression coefficients for the other impeller were not statistically significant. As an example, Fig 5.16 is a plot of the combined Rushton and concave turbine data from the 0.61 m tank.

Energy efficiency is important from the point of view of running costs, but the finding that the Rushton and concave turbines produce much the same gas hold-up for a given energy input and gassing rate would suggest that there is no gain in operating efficiency to be made by using a concave turbine instead of a Rushton turbine. This conclusion is in line with Chapman et al (1983) and Figueredo and Calderbank (1978), who found little difference in hold-up between different agitator types operating at the same power input. Workers studying mass transfer, such as Smith, van't Riet and Middleton (1977) have reached the same

conclusion. However, advantages might still be gained by retrofitting existing vessels with concave turbines to replace Rushton turbines. Since hold-up is a function of power input only (at constant superficial gas velocity) and the concave turbine has a flatter power curve than the Rushton turbine and might be expected to draw more power than a Rushton turbine of identical size, there could be gains in hold-up arising from the use of the concave turbine. The success of this approach depends on the absolute values of gassed power number at the operating conditions. In fact (Figs 5.17 and 5.18) in this set of tests the Rushton and concave turbines draw similar amounts of power at the same speed and gassing rates. This is illustrated in Figs 5.19 and 5.20 where results taken at identical speeds and gassing rates for the two agitators have been plotted against each other. With the exception of a few data points in the 0.61 m tank, the data lie within 10% of equal hold-up for the two agitators.

For the case of a mixing vessel sized to operate in ungassed conditions a more realistic approach to retro-fitting would be to use a concave turbine of larger diameter or larger blade width than the original Rushton turbine. This would draw the same ungassed power as the original impeller (the motor speed is generally fixed). The gassed power would then be considerably higher, since the concave turbine produces higher values of P_g/P_u at given flow number. At best, however, one could only double the gassed power demand by these measures.

Since $\epsilon_G \propto \bar{\epsilon}_m^{0.3}$ approximately, the amount of hold-up would be increased by only 20%. In critical applications, this may be worthwhile, but it would appear that the rewards of retro-fitting would be marginal in many cases.

Chapman et al (1983) included pitched blade and axial pumping agitators in a series of hold-up measurements, and found them to be correlated by the same equation (on a power input basis) as disc turbines. The results of combining the Rushton turbine and 4 blade PBT data in the 0.61 m and 2.67 m tanks are given in

Table 5.28. The correlation coefficients were very low indicating that differences in dispersion efficiency exist between pitched blade turbines and disc turbines. A plot of the data for the 0.61 m tank is given in Fig 5.21. At constant superficial gas velocity, the pitched blade turbine appeared to give higher hold-ups at low power inputs. Plotting the data at constant gas superficial velocity for each agitator in the 2.67 m vessel shows this trend (Fig 5.22). A similar plot for the 0.61 m tank is less revealing (Fig 5.23), the range of power inputs for the pitched blade turbine was somewhat restricted, and there were some experimental problems involved in measuring hold-up using the spill-over method with the pitched blade turbine in the small vessel caused by the extremely disturbed nature of the surface.

The data for the 2.67 m vessel indicate that the hold-up at constant power input and superficial gas velocity is of the same order for the two agitators, at least for the superficial gas velocity shown in Fig 5.22. The correlation coefficient for the combined 2.67 m data is probably low because of increased scatter about the correlation line as one attempts to account for agitators which behave in a fundamentally different way with a single correlation. The predictions of the correlations for the two agitator types at 0.61 m and 2.67 m scales combined are shown in Fig 5.24. The correlations suggest that as gas superficial velocity increases, the pitched blade turbine becomes energetically less favourable than the Rushton turbine. Inspection of individual data points for the 2.67 m vessel tentatively confirms this trend.

5.6 EFFECT OF PUMPING DIRECTION

A set of experiments was carried out using the 6PBT in electrolyte solution in both upwards and downwards-pumping modes. The raw data are given in Tables 5.9 and 5.10. The correlations based on specific power input and superficial gas velocity are given in Table 5.23. The most important feature of

the correlations is the drop in correlation coefficient for the combined correlation compared to the correlation for each pumping mode. A plot of the combined data shows a good deal of scatter, especially in the downwards-pumping data, about the correlation line. Once again, closer examination of the data is required to investigate the cause of this scatter.

Fig 5.25 shows data points plotted as lines of constant superficial gas velocity as functions of impeller specific power input. A clear trend is apparent, with the downwards-pumping agitator giving higher hold-up at constant power input than upwards-pumping at low superficial gas velocity, with the reverse being true at high superficial gas velocity. A parallel can be drawn with the correlation predictions discussed in the previous section, where the pitched blade turbine pumping downwards tended to produce less hold-up per unit energy input than the Rushton turbine as superficial gas velocity increased,

The differences between the upwards and downwards pumping agitators may tentatively be ascribed to the flow regime of the downwards-pumping agitator.

At low superficial gas velocities the downwards-pumping impeller is indirectly loaded at all impeller speeds, whilst at high superficial gas velocities the downwards pumping agitator is directly loaded. For indirect loading, bubbles are swept around strong liquid circulation loops and some bubbles may not pass through the agitator at all. This is a high efficiency mode of operation, from the experimental data, for reasons which are not entirely apparent. Gas is fairly well distributed over the whole vessel, and the bubbles are reasonably uniform (from visual observation). For direct loading, the gassed power demand becomes relatively independent of gassing rate, and there is a tendency for large pockets of gas to form which rise quickly to the surface, contributing little to the gas hold-up. The dispersion efficiency will tend to fall off with increasing gassing rate because, whilst the power input remains

constant, an increasingly large fraction of the gas throughput is not adequately dispersed by the agitator. Agitators which disperse the gas without flooding are clearly unaffected by this phenomenon, and become energetically more favourable than the downwards-pumping PBT at high gassing rates since they continue to disperse all the gas throughput as fine bubbles.

A total of 5 agitators were tested using electrolyte solution in the 0.61 m diameter tank. A plot of hold-up against power input at constant superficial gas velocity is given in Fig 5.26.

Whilst the impellers give very similar hold-up at low power input, the high power input data show considerable differences between the upwards-pumping PBT and the Rushton turbine of similar diameter. The difference between the two correlations was statistically significant. The repeatability of the electrolyte hold-up data is discussed in detail in Appendix 5A. Briefly, the PBT data are thought to be reliable, whilst the Rushton turbine data in Fig 5.26 are thought to underestimate the hold-up. Experimental error is not thought to be responsible for the differences in energy efficiency between the two agitator types. The data clearly show that the upwards-pumping PBT produces less hold-up than a disc turbine at the same power input.

5.7 EFFECT OF IMPELLER SIZE

The electrolyte data for the Rushton turbine with $D = T/3$ and $D = 0.47T$ allow the assessment of impeller size effects. The two sets of data were combined and correlated with reasonable success using power input and gas superficial velocity. There is therefore, no gain in energy efficiency to be obtained by increasing (D/T) as has been shown to be the case for other mixing processes, such as miscible liquid blending (Khang and Levenspiel, 1976) and solids suspension (Nienow 1985).

The data also allow the distinction to be made between the influence of power input and impeller tip speed on gas

dispersion. A correlation was attempted using :

$$\epsilon_G = \alpha U_s^B Q^Y \quad (5.39)$$

The correlation coefficient was 0.786, much lower than the value obtained using power input. This is perhaps not surprising when one considers the scale-up data, which are also correlated by power input per unit volume, but it is useful to confirm expected trends so clearly.

An attempt was also made to correlate the data using impeller speed (Table 5.24). This was very unsuccessful, which emphasises the need to account for impeller size effects when using this correlation technique. By including a term for impeller diameter, the following correlation was obtained :

$$\epsilon_G = 3.32 N^{1.0} Q^{0.48} D^{1.65} \quad (5.40)$$

The correlation coefficient was 0.928, still rather low when one considers that there are now four regression coefficients, and the fit to the data is poorer than the power input correlation which only uses three regression coefficients. The main point to note is that the scale term is very different from that obtained by varying D at constant D/T . The inevitable conclusion is that hold-up is not correlated by an expression such as (5.40) in a manner which accounts for the effects of both scale and (D/T) . It is also quite possible that the exponents on speed and flow rate themselves depend on (D/T) , further complicating the situation.

Clearly, in order to produce a comprehensive correlation for hold-up without using power input per unit volume as a correlating parameter, a set of experiments should be conducted at several vessel sizes using several values of (D/T) . This would be extremely time consuming, since it would also be necessary to investigate other geometric parameters such as

bottom clearance and degree of baffling to check their influence on hold-up.

In discussing scale-up effects, the ratio N/N_{CD} was shown to be approximately constant when scaling up using constant $\bar{\epsilon}_m$ and v_{SG} . This was taken as being indicative of a fundamental mechanism behind the success of this correlation approach when scaling up. A similar exercise for the case of varying D/T produces a different result :

$$\epsilon_G \propto (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (5.41)$$

From equation (5.17) :

$$N_{CD} = \frac{3.98Q^{0.5}T^{0.25}}{D^2} \quad (5.42)$$

Assuming that $N^3 D^5 = \text{constant} = X$, for constant power input,

$$\frac{N}{N_{CD}} = \frac{X^{1/3} D^{1/3}}{3.98Q^{0.5}T^{0.25}} \quad (5.43)$$

If Q is constant, N/N_{CD} increases considerably with increasing D , whilst a plot of the data at constant superficial gas velocity (Fig 5.26) shows that, if anything, the T/3 agitator tends to give higher hold-up at given power input. The assumption that, at constant power input, hold-up increases as a result of improved gas distribution at higher values of N_{CD} predicts the opposite trend. There are clearly influences on gas hold-up other than the influence of gas distribution.

5.8 COMPARISON OF LIQUID TYPES

For all vessels except the 0.61m tank at low power inputs, the octanol solutions produced higher hold-up than tap water at the same operating conditions. Over all three scales the correlations for the Rushton turbine were:

$$\text{OCTANOL} \quad \epsilon_G = 1.36 (\bar{\epsilon}_M)^{0.22} v_{SG}^{0.61} \quad (5.44)$$

$$\text{WATER} \quad \epsilon_G = 0.63 (\bar{\epsilon}_M)^{0.17} v_{SG}^{0.51} \quad (5.45)$$

The two liquid types were broadly similar in their sensitivity to gassed power input and superficial gas velocity, with the coalescing liquid being slightly less dependent on the two parameters than the coalescing liquid.

At first sight, the difference between the liquid types is simply a constant. The ratio of hold-up in non-coalescing liquid to that in coalescing liquid is virtually independent of power input and superficial gas velocity. However, if the hold-up ratio is plotted against gas flow rate (for the 1.83m tank), the ratio of hold-up falls quite sharply with increasing gassing rate (Fig 5.27). The data points were taken at identical impeller speed for the two liquid types. There is little or no evidence in the literature to provide a detailed explanation of this phenomenon. The fall in power demand with increasing gassing rate is similar for the two liquid types at constant impeller speed giving approximately the same conditions of liquid turbulence. It is possible that re-circulation effects for the non-coalescing medium become weaker as gas flow rate increases. The coalescing medium was not operating in a regime of gross re-circulation, so the gradual disappearance of re-circulation effects in the non-coalescing medium would account for the steady fall in hold-up ratio with increasing gassing rate seen in Fig 5.27.

The data of Hassan and Robinson (1977) show a very similar trend, as do the data of Davies (1986). Some systematic effect is clearly taking place, but no clear explanation appears to be currently available.

NOMENCLATURE : Chapter 6

<u>SYMBOL</u>	<u>MEANING</u>	<u>UNITS</u>
A_B	Interfacial area of bubble	m^2
C_{H_2O}	Concentration of water vapour in gas phase	$kmol/m^3$
C_G	Concentration of key component in gas	$kmol/m^3$
C_L	Concentration of key component in liquid	$kmol/m^3$
$\bar{C}_L$	Dimensionless liquid phase concentration	-
C_{LM}	Measured liquid concentration	$kmol/m^3$
C_{LX}	Concentration of component X in liquid	$kmol/m^3$
C_{LY}	Concentration of component Y in liquid	$kmol/m^3$
C_{NG}	Concentration of nitrogen in gas	$kmol/m^3$
C_{NL}	Concentration of nitrogen in liquid	$kmol/m^3$
C_{OG}	Concentration of oxygen in gas	$kmol/m^3$
C_{OL}	Concentration of oxygen in liquid	$kmol/m^3$
$\bar{C}_{OL}$	Dimensionless liquid oxygen concentration	-
$C_{O\infty}$	Concentration in gas phase at equilibrium at outlet from tank	$kmol/m^3$
$\bar{C}_O$	Dimensionless outlet concentration	-
C_{SAT}	Concentration of oxygen in liquid phase at equilibrium	$kmol/m^3$
C_X	Concentration of component X in gas phase	$kmol/m^3$
C_Y	Concentration of component Y in gas phase	$kmol/m^3$
d_B	Bubble diameter	m
D_L	Diffusion coefficient of gas in liquid	m^2/s
K_O	Equilibrium constant for oxygen	-
K_N	Equilibrium constant for nitrogen	-
k_{LX}	Liquid side mass transfer coefficient - component X	m/s
k_{LY}	Liquid side mass transfer coefficient - component Y	m/s
m	Number of elements in tank	-
p	Hydrostatic pressure	N/m^2
Q	Gas flow rate	m^3/s
t	Time	s

V_B	Volume of bubble	m^3
v_b	Bubble rise velocity	m/s
τ_C	Circulation time of liquid	s
τ_L	First order response time of liquid probe	s
θ	Temperature	$^{\circ}C$
ρ	Gas density	$kmol/m^3$

Subscripts

i	=	inlet to vessel
o	=	outlet from vessel
∞	=	at equilibrium

6. MASS TRANSFER

Few subjects have caused more controversy and uncertainty than the measurement of the mass transfer coefficient, $k_L a$, in stirred vessels. This chapter examines the existing theory behind the analysis techniques used in dynamic response experiments in an attempt to gain physical insight into the mass transfer process, particularly with regard to scale effects. A set of dynamic response experiments in the 0.61m and 2.67m vessels is also described along with the experimental data.

6.1 THEORY

It is instructive to consider the simplest possible situation in which gas liquid mass transfer takes place; the case of a single bubble rising through a liquid. To draw an analogy with the case of dynamic response experiments, it will be assumed that the bubble consists of a mixture of gases X and Y. The liquid is saturated with Y, but has zero concentration of X. The following processes take place, each of which have some influence on the rate of mass transfer:

- 1 Expansion of the bubble as it rises through the liquid. This increases the interfacial area for mass transfer but decreases the gas-phase concentrations of X and Y.
- 2 Mass transfer of X from gas to liquid and of Y from liquid to gas. This also influences the partial pressures of X and Y and the bubble interfacial area.

A mass balance on components A and B produces:

$$\frac{d}{dt}(V_B C_X) = - k_{LX} A_B (K_X C_X - C_{LX}) \quad (6.1)$$

$$\frac{d}{dt}(V_B C_Y) = - k_{LY} A_B (K_Y C_Y - C_{LY}) \quad (6.2)$$

Assuming that the bubble is composed of a mixture of perfect gases (which is realistic for air):

$$pV_B = nRT \quad (6.3)$$

$$\frac{d}{dt} \left(\frac{p}{RT} \right) = \frac{d}{dt} \left(\frac{n}{V_B} \right) \quad (6.4)$$

Since the process is assumed to be isothermal, an expression describing the volume change of the bubble can be obtained:

$$\frac{p}{RT} \frac{dV_B}{dt} = \frac{dn}{dt} - \frac{V_B}{RT} \frac{dp}{dt} \quad (6.5)$$

But p is just the local hydrostatic pressure, given by (using the co-ordinates of Fig 6.1)

$$p = p_O + \rho_L g (H - x) \quad (6.6)$$

Also, if the bubble is spherical:

$$V_B = \frac{\pi d_B^3}{6} \quad (6.7)$$

$$A_B = \frac{\pi d_B^2}{4} \quad (6.8)$$

The gas bubble reaches its terminal rise velocity almost immediately after formation, so that its motion is described by:

$$v_b = \frac{dx}{dt} \quad (6.9)$$

Finally, $\frac{dn}{dt}$ is the net rate of mass transfer from gas to liquid. By combining equations (6.1), (6.2) and (6.5) to (6.9), the following relations may be obtained for the unknown quantities C_X , C_Y and d_B :

$$d_B \frac{dC_X}{dx} + 3C_X \frac{d}{dx} (d_B) = - \frac{3k_{LX}}{2v_b} (K_X C_X - C_{LX}) \quad (6.10)$$

$$d_B \frac{dC_Y}{dx} + 3C_Y \frac{d}{dx} (d_B) = - \frac{3k_{LY}}{2v_b} (K_Y C_Y - C_{LY}) \quad (6.11)$$

$$\begin{aligned} 3v_b \frac{d}{dx} (d_B) &= \frac{RT}{p_O - \rho_L g(H-x)} \left[\frac{3k_{LX}}{2} (C_{LX} - K_X C_X) + \right. \\ &\quad \left. \frac{3k_{LY}}{2} (C_{LY} - K_Y C_Y) + \right. \\ &\quad \left. d_B (C_X + C_Y + C_{H_2O}) \rho_L g v_b \right] \\ &\quad \frac{1}{p_O + \rho_L g(H-x)} \end{aligned} \quad (6.12)$$

It is assumed that the bubble is saturated with water vapour, which then acts as an inert gas. This effect is only likely to be significant at high vapour pressures, and will not be discussed further.

An expression relating bubble rise velocity to diameter is all that is required to solve the equations (6.10) to (6.12). For bubbles of the size found in coalescing liquids in stirred tanks, the rise velocity may be taken as constant, an assumption already implicitly made in equation (6.9).

From this (comparatively) straightforward set of equations, the effects of depth and presence of other components on the overall transfer rate of key component X, may be predicted.

The most convenient way of expressing the mass transfer rate is as a mean value over the whole liquid depth:

$$\bar{R} = \frac{1}{H} \int_0^H k_{LX} \frac{\pi d_B^2}{4} (K_X C_X - C_{LX}) dx \quad (6.13)$$

Furthermore, by conceptually varying the value of rise velocity, the influence of residence time on the mean transfer rate may be predicted. This is to simulate (partially) the situation in a stirred vessel, when the bubble residence time will generally be considerably higher than for the case of free rise through a stagnant medium.

The initial conditions are given by an assumed initial bubble size and calculated values of C_X and C_Y for given gas composition at the inlet pressure. The liquid phase concentrations are assumed not to vary as a result of mass transfer, the liquid is taken as infinite in extent. The gases have been taken as oxygen and nitrogen with

$$k_{LX} \approx k_{LY} \approx 4 \times 10^{-4} \text{ m/s (van't Riet, 1979).}$$

The equations were solved using Runge-Kutta integration on a PDP 11/23 laboratory computer since no simple analytical solution appeared to be available.

The results of solving the equations for the full model including nitrogen transfer and depth effects are given in Table 6.1(a). The mean mass transfer flux has been included to enable a direct comparison of different bubble sizes. This is defined as:

$$\bar{F} = \frac{1}{H} \int_0^H k_{LX} (K_X C_X - C_{LX}) dx \quad (6.14)$$

The results clearly show the effects of depletion caused by mass transfer and of bubble size on the mean mass transfer rate. The bubble rise velocity has been varied to simulate residence time effects, and the rise velocities used in the simulations are in no way intended to be representative of actual rise velocities. The longer the bubble residence time, the greater the depletion of oxygen in the bubble, and the mean mass transfer rate is reduced accordingly. Similarly, for reducing bubble size, the mass transfer flux falls as a result of increased depletion of oxygen in the bubble.

Scale effects may be seen in Table 6.1(b), for which scale-up (increasing water depth) has been considered at constant bubble rise velocity (which would be the case if one conducted a set of experiments to study these effects) and at constant residence time, which is representative of scale-up techniques used for stirred tanks. The mean mass transfer rate and the mean flux both increase with increasing scale in each case. This is due to the increasing concentration driving force resulting from gas compressibility effects. Also, the bubble size at the inlet has been taken as constant, so the bubble contains more mass at large scale than at small scales. This will tend to reduce depletion effects, particularly for the simulations at constant rise velocity.

The effects of neglecting nitrogen transfer can be seen in Table 6.1(c). For the simulations investigated, the effect on the mean mass transfer rate was negligible, although some effect on the mean flux was visible. The effects of nitrogen transfer are:

- (i) To reduce the partial pressure of oxygen in the bubble by increasing the mole fraction of nitrogen.
- (ii) To increase the area available for mass transfer by expanding the bubble.

This explains the fact that some transfer rates accounting for nitrogen diffusion (such as $d_{BO} = 1\text{mm}$, $v_b = 0.001\text{ m/s}$) were higher than the predicted values neglecting nitrogen transfer. The general trend, however, is for these two effects to cancel out.

The effects of neglecting liquid depth are, however, more significant. Two types of simulation have been carried out:

- (i) Starting concentration of oxygen calculated at inlet hydrostatic pressure. (High inlet concentration).

- (ii) Starting concentration assumed to be equal to concentration at atmospheric pressure. (Low inlet concentration).

The results of these simulations are given in Tables 6.1(d) and 6.1(e), for a range of scales using constant rise velocity and constant gas residence time. The transfer rates using method (i) are reasonably close to those predicted using the full model, but tend to over-estimate the transfer rate, the error becoming larger with increasing scale. However, if the gas concentration is assumed to be that at normal atmospheric pressure, the error in mean transfer rate becomes very significant.

The conclusion from this exercise is that nitrogen transfer effects are probably negligible in mass transfer experiments, but depth effects must be included in any rigorous model of behaviour in the gas phase.

Lakin, Salzman, Oldshue and Gray (1978) presented a similar analysis to account for depth effects in subsurface aeration systems at large scale. The problem has also received considerable attention in the literature on bubble columns, such as the work of Jackson and Chia-Cheuh Shen (1978), but appears to have been neglected in work on stirred tanks, probably because many mass transfer experiments have been carried out in small scale equipment, and because the complex hydrodynamics of stirred vessels make the idealised case of a single bubble rising uniformly through an infinite medium without break-up or coalescence extremely unrealistic. Manfredini, Cavallera, Marim and Donati (1983) considered depth effects on equilibrium concentration of oxygen in a fermenter equipped with multiple agitators using a model in which the vessel was divided into four well-mixed cells, but this work is not applicable to large, single-stirrer vessels.

Diffusion of a second species, such as nitrogen, has been considered by Linek, Benes and Hovorka (1981) for dynamic

response experiments carried out using the perfect mixing assumption for the gas phase. The experimental data showed no influence of counter diffusion of nitrogen for values of $k_L a$ up to 0.11 s^{-1} . This conclusion does not necessarily hold for other gas phase mixing models and will be considered in parallel with the effects of liquid depth for each mixing model in turn.

a) No depletion model

The assumption that the gas inlet concentration is equal to the outlet concentration is widely accepted as unrealistic since it violates the oxygen mass balance. Depth effects compound the problem since even in the absence of mass transfer the constant concentration hypothesis is untrue because of isothermal expansion of the gas. The resultant error is a function of scale, but can be eliminated using the following argument:

- 1 The final equilibrium concentration of oxygen in the liquid phase is given by:

$$C_{L \infty} = K \frac{(C_{Gi} + C_{Go})}{2} \quad (6.15)$$

- 2 The assumption of no depletion of gas phase due to mass transfer can be combined with depth effects to give the following expression for dissolved oxygen concentration:

$$\frac{dC_L}{dt} = k_L a \left(K \frac{(C_{Gi} + C_{Go})}{2} - C_L \right) \quad (6.16)$$

This is reasonable since C_G is a linear function of depth.

- 3 Since the general practice is to use normalised liquid oxygen concentration responses, to avoid calibration of the probe, the familiar expression for the no depletion model results:

$$\frac{d\bar{C}_L}{dt} = k_L a (1 - \bar{C}_L) \quad (6.17)$$

The validity of step 1 of the argument is discussed in a later section.

Note that nitrogen transport has no influence on the no depletion model. This fact is of little significance, since the no depletion model is only a good approximation at low values of $k_L a$.

(b) Perfect mixing model

Linek et al (1981) have derived a set of equations to account for nitrogen transfer for the case of a well-mixed gas phase. The equations are as follows, with a correction for gas compressibility effects included in the overall mass balance:

$$\frac{dC_{OL}}{dt} = k_L a (K_O C_{OG} - C_{OL}) \quad (6.18)$$

$$\frac{dC_{NL}}{dt} = k_L a (K_N C_{NG} - C_{NL}) \quad (6.19)$$

$$\frac{dC_{OG}}{dt} = \frac{k_L a V_L}{V_G} (C_{OL} - K_O C_{OG}) + \frac{Q_i}{V_G} C_{OGi} - \frac{Q_o}{V_G} C_{OG} \quad (6.20)$$

$$\frac{dC_{NG}}{dt} = \frac{k_L a V_L}{V_G} (C_{NL} - K_N C_{NG}) + \frac{Q_i}{V_G} C_{NGi} - \frac{Q_o}{V_G} C_{NG} \quad (6.21)$$

$$\rho_i Q_i = \rho_o Q_o + k_L a V_L (K_O C_{OG} - C_{OL}) + k_L a V_L (K_N C_{NG} - C_{NL}) \quad (6.22)$$

Since $\rho_i = \frac{p_O + \rho_L g H}{RT}$ and $\rho_o = \frac{p_O}{RT}$, equation (6.22) may be

written as:

$$\frac{p_O + \rho_L g H}{RT} Q_i = \frac{p_O}{RT} Q_o + k_L a V_L (K_O C_{OG} - C_{OL}) + k_L a V_L (K_N C_{NG} - C_{NL}) \quad (6.23)$$

The equations for C_{OL} , C_{OG} , C_{NL} , C_{NG} and Q_o may be integrated numerically. A simplification would be to impose the condition

that:

$$C_{NG} + C_{OG} + C_{H_2O} = \text{constant} \quad (6.24)$$

This was used by Linek et al (1981); it becomes necessary to assume that the gas density in the tank is uniformly equal to the gas density at the outlet.

The effects of liquid depth and nitrogen transfer could be evaluated by integrating the differential equations numerically. There is clearly an error caused by neglecting the gas compressibility as previous authors have done in deriving equations for well-mixed gas phase models.

A considerable simplification can be made by assuming that mass transfer has negligible effect on the gas volumetric flow rate, and that the gas phase is in a pseudo-steady state. Equation (6.20) reduces to:

$$0 = k_L a \frac{V_L}{V_G} (C_{OL} - K_O C_{OG}) + \frac{Q_i}{V_G} \left[C_{OGi} - C_{OG} \frac{(p_o + \rho g H)}{p_o} \right] \quad (6.25)$$

Substituting for C_{OG} into equation (6.18) produces an equation of C_{OL} only, after a considerable amount of algebra:

From equation (6.24),

$$C_{OG} = \frac{k_L a V_L C_{OL} + Q_i C_{OGi}}{k_L a V_L K_O + Q_i \frac{(1 + \rho g H)}{p_o}} \quad (6.26)$$

Equation (6.18) becomes,

$$\frac{dC_{OL}}{dt} = \alpha - \beta C_{OL} \quad (6.27)$$

Where $\alpha = \frac{K_O Q_i k_L a C_{OGi}}{k_L a V_L K_O + Q_i (1 + \frac{\rho g H}{P_O})}$ (6.28)

$$\beta = \frac{k_L a Q_i}{K_O k_L a V_L + Q_i (1 + \frac{\rho g H}{P_O})} \quad (6.29)$$

The solution for equation (6.27) assuming zero liquid oxygen concentration at time $t=0$ is just:

$$C_{OL} = \frac{\alpha}{\beta} (1 - e^{-\beta t}) \quad (6.30)$$

But, $\frac{\alpha}{\beta} = \frac{K_O C_{OGi}}{1 + \frac{\rho g H}{P_O}}$ (6.31)

To be consistent with the assumption of perfect back-mixing of the gas phase, it must now be assumed that the liquid is in equilibrium with gas at the outlet condition. In other words,

$$C_{OL\infty} = \frac{\alpha}{\beta} \quad (6.32)$$

Equation (6.30) now becomes;

$$\bar{C}_{OL} = 1 - e^{-\beta t} \quad (6.33)$$

Now consider the case of a hypothetical experiment for which a dissolved oxygen transient response has been obtained. A semi-logarithmic plot of the data can be made provided all the assumptions stated earlier still hold. The effect of depth correction can be evaluated explicitly since if depth effects are neglected ($H=0$), β reduces to:

$$\beta_{H=0} = \frac{(k_L a)_O}{\frac{K_O (k_L a)_O V_L}{Q_i} + 1} \quad (6.34)$$

The slope of the semi-logarithmic plot, as an experimentally determined quantity, is fixed, so the estimate of $k_L a$ obtained

using depth correction can be compared to the estimate of $k_L a$ neglecting depth correction by setting $\beta = \beta/H=0$. It can be shown that the two estimates of $k_L a$, $(k_L a)_H$ and $(k_L a)_O$ are related by:

$$(k_L a)_H = \frac{(k_L a)_O}{\frac{K_O V_L (k_L a)_O}{Q_i} \left[1 - \frac{1}{1 + \frac{\rho g H}{p_O}} \right] + 1} \quad (6.35)$$

The error caused by neglecting depth effects is a function of scale and of $k_L a$. The assumptions behind equation (6.35) restrict its validity to low values of $k_L a$ (this is true for the well-mixed model generally).

The quantity $\frac{V_L}{Q_i}$ is the gas residence time based on liquid

volume, which is generally held constant on scale-up in industrial applications. A value of 1 VVM combined with the maximum reasonable $(k_L a)_O$ value of 0.1 seconds gives:

$$\frac{(k_L a)_H}{(k_L a)_O} = \frac{1}{1 + 0.2 \left(\frac{\rho g H}{p_O + \rho g H} \right)} \quad (6.36)$$

For small scale experiments ($H=0.61\text{m}$) the error in $k_L a$ is approximately 1%. For large scale experiments ($H=2.7\text{m}$), the error in $k_L a$ is approximately 4%. This is within the limits of experimental error and depth effects may be neglected. There is the proviso however, that the liquid equilibrium concentration is accurately described by the well-mixed gas phase assumption. In reality, even for the conditions of intense gas re-circulation in which the well-mixed model is most likely to hold, this will not be true since gas concentration gradients will still exist as a result of pressure effects. The liquid is very unlikely to be in equilibrium with gas at the outlet concentration, so the actual error in $k_L a$ as a result of neglecting depth effects will be greater than predicted by equation (6.35).

(c) Plug flow model

Davies (1986) gave a set of equations describing a system with the gas in plug flow, but not accounting for depth effects. A more rigorous approach, accounting for density changes due to depth, is described by:

$$\frac{\partial}{\partial x} (QC_G) = -A(1 - \epsilon_G) k_L a (KC_G - C_L) + A \epsilon_G \frac{\partial C_G}{\partial t} \quad (6.37)$$

$$\frac{\partial}{\partial x} (\rho Q) = -A(1 - \epsilon_G) k_L a (KC_G - C_L) \quad (6.38)$$

$$\frac{\partial C_L}{\partial t} = \frac{k_L a}{H} \int_0^H (KC_G - C_L) dx \quad (6.39)$$

The following relation also holds for gas density:

$$\rho = \frac{p_O + \rho g (H-x)}{RT} \quad (6.40)$$

The partial differential equations must be converted to ordinary differential equations to allow a straightforward solution.

Davies (1986) discretised the vessel into 10 elements, producing a set of 11 simultaneous equations for the gas concentration in each element and the liquid oxygen concentration. In fact, care must be taken when choosing the number of elements used. Too many elements give numerical instabilities because of the very large concentration gradient in the bottom element at times close to the start time. Too few elements will lead to inaccuracy. Ten is a reasonable compromise number of elements; this was verified using numerical integration.

Using the discretisation technique, a set of difference equations may be formed by re-writing equations (6.37) to (6.39):

$$\frac{\partial C_G}{\partial x} = \frac{C_{Gn} - C_{Gn-1}}{\Delta x} \quad (6.41)$$

$$\frac{\partial Q}{\partial x} = \frac{Q_n - Q_{n-1}}{\Delta x} \quad (6.42)$$

For the n^{th} element, $n > 1$:

$$\frac{dC_{Gn}}{dt} = Q_n \frac{(C_{Gn} - C_{Gn-1})}{\Delta x} + C_{Gn} \frac{(Q_n - Q_{n-1})}{\Delta x} - A(1-\epsilon_G) k_L a (KC_{Gn} - C_L) \quad (6.43)$$

For the lowest element, $n = 1$:

$$\frac{dC_{Gn}}{dt} = Q_n \frac{(C_{Gn} - C_i)}{\Delta x} + C_{Gn} \frac{(Q_n - Q_i)}{\Delta x} - A(1-\epsilon_G) k_L a (KC_{Gn} - C_L) \quad (6.44)$$

The overall mass balance, using equation (5.39) becomes:

$$Q_n = \frac{-RTk_L a A(1-\epsilon_G)(KC_{Gn} - C_L)\Delta x + Q_{n-1}(p_o + \rho_L g(H-x))}{p_o + \rho_L g(H-x) - \rho_L g\Delta x} \quad (6.45)$$

Again, $Q/n = 0 = Q_i$.

Finally the expression for the liquid concentration is (for m elements):

$$\frac{dC_L}{dt} = \frac{k_L a}{L} \sum_{n=0}^m (KC_{Gn} - C_L) \Delta x \quad (6.46)$$

This can be solved to give the liquid concentration time history. Again, the errors involved in neglecting depth effects can only be evaluated numerically. Even the case where the gas concentration reaches steady state quickly ($\frac{dC_{Gi}}{dt} = 0$) produces

very complex equations with no ready analytical solution.

The effects of nitrogen transfer can be accommodated by setting a simultaneous system of equations for nitrogen, which are coupled with the equations for oxygen through the effect of mass transfer on gas flow rate. Solution of the resulting equations showed that for the range of $k_L a$ values covered by this study, the difference in $k_L a$ resulting from the inclusion of nitrogen transfer was of the order of 1%. Nitrogen transport has been neglected in all subsequent analysis using the plug flow model.

(d) Double Response Model

The residence time independent technique for analysis of transient response data was initiated by Chapman, Gibilaro and Nienow (1981), whilst the most rigorous analysis so far has been presented in Davies' derivation. The detailed derivation is not of concern here, it suffices to say that by assuming only perfect liquid phase mixing and negligible resistance to mass transfer in the gas phase, the following equation may be obtained:

$$\frac{dC_L}{dt} = \frac{k_L a}{V_G} \int_{V_G} KC_G dV_G - k_L a C_L \quad (6.47)$$

Davies and Chapman et al then used the following equation for the overall mass balance:

$$Q \int_0^t C_{Gi} - C_{Go} dt = V_L C_L + \int_{V_G} K C_G dV_G \quad (6.48)$$

If one considers compressibility effects, this should be rewritten as:

$$\int_0^t (Q_i C_{Gi} - Q_o C_{Go}) dt = V_L C_L + \int_{V_G} K C_G dV_G \quad (6.49)$$

Then unknown integral of the gas phase concentrations can now be eliminated, producing:

$$\frac{dC_L}{dt} = \frac{k_L a K}{V_G} \int_0^t Q_i C_{Gi} - Q_o C_{Go} dt - \frac{k_L a K V_L}{V_G} C_L - k_L a C_L \quad (6.50)$$

This may be simplified to:

$$\frac{dC_L}{dt} = k_L a \left[\frac{Q_i}{V_G} K \int_0^t C_{Gi} - \frac{Q_o}{Q_i} C_{Go} dt - \left(1 + \frac{K V_L}{V_G} \right) C_L \right] \quad (6.51)$$

At this stage, it is necessary to normalise the equation, in the manner first described by Chapman et al (1981). Another assumption must be made: that the gas volumetric flow rate is unaffected by mass transfer, or that $Q_o = \text{constant}$. If this assumption were not made, severe experimental problems of measuring the outlet gas flow as a function of time would be encountered. As with the ideal models, it is necessary to have an expression for the liquid equilibrium concentration in absolute terms because equation (6.51) must be written as:

$$\frac{d\bar{C}_L}{dt} = k_L a \left[\frac{Q_i K C_i}{V_G C_{L\infty}} \int_0^t 1 - \frac{Q_o C_{Go}}{Q_i C_{Gi}} dt - \left(1 + \frac{K V_L}{V_G} \right) \bar{C}_L \right] \quad (6.52)$$

If the liquid equilibrium concentration is the concentration in equilibrium with gas at half the liquid depth, and since,

$$\frac{Q_o}{Q_i} = \frac{H_o + H}{H_o}$$

and,

$$C_{o\infty} = \frac{C_{Gi} H_o}{H + H_o}$$

equation (6.52) now becomes:

$$\frac{d\bar{C}_L}{dt} = k_L a \left[\frac{Q_i}{V_G} \frac{2C_i}{C_i + C_{o\infty}} \int_0^t \frac{(1-C_o)}{C_{o\infty}} dt - (1+KV_L) \frac{\bar{C}_L}{V_G} \right] dt \quad (6.53)$$

The term $C_o/C_{o\infty}$ is just the normalised gas phase concentration at outlet, $\bar{C}_O$, so the only difference between equation (6.53)

and earlier expressions is the factor $\left(\frac{2C_i}{C_i + C_{o\infty}} \right)$, which

corrects for depth effects.

As a consistency check, it may be shown that if $C_{L\infty} = KC_i$, the same result as earlier workers is obtained. If $C_{L\infty} = KC_o$, the factor becomes $\left(\frac{C_i}{C_{o\infty}} \right)$. If the original assumption about liquid

equilibrium concentration holds, then

$$\begin{aligned} f &= \frac{2}{1 + \frac{C_{o\infty}}{C_i}} \\ &= \frac{2}{1 + \frac{H_o}{H + H_o}} \end{aligned} \quad (6.54)$$

For $H_o = 10.3m$, $f = 1.03$ for $H = 0.61m$ and $f = 1.12$ for $H = 2.7m$. There may be appreciable systematic errors in large scale experiments if depth effects are neglected. It could be argued that at very large scale, the method becomes invalid in any case

since the assumption of perfect liquid phase mixing is no longer tenable. This point is discussed in a later section.

It is difficult to see how the double response method can be modified to account for nitrogen transfer. The equations for other models which account for nitrogen transfer are coupled through equations describing the change of gas flow due to mass transfer. In deriving the double response equation, the change of gas flow due to mass transfer has been implicitly neglected, hence it is not possible to write a set of equations accounting for species other than the key component.

6.2 OXYGEN CONCENTRATION AT EQUILIBRIUM

At first sight, the assumption that the liquid phase is in equilibrium with gas at half the liquid depth appears reasonable, but further thought is necessary to justify the assumption. Lakin et al (1978) pointed out that the equilibrium oxygen concentration is not simply that at equilibrium at the mid-depth. Their expression for the saturated liquid concentration was:

$$C_{SAT} = \frac{K}{RT} \frac{\left(\frac{p_i - p_o - \zeta p_i}{\ln \left(\frac{p_i}{p_o} \right)} \right)}{1 - \zeta} \quad (6.55)$$

where

$$\zeta = \left[1 - \frac{\Omega}{\rho g} \frac{p_i - p_o}{\ln \left(\frac{p_i}{p_o} \right)} \right] \exp \left(\frac{\Omega p}{\rho g} \right) \quad (6.56)$$

$$\text{and } \Omega = \frac{k_L a RT}{(\text{He}) v_b}$$

He = Henry's law constant

The log-mean concentration is therefore the limiting case for $z = 0$. z cannot be evaluated other than by experiment, which involves measuring the saturation concentration and calculating z for use in analysis of the transient. Equation (6.55) is not therefore of use for predicting the saturation level in stirred tanks.

Jackson and Shen (1978), in considering aeration of deep tanks by sparging alone, arrived at the conclusion that for saturated conditions, the mid-depth gas phase concentration determines the saturation value. This was not based on any detailed modelling work.

The situation is rather unsatisfactory, but can be resolved in two ways.

- 1 The plug flow model of the gas phase can be solved for known $k_L a$ and other parameters to give the saturation value of oxygen concentration at equilibrium. This accounts for the effects of absorption of oxygen at lower levels and desorption of oxygen at higher levels.
- 2 The use of absolute dissolved oxygen measurement in the largest available vessel ($T = 2.7\text{m}$) to compare with the model predictions.

The results of the mathematical modelling procedure are given in Table 6.2, for values of $k_L a$ and ϵ_G inferred from existing correlations in the 2.7m vessel in tap water. Impeller speed and gas flow rate were measured at the time of carrying out the D.O. measurements. The D.O. concentration was measured using a Yellow Springs Instruments Probe, Model No 5839, with built in temperature and pressure compensation. These probes are relatively slow to respond but give measurements of dissolved oxygen concentration accurate to approximately ± 0.1 ppm by weight. The probe was mounted on a sliding arrangement attached to a baffle, and was held downwards at 45° facing the agitator

to prevent air bubbles attaching to the probe membrane. A protective cage around the membrane was removed to prevent air being trapped on the membrane. The reading with gas bubbles present was found to be the same as the ungasged reading, validating the method. The pressure compensation device was a membrane located on the side of the probe body (Fig 6.2). The distributors of the instrument stated that the pressure compensation was present to provide a direct reading of local oxygen concentration. From the construction of the probe, the pressure compensation appears to be present only to prevent distortion of the measuring membrane, which may affect the reading.

The probe was calibrated in humid air before use. The results are summarised in Table 6.3. The readings showed definite supersaturation of the liquid with oxygen. There was a slight tendency for the D.O. concentration to increase with increasing impeller speed at constant gassing rate.

This is expected from differentiating equation (6.55) with respect to ζ , assuming that increasing $k_L a$ leads to increases in ζ . There was also a tendency for the concentration to fall slightly as gassing rate increased.

The model predictions show that the equilibrium concentration should be virtually independent of gassing rate and $k_L a$. The trends shown by the experimental data are minor, and can be neglected for the range of operating conditions under consideration. Only the absolute value of equilibrium concentration is of interest.

The experimental data show approximately 9% higher oxygen concentrations than the value predicted at atmospheric pressure. This is consistent with gas being dispersed only in the region above the agitator. The distribution of gas must play some part in determining the degree of supersaturation, since the lowest point to which gas is dispersed determines the

maximum oxygen concentration. There is no way of predicting this phenomenon from existing data; the simplest realistic approach is to use the gas concentration at the mid point between the impeller and the dispersion surface to calculate the equilibrium liquid concentration.

The situation is clearly very complex, since, strictly speaking, the saturation concentration should be calculated from a first estimate of $k_L a$. This would then be used to calculate a $k_L a$ value from a best fit to the measured response. The new $k_L a$ value would then be used to calculate a new saturation concentration and so on. It is proposed instead that a simple estimation of the saturation concentration be made using the assumption outlined above, followed by a curve fitting procedure to obtain $k_L a$. The dependency of saturation concentration on $k_L a$ appears to be weak, and one should be wary of introducing too many refinements when major uncertainties such as the state of gas mixing remain.

6.3 EXPERIMENTAL TECHNIQUE

It was originally intended to use the double response technique for all the mass transfer experiments. The experimental procedure for the 0.61m and 2.67m tanks is described in detail in this section.

6.3.1 0.61m vessel

The liquid charge was stripped of oxygen using a stream of pure nitrogen. When the dissolved oxygen concentration had been reduced to approximately zero, the nitrogen flow rate was set to the test value and the output from the oxygen probes was sampled by computer over a one minute period to establish a zero level and to check that the signal noise was within acceptable limits.

If the noise level was acceptable the gas supply was switched from nitrogen to air at the same flow rate (Fig 6.3). This was

achieved by using two solenoid valves which operated simultaneously; one to close off the nitrogen supply, the other to open up the air supply. The computer sampling routine was also triggered simultaneously to sample the oxygen concentration transients. The entire operation was carried out using a single electrical signal to ensure that the valves operated at the same time that the computer sampling began.

The transients were sampled over a five minute period to ensure that the system reached equilibrium. 4096 data points were taken for each channel, corresponding to a sampling frequency of 12.5 Hz.

A check can be made on whether the length of the sampling period was adequate. The minimum $k_L a$ measured was $0.02s^{-1}$. At such low values of $k_L a$, the no depletion model can be assumed to apply, and the normalised probe response is given by:

$$\bar{C}_L = 1 - e^{-k_L a t} \quad (6.57)$$

For a first order process with $\tau = \frac{1}{k_L a} = 50$ seconds, the

response for the i^{th} multiple of the characteristic time is:

$$\bar{C}_L = 1 - e^{-i} \quad (6.58)$$

$$\text{Since, } \bar{C}_L = 1 - e^{-t/\tau} \quad (6.59)$$

For a sample period of 327 seconds, $i=6$, giving a normalised response value of 0.998. This is sufficiently close to 1 to be acceptable, in view of the inherent measuring error from the oxygen electrodes. The equilibrium value was taken as the average reading over the last 400 samples (corresponding to 32 seconds, 10% of the total response time).

The instrument chain is illustrated in Fig 6.4. The oxygen probe produced a 0-10mV output proportional to oxygen

concentration which was converted in an amplifier unit to a signal in the approximate range -5V to -4V to be compatible with the analogue to digital converter on the computer. The signal was then filtered at 1-2Hz to remove electrical noise. The full scale deflection of the oxygen probe corresponded to approximately 800 bits of the A-D converter, giving a resolution better than the signal noise. Further amplification of the signal would have been counter-productive, since the additional signal resolution would be accompanied by a corresponding increase in noise.

The liquid temperature was recorded by a thermocouple probe at the beginning and end of each experiment. The liquid temperature for the tests in electrolyte was not recorded for all tests.

A simple calculation showed that for water at normal ambient temperatures, a power input of 2.0 W/kg (beyond the maximum achievable in these experiments), the temperature rise over a 5 minute period is approximately 0.15°C , this neglects heating effects due to the gas. The gas, being compressed from atmospheric pressure, has therefore a temperature above ambient. Heat is therefore transferred from gas to liquid. On the other hand, some water will probably evaporate into the gas stream, which produces a cooling effect. Even when combined with the isothermal expansion power of the gas, these effects are small in comparison to the power dissipated by the agitator. For the electrolyte tests, variation in temperature should not have been sufficiently large to produce spurious trends in the data.

The oxygen probe positions are shown in Fig 6.5. The liquid probe was placed at the same height as the agitator to be in the region of maximum liquid velocity. This minimised the chance of bubbles attaching to the probe membrane. Air bubbles cause noisy signals, since when the water is only partly re-oxygenated, an air bubble attaching to the membrane causes the

apparent local partial pressure of oxygen to rise suddenly to saturation, and to fall back to the partly saturated value when the bubble disengages. The actual effect is not so dramatic as this description implies, since the membrane is unlikely to be completely covered in gas, and the probe response time tends to damp out the fluctuations.

6.3.2 2.7m vessel

Every effort was made to use the same experimental procedure as in the 0.61m tank. A difficulty arose because of the nitrogen requirement. The small scale experiments were carried out using bottled nitrogen, which was clearly impractical for a vessel 85 times the volume. A mobile liquid nitrogen storage facility was hired from the British Oxygen Company (BOC). The ambient heat exchangers used to evaporate the liquid nitrogen were modified to increase the flow rate available. Ice formation on the evaporators did not prove a problem and steady flow rates were obtainable.

Owing to the cost of liquid nitrogen it was not possible to use pure nitrogen to de-oxygenate the vessel. A mixture of approximately 50:50 nitrogen:air was used. Fig 6.6 illustrates the gas supply system

The procedure was as follows:

- 1 Start up nitrogen flow with valve B closed.
- 2 Control nitrogen flow to desired value using valve C. Measure flow using orifice plate D.
- 3 Open valve B and control air flow using valve B until desired total flow rate is reached.
- 4 Switch three-way valve to air only and close valve B to prevent nitrogen entering air stream. Simultaneously

trigger computer sampling by manual electrical switch. The air flow, measured by orifice plate E, was previously set using control valve A to match the mixture flow rate.

Note that the air and nitrogen flow rate in the mixture were not separately monitored due to difficulties of arranging the pipework to allow proper flow measurement. The composition of the mixture was not therefore available from the flow measurements. Simply subtracting the nitrogen flow measured in Step 2 from the total mixture flow is not sufficiently accurate. The nitrogen flow must be reduced by adding air because the pressure just downstream of valve C will increase as a result of the increased gas flow through the three-way valve which appeared to control the pressure drop in the system. The pressure upstream of valve C was controlled by regulators.

Fortunately, the exact mixture composition is not significant as long as the starting oxygen concentration in the vessel is accurately known. This information was provided by the oxygen probes.

The liquid oxygen probe was mounted on a holder which slid up and down one of the baffles on a rail (Fig 6.7). This was done to allow the probe to be inspected without draining the vessel. The probe was positioned at the same height as the agitator, and the membrane was at about the same radius as the inner edge of the baffle. The arrangement minimised intrusion into the vessel which may have been a problem had an alternative position been chosen, and was virtually the only practical option. It was not possible to mount the probe in the vessel wall, as was the case in the 0.61m tank.

6.3.3 Oxygen concentration

The oxygen concentration in both gas and liquid was measured using fast-response polarographic electrodes manufactured by Electrosense Ltd, Frodsham, UK, and were identical in

construction to those used by Davies (1986). The probe is illustrated schematically in Fig 6.8.

The probes measure oxygen concentration as a result of the reaction of oxygen at the cathode when a polarising voltage is placed across the cell. The current produced as a result of the reaction is proportional to the rate of reaction of oxygen, which is replenished by diffusion through the membrane. The rate of diffusion is itself governed by the concentration gradient of oxygen from the measuring medium to the cathode. Linek and Benes (1977) considered the properties of this type of oxygen probe, especially with respect to the transient response to a step change in oxygen concentration. The cell provides a direct measure of oxygen concentration in the medium, and whilst the true transient response to step changes in oxygen concentration is governed by a complex set of resistances to diffusion, the response is, to a good approximation, first order, as shown by Davies (1986). The more complex model of Linek and Benes can be reduced to a first order model provided that diffusion across the membrane is the limiting factor. The ElectroSense probes are constructed such that an extremely thin layer of electrolyte exists between the membrane and the cathode, minimising diffusional resistance in the electrolyte.

The existence of a liquid boundary layer must also be considered. A concentration boundary layer in the liquid phase provides another series resistance to diffusion, which affects the transient response of the system. For water-like substances, this resistance is implicitly assumed to be negligible; problems can and do arise in liquids of even moderate viscosities.

Experience showed that considerable care was required in assembling the probes, and various practical difficulties were encountered. Briefly, there were unexpected corrosion products deposited on the electrodes which had adverse effects on the probe, causing loss of sensitivity and very slow response to

changes in oxygen concentration. The probes also showed a tendency to drift. Various solutions were explored, including investigating the electrolyte within the cell, but the problem was eventually solved by reversing the polarity of the probe for a few minutes before use. This gave fast, reproducible responses which were free of drift.

The time constant of the probes was measured by placing them in a stream of pure nitrogen. When the reading reached zero, a computer sampling program was automatically triggered. The probe was then removed suddenly from the nitrogen stream into air, and the transient recorded on the computer. The data was then analysed using regression analysis to obtain the time constant of the probe. The computer program also performed rudimentary tests for drift and noise. The probes were also checked for drift by visual inspection of the output registered on the computer over at least the length of time taken by one experiment. Only when these checks on the probe had been successfully completed was the probe used for mass transfer experiments.

The probe first order response times were in the range 0.5-1.3 seconds, with 1.0 second being a representative average.

6.3.4 Experimental problems

Largely as a result of the difficulties encountered with the oxygen electrodes, it was not possible to acquire data using the double response method in the 0.61m tank. The gas phase response was very difficult to reproduce, and the oxygen probe appeared to be drifting over the course of the experiment. Although various steps were taken to investigate the problem, it was not possible to obtain gas phase responses of the required standard. In particular, the oxygen mass balance was not satisfied. By measuring the gas phase oxygen concentration transient, the mean residence time theorem may be invoked to check the oxygen mass balance:

$$V_G C_G + V_L C_L = \int_0^t (Q_i C_{Gi} - Q_o C_{Go}) dt \quad (6.60)$$

By assuming that $Q_i = Q_o$ and $C_{L\infty} = KC_i$:

$$\frac{KV_L + V_G}{Q_i} = \int_0^\infty (1 - \bar{C}_{Go}) dt \quad (6.61)$$

A linear correction for drift was attempted which involved cutting off the gas phase response at various times until equation (6.61) was satisfied. The drift correction was so large that it was eventually concluded that the data were probably unreliable in any case. The problems with the electrodes probably contributed to the drift of the gas phase probe.

However, the liquid probe responses were drift free, enabling $k_L a$ values to be obtained by assuming an ideal model for the gas phase RTD.

As a result of the limitations on power input, the value of $k_L a$ measured in this work never exceeded 0.1 s^{-1} . The amount of depletion in the gas phase was therefore relatively small, and it is the amount of depletion which determines the error arising from assuming one of the ideal gas phase RTD models.

The plug flow model was chosen for the following reasons:

- 1 The data of Davies (1986) show that the plug flow model is a reasonable approximation to the double response model for $k_L a$ values less than 0.1 s^{-1} . Furthermore, Fig 6.9 shows that the plug flow model is the best approximation of the three available even at higher values of $k_L a$. This figure is a little misleading, not all Davies' data showed such good agreement at high $k_L a$ values, but the plug flow model appeared to be free of the spurious trends shown by the no depletion and well-mixed models.

- 2 It was desirable to be consistent with the large scale experiments, for which plug flow of the gas is likely to be the most reasonable approximation as a result of the tendency for gas re-circulation to decrease with increasing scale.

The large scale experiments were also carried out with the intention of using the double response method. Once again, it was possible to check the overall oxygen mass balance.

Neglecting depth effects, the mass balance may be written as:

$$\int_0^{\infty} (1 - \bar{C}_{Go}) dt = \frac{KV_L(1 - \bar{C}_{Li}) + V_G(1 - \bar{C}_{Gi})}{Q_i} \quad (6.62)$$

$\bar{C}_{Li}$, $\bar{C}_{Gi}$ are the initial normalised values of the liquid and gas

probes respectively, since the starting condition of the experiments was a non-zero oxygen concentration. If depth effects are considered, the equation becomes:

$$\frac{2}{1 + \frac{H_o}{H + H_o}} \int_0^{\infty} (1 - \bar{C}_o) dt = \frac{KV_L(1 - \bar{C}_{Li}) + V_G(1 - \bar{C}_{Gi})}{Q_i} \quad (6.63)$$

Integrating the gas phase response produces a check on the mass balance. Once again, the mass balance did not hold; examination of the response showed that there was a small amount of drift, although the probe had been extensively checked outside the vessel for drift, and the technique of reversing the polarity of the electrodes had been introduced at the time of carrying out the experiments. The response was integrated from a start time calculated from the trace, not from the start time of the experiment, so the error was not due to any pure lags in the system.

Increased values of the first order response time of the gas probe τ_G , were used in a similar way to the method of Chapman, Gibilaro and Nienow (1981), who used a much deeper funnel which produced back-mixing, tending to increase the effective first

order response time of the gas probe. This technique did not succeed in this instance, largely because large amounts of area were in regions of slow change of signal, so de-convoluting with increased values of τ_G had little effect on the total area.

The plug flow model already discussed was used to predict the outlet gas concentration values as a function of time. The time taken for the outlet concentration response to begin changing rapidly was virtually identical to that shown by the experimental response. This is a piece of supporting evidence for the hypothesis that the gas is in plug flow since for significant amounts of back-mixing in the residence time distribution, the breakthrough would occur more quickly for the experimental response than predicted by the plug flow model.

6.4 PERFECT LIQUID PHASE MIXING

The assumption that the liquid phase is fully back-mixed lies behind all the analysis techniques. The problem has been considered for bubble column contactors by Andre, Robinson and Moo-Yung (1983), who assumed a plug flow RTD for the gas phase and modelled the liquid phase as a chain of stirred tanks in series. By solving the resulting sets of differential equations, the following criterion for the validity of perfect liquid phase mixing was obtained:

$$k_L a \tau_C \leq 2 \quad (6.64)$$

It is thus possible, with a knowledge of the value of the liquid circulation time, to judge the validity of the well-mixed liquid assumption.

There is extensive literature on single phase mixing times and circulation times, such as the work of Khang and Levenspiel (1976), Ruszkowski and Muskett (1985) and Middleton (1977). Little work has been published for two-phase gas-liquid systems, with the exception of Pandit and Joshi (1983) and Einsele and

Finn (1980) on mixing times, and Bryant and Sadeghzadeh (1979) on circulation times.

Pandit and Joshi (1983) considered the circulation velocity generated by the impeller and by the action of gas sparging. At low agitator speeds, the mixing time decreased with increasing gassing rate, as a result of the increased liquid circulation induced by the gas. At high agitator speeds, the mixing time increased with increasing gassing rate as a result of the decreased pumping action of the agitator. Einsele and Finn (1980) also measured mixing times in gassed systems, and found that, for a tank of 0.76m diameter:

$$\frac{\theta_a}{\theta_m} = 1 + 7.5 \mu^{0.27} \epsilon_G \quad (6.65)$$

θ_a = aerated mixing time

θ_m = unaerated mixing time

Hence, the aerated mixing time tended to increase with increasing gassing rate. For water at a gas hold-up of 0.1, the increase in mixing time due to gassing would be in the region of 10%. The data of Pandit and Joshi show that much larger increases in mixing time are possible. Bryant and Sadeghzadeh (1979) found consistent increases in circulation time with increasing gassing rate.

The overall picture from the literature is that for most regimes of practical interest, the liquid circulation time increases with increasing gassing rate, but the magnitude of the increase is very difficult to predict, as are effects of scale.

As a result of these uncertainties, it was decided to conduct some experiments in the 0.61m vessel to investigate the effect of gassing on liquid mixing times. A conductivity method was used, described by Ruszkowski and Muskett (1985). The main experimental problem was the tendency for gas bubbles to attach

to the conductivity probe (Fig 6.10), which was avoided by placing the probe at the lowest point in the vessel and using tap water. The probe therefore did not come into contact with large gas bubbles and a noise-free signal was obtained.

A 2.4 M solution of sodium chloride was used as a tracer. The quantities of tracer used did not appear to affect the coalescence properties of the liquid: no changes in gas hold-up or consistent trends in the mixing time measured by successive injections were observed. It was not possible to measure the mixing time in a non-coalescing system, because of the presence of small gas bubbles in the region of the conductivity probe.

The tracer was injected by rapidly emptying 25cm³ of solution onto the dispersion surface and triggering the sampling routines on a PDP 11/23 computer using a manual switch. Subsequent experiments in a 0.945m diameter vessel using a solenoid valve and automatic triggering at the moment the tracer was delivered to the surface produced similar results to the manual method. The experiment was repeated 6 times for each data point to remove the random errors caused by the use of such a relatively crude experimental method.

The conductivity probe response was analysed to give an end-point when the variance of the tracer about the final mean level fell to a value corresponding to 95% of the total step change in conductivity having taken place.

The results are shown in Fig 6.11. The surprising feature of the data was the tendency for the mixing time to reach a maximum at low gassing rate, and to fall to an asymptotic value which was higher than the ungassed mixing time at high gassing rates. A possible explanation for this behaviour is the cavity structure behind the agitator blades. At low gassing rates, clinging cavities form which grow rapidly with increasing gassing rate, producing a sharp fall in power demand and, presumably, a sharp fall in impeller pumping capacity. Once the

"3-3" cavity structure has formed the increased liquid circulation caused by the gas flow may then reduce the mixing time from its maximum value. One experiment was conducted with a stationary agitator to check the influence of the impeller on the mixing time. Even though, at the low speed, the impeller was nearly flooded, the mixing time without mechanical agitation exceeded the agitated mixing time by a factor of 2.

In fact, if one applies the correlation of Warmoeskerken (1986) for the point at which the transition from clinging to "3-3" cavities takes place to the data of Bryant and Sadegkyadeh and Pandit and Joshi the trends in the reported data cannot be adequately explained in terms of cavity structure. The interaction between the various phenomena are clearly very complex and a great deal of further work would be required to elucidate the mechanisms fully. For the purposes of the present exercise, if one assumes that the gassed mixing time exceeds the ungassed mixing time by a factor of two, the estimate will be a rather conservative one.

For ungassed systems, confidential data from BHRA's Fluid Mixing Processes project (FMP) has shown that at least for vessels up to 2.67m in diameter:

$$No_m = \text{constant} \quad (6.66)$$

From the ungassed, 0.61m data, the constant is approximately 27.

Norwood and Metzner (1960) and Holmes, Voncken and Decker (1964) used the following relationship between mixing time and circulation time:

$$\theta_m = 5\tau_c \quad (6.67)$$

In this project, the 95% mixing time was measured, whereas Norwood and Metzner and Holmes et al used the 99% mixing time.

From the theoretical relationship between mixing time and criterion for mixing, it may be shown that:

$$\theta_{95} = \frac{\theta_{99}}{1.54} = 3.247\tau_C \quad (6.68)$$

This leads to:

$$0.31 k_L a \theta_{95} \leq 2 \quad (6.69)$$

or

$$k_L a < 0.24N \quad (6.70)$$

This relationship was indeed satisfied for all the 2.67m vessel data. The assumptions behind the modified condition for complete liquid phase mixing are extremely conservative, especially the result that the gassed mixing time is equal to twice the ungassed mixing time. This must be balanced against the fact that the original work on which the criterion was based was purely theoretical and involved the assumption of a gas phase RTD.

It is therefore concluded that the perfect liquid mixing assumption is valid for the experiments conducted in this study.

6.5 DATA ANALYSIS

The differential equations describing the plug flow model have already been discussed. The value of $k_L a$ from the liquid phase oxygen time history was calculated by solving the equations numerically to produce a calculated response. The value of $k_L a$ was optimised to give the minimum error sum of squares between the calculated and experimental responses.

Briefly, the computer program operated as follows:

- 1 Calculate starting value of $k_L a$ by assuming no depletion and using linear regression on the measured response.
($k_L a$)₁.

2 Calculate second starting value of $k_L a$ using plug flow model with no gas transient $(k_L a)_2$.

3 Calculate third starting value of $k_L a$ using

$$(k_L a)_3 = (k_L a)_2 + [(k_L a)_2 - (k_L a)_1]$$

4 Calculate error sum of squares for each starting value.

5 Carry out search to check that the minimum value of the error sum of squares is at $(k_L a)_2$.

6 Fit parabola to calculated values of error sum of squares and differentiate to find minimum.

7 Discard value of $k_L a$ furthest from minimum and repeat fitting procedure.

8 Repeat steps 6 and 7 until convergence is achieved.

It was essential to obtain a reasonably accurate estimate of the start time of the experiment since this affected the quality of the fit to the experimental data. The subroutine to establish the start time is illustrated in Fig 6.12. Great care was taken to avoid false starts due to freak noise.

The point at which the fitting process ended was also investigated. The transient up to 90% of the total step change was used for comparison with the calculated value to save computation time. Comparison with a similar analysis up to 98% of the total step change showed very little change in $k_L a$, so the 90% method was used since this approximately halved the calculation time.

It was observed that on some occasions the calculated value of $k_L a$ using the plug flow model was lower than that obtained by linear regression of the response assuming a no depletion model. This was anomalous because the no depletion value of $k_L a$

is the minimum possible value of $k_L a$ which can be obtained from an experiment. Curve fitting using the no depletion model was employed, and the value of $k_L a$ was lower than that obtained using the curve fitting plug flow technique. There were differences between the $k_L a$ values calculated by regression and those obtained by curve fitting of up to 10%. The regression values varied by about 5% depending on which part of the transient was used for the regression. Of course, the regression technique neglects the probe response time, and the curve fitting procedure was carried out with the equation to de-convolute the probe response:

$$C_L = C_{LM} + \tau_L \frac{dC_{LM}}{dt} \quad (6.71)$$

Equation (6.71) was solved simultaneously with the equations describing the gas and liquid phase responses, and the transient of C_{LM} was actually fitted to the experimental data.

The differences between the $k_L a$ values obtained by the two methods are considered to have been of only minor significance since the only function of the regression calculation was to generate a reasonable set of starting values for $k_L a$.

The numerical solution of the differential equations describing the plug flow model was effected using a fourth-order Runge-Kutta technique.

A preliminary sensitivity analysis was carried out to check that errors were not introduced as a result of the time step being too long (which can produce instabilities) or as a result of the number of sections used in the discretisation of the tank. The predicted liquid response was fairly insensitive to these parameters, although differences could be observed in the predicted gas concentrations. It was concluded that the model was reasonably robust with respect to small changes in the numerical technique and was not likely to suffer from instability.

6.6 TEMPERATURE CORRECTION

Smith, van't Riet and Middleton (1977) proposed the use of a temperature correction of the following form:

$$\frac{k_L a \mu}{(\theta + 273)} = \text{constant} \quad (6.72)$$

This was confirmed by Smith et al experimentally, and was used to correct their data to 20°C. A standard temperature of 15°C is more appropriate to this work, and the temperature correction is given by:

$$(k_L a)_{15} = (k_L a)_\theta \frac{\mu_{15}}{\mu_\theta} \left(\frac{273 + \theta}{288} \right) \quad (6.73)$$

The correction is consistent with the Stokes-Einstein relation:

$$\frac{D_L \mu}{273 + \theta} = \text{constant} \quad (6.74)$$

provided that $k_L a \propto D_L$, which is the film diffusion model of mass transfer. Equation (6.75) has been confirmed by Wilke and Chang (1955) and Sridhar and Potter (1977) as a special case of wider correlations involving solute properties such as molar volume. Thus equation (6.72) appears to have a sound theoretical and experimental basis. It has been used in preference to the purely empirical correction used by workers such as Jackson and Shen (1978) and Lakin et al (1978):

$$(k_L a)_{20} = 1.024^{(20 - \theta)} (k_L a)_\theta \quad (6.75)$$

Equation (6.73) is considered to be applicable over a wider range of temperature and is therefore more acceptable than equation (6.75).

6.7 SURFACE AERATION

The effect of surface aeration on the measurement of $k_L a$ has been neglected in deriving the equations to predict the oxygen concentration transient in the liquid phase. This is in accordance with the general view in the literature.

Fuchs, Ryu and Humphrey (1971) found that surface aeration produced a far smaller contribution to $k_L a$ than sparging, and stated that the contribution declined with scale. Chapman, Nienow and Middleton (1980) concluded that even in a vessel of 0.22m diameter, surface aeration was negligible.

Gassing reduces the pumping capacity of the agitator and the presence of gas bubbles affects the formation of surface vortices. Surface aeration is suppressed for a gas liquid system in comparison to an ungassed system at the same agitator speed.

6.8 RESULTS

Mass transfer data was obtained for the following experimental conditions:-

Vessel Diameter (m)	Impeller Type	Liquid Type	Table Number for Raw Data
0.61	Rushton (T/3)	Water	6.4
		Electrolyte	6.5
	Concave (T/3)	Electrolyte	6.6
2.67	Rushton (T/3)	Water	6.7
	Concave (T/3)	Water	6.8

The data have been correlated using the $\bar{\epsilon}_m$, v_{SG} method (Tables 6.9 and 6.10) and the N, Q method (Tables 6.11 and 6.12).

Generally, the individual data sets were correlated very well, with high correlation coefficients and maximum errors within 10%, using both the $\bar{\epsilon}_m$, v_{SG} method and the N, Q method. The exception was the 2.67 m data obtained with the Rushton turbine. Although the maximum error was only about 10%, the correlation coefficient was very low as a result of scatter in the data. The time histories of the liquid phase oxygen concentration showed considerable signal noise. This was thought to be caused by:

1. Inhomogeneity of the liquid phase. Although the criterion of Andre, Robinson and Moo-Yung (1983) was satisfied, it is quite possible that local inhomogeneity of the liquid phase with respect to oxygen concentrations at the large scale caused signal noise.
2. Gas bubbles may have attached to the oxygen electrode. This is known to be a cause of noise problems. The probability of a gas bubble attaching to the electrode is governed at least partly by the local liquid velocity. For increasing scale, the liquid velocity tends to increase as the impeller tip speed increases, neglecting the effects of gassing on pumping rate. On that basis, the large scale data should have been less prone to noise caused by gas bubbles than the small scale data, although it should not be forgotten that the relative positions of the liquid oxygen probe were different on the two scales. At small scale, the probe was set into the vessel wall mid-way between the baffles at the same height as the agitator. For the large scale work the probe was, effectively, attached to a baffle at the same height as the agitator.

A notable feature was that the noise became progressively worse as the gas flow rate increased. Since the tests were carried out in tap water, one would expect that near the flooding point, there would be very little gas in the vicinity of the probe. The flow pattern changes of Figure 2.2 show this effect. This

implies that the noise problems were not the result of gas bubbles attaching to the oxygen probe, but were caused by poor liquid mixing as a result of the reduced impeller pumping capacity near the flooding point.

The same problems were encountered in the 0.61 m vessel at the lowest agitator speed (3.5 rps). Some of the liquid oxygen concentration time histories were very noisy and were rejected for correlation purposes. This may also have been caused by poor liquid phase mixing.

The flooding correlation of Nienow et al (1985) (Equation 2.9) was used to predict the gassing rate at which the agitator would flood for a given gassing rate. It was confirmed that the gassing rate above which the noise problems became serious was in the region of the predicted flooding point. It is difficult to discern clear trends from the Rushton turbine data, which are summarised in Table 6.13 to show the presence or absence of serious noise problems. There is a clear tendency for the noise problems to become serious at higher gassing rates as speed is increased, which is consistent with the hypothesis that the problems were caused by poor liquid phase mixing close to the flooding point. The same behaviour was shown by the concave turbine (Table 6.14).

Naturally, the signal noise presented difficulties for the data analysis. The curve fitting technique was particularly prone to give inconsistent values of $k_L a$ for repeated experiments at the same operating conditions. For the Rushton turbine, the curve fitting method was compared, for the response considered to be sufficiently free of noise to give acceptable data, with the pseudo-steady state plug flow model which uses the slope of a semi-logarithmic plot of the data to calculate $k_L a$. The agreement between the two methods was good, and when linear regression was used on the data, the statistical hypothesis that the two methods gave the same value of $k_L a$ was found to be valid. The complex transient plug flow model with depth effects produced the same value of $k_L a$ as a much simpler model using

semi-logarithmic plots and requiring far less computational effort. One cannot generalise this finding to all experiments since the equivalence of the methods is probably due to errors in the deviation of the simple plug flow model, such as neglecting the probe response time and neglecting depth effects, cancelling each other out. However, for this particular set of data, the simple plug flow model has been used to obtain the correlation shown in Table 6.10. The curve fitting technique produced a correlation which was not statistically significantly different from the simple plug flow model but which had a lower correlation coefficient (0.81) as a result of scatter. The choice of which data set to use was purely arbitrary : the same conclusions can be drawn from using either set of data.

6.9 RELATIONSHIP BETWEEN $k_L a$ AND HOLD-UP

There is a considerable incentive to investigate the relationship between $k_L a$ and gas hold-up from the point of view of the experimenter. Hold-up experiments are very much easier to perform than mass transfer experiments, especially at large scale. It must be established whether or not the gas hold-up is a reliable indicator of the mass transfer performance for a particular set of operating conditions.

Andrew (1983) set out the following relationship between $k_L a$ and hold-up, for dirty water:

$$\frac{k_L a}{\epsilon_G} = 0.4 \quad \frac{P}{V} < 1.0 \text{ kW/m}^3 \quad (6.76)$$

$$k_L a = k \left(\frac{P}{V} \right)^{5/6} \quad \frac{P}{V} > 1.0 \text{ kW/m}^3 \quad (6.77)$$

Most of the FMP data lie in the region of $\frac{P}{V} < 1.0 \text{ kW/m}^3$. $k_L a$ is proportional to hold-up provided that the gas bubble size is approximately constant, since for uniform dispersions of spherical bubbles:

$$a = \frac{6\epsilon_G}{d_{sv}} \quad (6.78)$$

If k_L is independent of (P/V) (this point was discussed by Figueiredo and Calderbank, 1979), then $k_L a$ is proportional to hold-up provided that ϵ_G and d_{SV} depend on power input and superficial gas velocity in a similar way.

The relationship has been tested by plotting $k_L a$ against hold-up for identical operating conditions of speed and gas flow rate. Figures 6.13 to 6.15 show the data for the 0.61 m tank, and Figures 6.16 to 6.17 show the data for the 2.67 m tank. The plots for the 0.61 m tank include the linear regression line forced through the origin. All the data showed positive intercepts, in other words finite $k_L a$ at zero hold-up, which does not fit Equation (6.76), but may have some physical meaning. For the case of the vessel acting purely as a bubble column, measurable values of $k_L a$ can be obtained at almost negligible hold-up.

Almost all the small scale data lay within $\pm 20\%$ of the regression, but there was a definite effect of agitator speed, as careful inspection of Figures 6.13 to 6.15 shows. The data points tend to lie above the correlation line at high agitator speeds and below the line at low speeds. There was also a tendency for $k_L a / \epsilon_G$ to decrease with increasing hold-up at constant speed. This is equivalent to decreasing $k_L a / \epsilon_G$ at increasing gas flow rate. This may be re-expressed as a dependency of $k_L a / \epsilon_G$ on power input, since power input decreases with increasing gassing rate at constant agitator speed.

This can be seen by considering the correlation for $k_L a$ and ϵ_G in the 0.61 m tank:

$$\text{Rushton in electrolyte : } \epsilon_G = 1.46 (\bar{\epsilon}_m)^{0.18} v_{SG}^{0.63} \quad (6.79)$$

$$k_L a = 0.83 (\bar{\epsilon}_m)^{0.42} v_{SG}^{0.5} \quad (6.80)$$

$$\frac{k_L a}{\epsilon_G} = 0.57 (\bar{\epsilon}_m)^{0.24} v_{SG}^{-0.13} \quad (6.81)$$

$$\text{Rushton in water} : \epsilon_G = 1.46 (\bar{\epsilon}_m)^{0.24} v_{SG}^{0.69} \quad (6.82)$$

$$k_L a = 0.61 (\bar{\epsilon}_m)^{0.46} v_{SG}^{0.6} \quad (6.83)$$

$$\frac{k_L a}{\epsilon_G} = 0.42 (\bar{\epsilon}_m)^{0.22} v_{SG}^{-0.09} \quad (6.84)$$

For Equation (6.76) to be valid, the exponents on $\bar{\epsilon}_m$ and v_{SG} would have to be zero.

The assumption that $k_L a$ is proportional to hold-up is a fairly crude approximation to the current data. There are probably effects of gas distribution as well as bubble size to account for the poor fit to the data. Calderbank (1958) found that the Sauter mean diameter of the bubbles in a stirred tank for a coalescing liquid was given by:

$$d_{SV} = 0.0415 \frac{\sigma^{0.6}}{\rho^{0.2} \left(\frac{P}{V}\right)^{0.4}} \sqrt{\epsilon_G} + 9 \times 10^{-4} \quad (6.85)$$

For sufficiently large bubbles, one can approximate this relationship as follows:

$$d_{SV} \propto \frac{\epsilon_G^{0.5}}{\left(\frac{P}{V}\right)^{0.4}} \quad (6.86)$$

For the tap water data:

$$\epsilon_G^{0.5} \propto (\bar{\epsilon}_m)^{0.12} \quad (6.87)$$

$$\therefore d_{SV} \propto (\bar{\epsilon}_m)^{-0.28} \quad (6.88)$$

Hence the group $\frac{k_L a d_{SV}}{\epsilon_G}$ is almost independent of power input,

but is still a function of superficial gas velocity. It may well be the case that bubble size is a function of gassing rate. Without making actual bubble size measurements (which is a difficult and time consuming process), no firm conclusion can be drawn about the reason for $k_L a$ not being truly proportional

to hold-up.

The results in tap water present the closest approximation to the result quoted by Andrew (1983). Approximately:

$$\frac{k_L a}{\epsilon_G} = \frac{2}{3} \quad (6.89)$$

This was for clean tap water. The difference between this empirical equation and that of Andrew is that for "dirty water" the value of $k_L a$ is probably reduced by contaminants compared to the value for clean water.

By using: $a = \frac{6\epsilon_G}{d_{SV}}$ (6.78)

it is possible to calculate approximate mean bubble diameters for the agitators:

$$d_{SV} = 6 k_L \left(\frac{\epsilon_G}{k_L a} \right)_{\text{exp}} \quad (6.90)$$

The value of k_L has been assumed to be 4×10^{-4} m/s after van't Riet (1979). This is a fairly tenuous hypothesis, since surface properties may well change with the addition of electrolyte, affecting the value of $k_L a$, but provides a working approximation.

The bubble diameters calculated on the basis of the mean slope of the $k_L a - \epsilon_G$ plots passing through the origin are:

Rushton, water, 0.61 m tank : $d_{SV} = 3.6 \times 10^{-3}$ m
 Rushton, electrolyte, 0.61 m tank : $d_{SV} = 2.61 \times 10^{-3}$ m
 Concave, electrolyte, 0.61 m tank : $d_{SV} = 2.55 \times 10^{-3}$ m

As expected, the electrolyte values are considerably smaller than the water result.

The calculated Sauter mean bubble diameters for tap water and electrolyte are much closer in magnitude than the measurements

of Koetsier and Thoenes (1972) indicated. The presence of electrolyte clearly influences the value of k_L , since for electrolyte compared to water, the gas hold-up is approximately 40% higher and the bubble size is lower by a factor of at least two (Koetsier and Thoenes). The measured values of $k_L a$ in electrolyte exceed those in water by a factor of about 1.9, so the value of k_L must have been reduced by the electrolyte.

The large scale data (Figures 6.15 and 6.16) are considerably more difficult to interpret than the small scale data, largely because of the very restricted range of operating conditions for which $k_L a$ could be measured accurately. The linearity of the plots of $k_L a$ against hold-up is excellent, but again there is a finite positive value of $k_L a$ at zero hold-up. The data for the two impellers lie on virtually the same line, implying that for a coalescing liquid the bubble size produced by the two different agitator types is similar. The slope of the large scale plots are different from the small scale tap water data obtained with the Rushton turbine, with the large scale results showing consistently lower values of $k_L a$ at given gas hold-up at gas hold-ups $> 6\%$. These are probably scale effects on bubble size for a coalescing system since circulation times increase with increasing scale, this allows more time for coalescence to occur, and the bubble size will tend to increase. The characteristic time for coalescence is a critical parameter in determining the validity of this argument. If it is very much shorter than the circulation time, scale effects or circulation time will not influence bubble sizes.

A second hypothesis to explain this apparent scale effect is that the distribution of energy dissipation (and, thus, according to Kolmogoroff's theory of isotropic turbulence, assuming that this holds for a stirred tank) the local turbulent length scales may well change with scale. In regions of the vessel away from the impeller, especially when the gas is in plug flow, the bubble size will be determined by the relative rates of bubble break-up and coalescence. If break-up is

determined by local turbulence, then the bubble size could well be affected by changes in the distribution of energy dissipation rate. There is little firm evidence to support this argument but it is a more probable explanation of the trends shown by the data than the explanation based on circulation time.

6.10 COMPARISON OF AGITATOR TYPES

The data allow the direct comparison of the Rushton and concave turbines.

In the 0.61 m vessel, in electrolyte, the concave turbine data were statistically significantly different from the Rushton turbine data, whilst the hold-up correlations were very similar.

This is reflected in the values of $k_L a / \epsilon_G$ which were generally higher for the concave turbine than the Rushton turbine, implying that although the hold-ups were similar, the concave turbine was producing slightly smaller bubbles than the Rushton turbine. This is shown in the results of the previous section on effective bubble sizes.

Figure 6.18 is a plot of the $k_L a$ data as a function of power input at constant superficial gas velocity for the two agitator types, showing that the increased values of $k_L a$ obtained with the concave turbine, although statistically significant, are of little practical importance.

In the 2.67 m vessel, in water, there was a small, but statistically significant difference between the Rushton and concave turbines, whilst the hold-up correlations were not significantly different from each other.

6.11 SCALE-UP

As for the hold-up experiments, the $k_L a$ data were correlated successfully by the $\bar{\epsilon}_m, v_{SG}$ method when the data for both the

0.61 m and 2.67 m scales were combined. The 2.67 m correlation was significantly different from the combined correlation. The $k_L a$ data for the two scales were also correlated successfully by (Tables 6.11 and 6.12):

$$k_L a = \alpha N^\beta Q^\gamma \quad (6.91)$$

The correlation coefficient for this approach was just as high as for the $\bar{\epsilon}_m, v_{SG}$ method. This was possibly the result of the heavy bias in number of data points towards the small scale data. No distinction between the two correlation methods can be made on the basis of this data.

The correlation techniques discussed in the chapter on hold-up have been applied to the $k_L a$ data. The results of including scale terms for the two correlating methods were:

$$k_L a = 0.424 (\bar{\epsilon}_M)^{0.41} v_{SG}^{0.50} D^{0.03} \quad (6.92)$$

$$k_L a = 0.056 N^{1.24} Q^{0.39} D^{0.14} \quad (6.93)$$

The improvements in fit to the data were marginal in each case, and the inclusion of a scale term cannot be justified on statistical grounds.

An argument was put forward in the hold-up chapter to show that the exponent on scale in Equation (6.93) is given by the following approximate relationship:

$$\delta = 3 (\beta - \gamma) \quad (6.94)$$

δ = exponent on scale in Equation (6.93)

β = exponent on power per unit mass

γ = exponent on superficial gas velocity

Since for virtually all the experiments reported here, $\gamma > \beta$, δ should be negative. This is not the case

(Equation 6.93). The hold-up data also showed deviations in the exponent on scale from the predicted value. The effects of scale and gassing rate on power input are probably the cause of the discrepancies, since it has been assumed that:-

$$\bar{\epsilon}_m = A N^3 D^2 \quad (6.95)$$

$A = \text{constant}$

The actual value of power input depends on the value of P_g/P_u , which varies sufficiently widely to cause one to question the true validity of assuming $N^3 D^2$ to be constant for given power input per unit mass.

7. FLOODING

7.1 MEASUREMENT

The flooding point of Rushton turbines has been investigated by studying the gassed power curves. At the transition from large cavities to ragged cavities, the gassed power consumption increases (as a step change). To pinpoint such a step change exactly requires a large number of experiments, which when working at large scale is extremely inconvenient. Therefore, the flooding point has been located only over a range of gas flow rates.

A typical set of gassed power curves from which flooding points were inferred is shown in Fig 4.8. The magnitude of the step change was rather variable, presumably depending on the exact point at which the first "flooded" power measurement was made. The maximum step change observed was approximately 10% of the signal, which is outside the experimental error. The power curves were otherwise quite smooth especially at gas flow rates approaching the flooding point. These two observations reduce the doubt that the step changes were due to changes in the signal due to extraneous factors. Unfortunately, time did not permit repetitions of the experiments, which would have provided another check.

Visual observation was only of limited value since non-coalescing liquids were being studied for the majority of the tests, and in the 2.67m vessel sideways viewing of the agitator was not possible due to the construction of the tank. However, it was possible to observe the surface of the dispersion. At gas flow rates below those required to flood the agitator, the surface is relatively quiescent, with bubbles disengaging gently across the whole cross-sectional area of the tank. When the agitator is flooded, the surface is disturbed, and large pockets of gas erupt around the impeller shaft. Between these two states there is a transition region, in which pockets of gas may

be seen intermittently breaking through the dispersion surface. The frequency of the appearance of these gas pockets increases as the gas flow rate is increased, until the break-through becomes a virtually continuous process. Careful observation show that the initial break-through of gas was virtually coincident with the flooding point measured from the gassed power curves. Although not conclusive proof, this at least corroborates the evidence of the step changes in the gassed power curves.

Finally, it should be noted that the experiments were always conducted with gassing rate increasing. After each data point the gas supply was shut down and the mixer stopped to check the zero level on the strain gauge output. The question of hysteresis of the flooding point has not been addressed.

7.2 RESULTS

The estimated flow number at the flooding point is given in Tables 7.1 and 7.2 for the 1.83m and 2.67m tanks. Since relatively few data points were available, and because the probable error in the measured flooding point was relatively large, it was not possible to produce a flooding correlation. However, the data have been compared with the correlation of Nienow et al (1985):

$$(Fl)_F = 30 \left(\frac{D}{T}\right)^{3.5} (Fr)_F \quad (7.1)$$

The fit of the correlation to the data is within 20% (Fig 7.1); the correlation tended to underestimate the large scale data, but applying an F-test to the data showed that the difference in fit between the linear regression line and the correlation line was not statistically significant. One would not expect the fit to be very much better than $\pm 20\%$ since there are likely to have been many differences of impeller and sparger geometry between the tests reported here and the experiments on which the correlation was based. Warmoeskerken (1986) reported relatively

minor effects of sparger type and spacing on the flooding point. Increasing the sparger spacing from the impeller caused the flow number at flooding to increase slightly. The sparge ring for these experiments was facing downwards, giving a relatively large sparger spacing. On this basis, the data from this study are taken as confirmation of the flooding correlation for scales up to 2.67m in diameter. This is an important extrapolation of the range of the correlation.

Bujalski (1987) produced a modified form of the flooding correlation, including a term to account for the blade thickness:

$$(Fl)_F = 13.3 \left(\frac{x_1}{D}\right)^{-0.472} \left(\frac{D}{T}\right)^{4.63} (Fr)_F \quad (7.2)$$

For the impeller used in these tests, the prediction of the modified correlation was within 10% of the prediction of the original correlation, but the tendency of the original correlation slightly to underpredict the large-scale data is exaggerated when Bujalski's correlation is used. The effect of impeller diameter is important and further work is recommended to establish the exponent of (D/T) more firmly. The disagreement between the two forms of the correlation becomes rather large for $D/T = 2$. According to Bujalski's equation, the original correlation underestimates the flow number at flooding by more than 40% for a typical (x_1/D) at given Fr . This should be investigated further.

7.3 EFFECT OF VISCOSITY

A few experiments were conducted in the 0.61m vessel with a solution of 1% Sodium carboxymethylcellulose (CMC), grade 7H3 SXF, manufactured by Hercules Ltd, France. The solution was biologically stabilised using Kathon CG, a commercial biocide manufactured by Rohm and Haas Ltd. The range of experiments carried out in this solution was limited to measurements of gassed power, but the opportunity was taken to make a

preliminary assessment of whether a flooding point for viscous liquids could be defined in the same way as for aqueous solutions. The apparent viscosity of this non-Newtonian system at 20.5°C was given by:

$$\mu_{APP} = 3.55 \dot{\gamma}^{-0.55} \quad (7.3)$$

The solution showed no obvious signs of visco-elasticity such as climbing up a rotating shaft, but no accurate measurements to detect elasticity were attempted due to the lack of availability of suitable equipment.

The gassed power measurements, shown as power number against flow number at constant agitator speed, are shown in Fig 7.2. The overall shape of the curve is typical of the results of workers such as Kipke (1978) who has studied gassed power demand in viscous liquids. The rapid fall is caused by the formation of stable cavities shaped like anvils (van't Riet, 1975). The anvil cavities do not change significantly in size and shape as gassing rate increases, and the result is that gassed power number is virtually independent of flow number at constant impeller speed. It can be seen from Fig 7.2 that the scatter of the power data about the mean tended to increase with increasing gassing rate, but no step change could be seen, even though very high gas flow rates were attained in the experiments.

It was possible to see changes in the nature of the dispersion as the gassing rate was increased, but again, no definite transition could be seen. As gassing rate increased, large spherical cap bubbles became increasingly predominant, and increased in size as gassing rate increased. The maximum diameter observed was of the order of 100mm, with a very high rise velocity indeed. Clearly such bubbles cannot contribute significantly to mass transfer in the vessel because of their short residence times and low surface area to volume ratios. Qualitative observations of bubbles in the area of the vessel wall showed that the liquid velocities in regions away from the

agitator decreased as gassing rate increased, suggesting a fall in the liquid pumping capacity of the agitator.

This set of experiments showed that, at extreme gas flow rates, a condition analagous to flooding in low viscosity liquids is present. The transition to this state, however, was gradual and no distinct flooding point could be defined.

7.4 FLOODING OF CONCAVE TURBINE

Study of the gassed power curves for the concave turbine (Figs 4.16 to 4.18 in octanol and Fig 4.25 in electrolyte) shows that at high flow numbers, upwards step changes can be seen, at least for the 0.61m and 2.67m vessels. Once again, there was some breakthrough of gas at approximately the same flow number as the step change. One hesitates to define this as a flooding point since there is so little information on cavity transitions for concave turbines, and there is no supporting evidence that the point at which the step change occurs is coincident with any kind of flow regime transition. There was an unmistakeable tendency for the concave turbine to give step changes at slightly higher flow numbers at constant impeller speed than the Rushton turbine. The estimated ranges for the flow number at the sudden increase in power are shown in Table 7.2, compared with the corresponding estimates for the Rushton turbine.

It must be pointed out that no such step changes were observed in the 1.83m tank, which casts a shadow of doubt on the validity of the tests in the 2.67m tank. However, the large scale data were obtained in two completely separate sets of experiments which produced fairly consistent results.

It would be wisest to conclude that the gas handling capacity of the concave turbine appears to be greater than the Rushton turbine, but since firm experimental evidence is lacking, the Rushton turbine flooding correlation is an adequate, if rather conservative, estimate of the flooding point for a concave turbine. Further experiments are required to confirm the

apparent trend shown by this data. In particular, it would be useful to use a flexible vane sensor or a velocity-measuring device in the impeller outlet stream in conjunction with visual observation.

7.5 FLOODING OF PITCHED BLADE TURBINES

The gassed power and hold-up measurements with the upwards-pumping 6PBT showed that, for the operating conditions originally tested, there was no indication of flooding. The flooding of upwards-pumping agitators has been studied by Bujalski (1987), who reported a sharp transition from gas rising almost vertically through the vessel to gas being swept into liquid re-circulation loops. Impeller speed was increased at constant gas flow rate.

It was decided to investigate flooding in this study using visual observation through the vessel wall. The gas flow rate was set and the impeller speed was increased in small increments from zero. Tap water was used as the test liquid to allow observation of the bulk flow patterns in the liquid.

At low impeller speed, the gas was clearly not dispersed by the agitator. As impeller speed increased, the plume of gas increased slightly in diameter, until the point was reached at which gas bubbles were clearly being swept considerable radial distances away from the impeller tip by liquid re-circulation. This was the flooding point defined by Bujalski. At this stage a problem arose. It was noticed that the flow pattern oscillated irregularly from "flooded" to "dispersed". The period of the oscillation was quite long (tens of seconds), and for short times the "flooded" flow pattern predominated in one part of the tank, and the "dispersed" flow pattern in another.

The oscillations were only noticed by careful observation over a period of at least a minute, but were quite definite.

The problem was essentially one of definition. The oscillations took place over a range of about 5 RPM at speeds of about 90 RPM, so a considerable error could be introduced by inconsistent application of the criterion for flooding. It was decided to define the flooding point as the speed for which no oscillation to the "flooded" flow pattern was observed over a period of at least one minute.

The results of this set of experiments are given in Table 7.3. The data showed that:

$$N_F \propto Q_F^{0.25} \quad (7.4)$$

Bujalski gave the following equation for the flooding point:

$$(Fl)_F = 6000 (Fr)_F^{2.7} \left(\frac{T}{T_O}\right)^{1.55} \quad (7.5)$$

T_O = reference tank diameter = 1.0m

From this correlation, it may be seen that:

$$N_F \propto Q_F^{0.16} \quad (7.6)$$

This is rather weaker dependency than the result of this study. Bujalski used an impeller with a wider blade than used in this work, which may account for the differences in behaviour. There is also the rather crude nature of the technique used in these experiments, although the observations were very carefully made. Also, the absolute values of the speed at flooding are underestimated by Bujalski's correlation. Once again, the impeller blade width was probably the cause of the difference, but this result suggests that the flooding point is very sensitive to changes in impeller geometry.

To put this result in perspective, the power input per unit volume to avoid flooding was still well below normal operational limits.

Even at a gas flow rate of 2.8 VVM, the flooding point corresponded to an ungassed power input of 0.25 W/kg. The true gassed power input could not be measured at the time of the test, but is thought to have been around 90% of the ungassed value.

NOMENCLATURE : Chapter 8

SYMBOL	MEANING	UNITS
A_p	Annual production	tonne
C_A	Cost of agitator	£
C_E	Equivalent annual cost	£
C_p	Specific heat at constant pressure	J/kgK
C_T	Total present value	£
C_V	Specific heat at constant volume	JgK
d	Shaft diameter	m
d_p	Particle diameter	m
hP	Prime mover power	hP
i	Rate of return on capital (annual basis)	-
I_N	Net annual income	£
L	Length of agitator shaft	m
$\dot{m}$	Gas mass flow rate	kg/s
n	Mixer lifetime	years
N_B	Number of batches per year	-
P_G	Power required to compress gas	W
P_K	Prime mover power	kW
P_R	Present value, of running cost	£
p_1, p_2	Pressure at inlet and outlet to gas compressor	N/m ²
R	Running cost (annual basis)	£
r	Ratio of specific heats	-
t_B	Time required for one batch	hr
T_1, T_2	Temperature at inlet and outlet to gas compressor	K
U_c	Energy cost	£/kWh
U_T	Utilisation factor	hr/yr
V_C	Cost of vessel	£
V_D	Volume of dispersion	m ³
X	Ratio of weight of solids to weight of liquid	%
Z	Added value of product	£/te
σ_w	Working stress of shaft material	N/m ²
μ	Shaft bending moment	Nm
θ_m	Mixing time	s
$\Delta \rho$	Density difference between liquid and solid	kg/m ³

Subscripts

JS Just suspended (for solid suspension)

8. ECONOMICS OF MIXER DESIGN

A considerable amount of research published in the mixing literature has been aimed at identifying the most energy efficient agitator to produce the desired process result. There has been considerable controversy about the practical value of this work on economic grounds. It is now well established that for blending of miscible liquids and suspension of solids, that the larger the impeller diameter in relation to the tank diameter, the lower the energy requirement of the mixer. The drawback to the larger agitators is that they produce higher shaft torque than the smaller impellers, with the result that capital cost is increased. There has been little published work on the objective economic comparison of the various agitation options available to the designer. The basis of such a comparison is set out in this chapter after discussion of the technical background.

8.1 BLENDING

The task of blending miscible liquids is the most convenient to consider as the basis for an economic analysis. There is substantial evidence in the literature (Khang and Levenspiel, 1976; and Ruszkowski and Musckett, 1985) on the relationship between blend time, θ_m , and the impeller diameter, for low viscosity liquids:

$$(N\theta_m) (Po)^{1/3} \left(\frac{D}{T}\right)^a = b \quad (8.1)$$

a and b are experimentally determined quantities. b is especially dependent on the measuring technique and the criterion for complete mixing used in the experiments. The exact numerical value of b is, however, unimportant for this analysis.

If $a = \frac{5}{3}$ and since $Po = \frac{P}{\rho N^3 D^5}$, Equation (8.1) reduces to (assuming that Po is independent of D/T):

$$\frac{\theta_m P^{1/3}}{\rho^{1/3} T^{5/3}} = b \quad (8.2)$$

In other words, the energy required at given tank size to achieve the desired θ_m would be independent of impeller diameter. In practice, the experimental data show that $a > 5/3$, although the exact value depends on the agitator type being used for blending. This means that for given tank size, and taking a value of 2.17 for a (from FMP data):

$$P \propto \frac{D}{\theta_m^3}^{-1.5} \quad (8.3)$$

The improvement in energy efficiency is thought to be due to the improved distribution of energy dissipated by the agitator. Okamoto, Nishikawa and Hashimoto (1981) showed that the energy dissipation rates within a stirred vessel are extremely inhomogeneous, with much of the energy dissipated as turbulence in the impeller region. Improving the distribution of energy dissipation probably increases mixing rates in the regions away from the agitator, which are rate controlling for overall blending.

In fact, the same trends are shown in commercial design methods, such as that of Chemineer (Hicks, Morton and Fenic, 1976). The criterion for process result is the Chemscale, and tables are provided to establish the prime mover power and agitator speed combination to achieve a given Chemscale. Table 8.1 is an example of the mixer options available for a 30,000 gallon vessel with a Chemscale of 4. The motor power is selected from a fixed range of options, and the speed is determined by the gearbox option. Table 8.1 clearly shows that the power required to achieve a given process result falls quite sharply as impeller diameter increases. This result is supported by Chemineer's data on the liquid flow rate produced by the agitator, which for increasing impeller diameter, increases at constant power input. Thus mean circulation time is reduced, reducing blend time.

The similarity between the Chemineer design criteria for blending and fluid motion and the mixing time research implies that, whilst mixing time as such is seldom used in industrial design, it is a sufficient indicator of process result for relatively straightforward applications to be useful in making an economic analysis. Once complex chemical reaction schemes are encountered, equal mixing times is no longer a sufficient criterion for equal process result.

8.2 SOLIDS SUSPENSION

A great deal of work has been published on the suspension of solids in stirred tanks, relating the minimum speed for suspension to impeller geometry and properties of the solids.

Zwietering (1958) proposed a correlation for the minimum speed requirement to just suspend solids, N_{JS} , which has been recommended by Nienow (1985), after an extensive review of the literature in the following form:

$$N_{JS} = \frac{S v^{0.1} d_p^{0.2}}{D^{0.85}} \left(\frac{g \Delta \rho}{\rho} \right)^{0.45} X^{0.13} \quad (8.4)$$

The parameter, S , is sensitive to impeller size and clearance. As an example, Nienow recommended that:

$$S \propto (T/D)^{1.3} \quad (8.5)$$

At constant tank size, this leads to:

$$P_{JS} \propto D^{-1.45} \quad (8.6)$$

This is a very similar relationship to that derived for blending applications (Equation 8.3). Provided one remains strictly within the limits of the suspension correlations, the treatment for blending processes will apply to solids suspension, although a good deal of care must be taken in carrying out the solids

suspension analysis since the process is inherently more sensitive than blending to minor geometric changes.

The minimum speed for suspension has been shown to be the most economic operating condition for processes depending on mass transfer from solid to liquid or vice versa. Thus, the quantity studied in research is directly related to process result, which is not the case for blending.

8.3 GAS-LIQUID MIXING

It has already been shown in this thesis that for gas-liquid dispersion, the energy efficiency of the process is not related to D/T. However, there are performance criteria available from the current work and from the literature which allow the extension of economic analysis to gas liquid mixing.

Consider the performance of a hypothetical process which is mass transfer limited, for example, a hydrogenation. The reaction is carried out batch-wise, and calculations show that the rate of production will be determined by the hydrogenation step. In other words, the rate of production is determined by $k_L a$, assuming constant reactor pressure.

It is well established that:

$$k_L a = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (8.7)$$

Assuming that the power input per unit volume is fixed gives:

$$k_L a = A' v_{SG}^\gamma \quad (8.8)$$

A reasonable value for γ from this project and from van't Riet (1979) would be 0.5. The maximum superficial gas velocity is limited by the flooding point of the agitator, and the work of Pollard (1978) has shown that the most energy efficient operating point is close to the flooding point. The assumption

is that flooding is an undesirable condition due to poor gas distribution.

The flooding point of the agitator for a disc turbine is predicted by:

$$Fl_F = 30 \left(\frac{D}{T}\right)^{3.5} Fr_F \quad (8.9)$$

In terms of superficial gas velocity:

$$v_{SG} = 38.2 \left(\frac{D}{T}\right)^{5.5} \frac{N^3 D^2}{g} \quad (8.10)$$

Since power input per unit volume is to be kept constant, assuming that P_g/P_u is approximately constant:

$$N^3 = \left(\frac{P_u}{P_g}\right) \frac{P}{\rho P_o D^5} = \frac{A}{D^5} \quad (8.11)$$

Substituting for N^3 in Equation (8.10) leads to:

$$v_{SG} = \frac{38.2 A}{D^3 g} \left(\frac{D}{T}\right)^{5.5} \quad (8.12)$$

Expressing v_{SG} in terms of impeller diameter:

$$v_{SG} = A' \frac{D^{2.5}}{T^{5.5}} \quad (8.13)$$

Increasing v_{SG} also has the additional benefit for non-pure gases that the concentration driving force increases since depletion will be reduced. This does not apply to hydrogenations.

The outcome of this exercise is that plant production is increased at the expense of higher running costs (increased gas compression costs) and of higher capital costs. The balance is now between higher capital costs and running costs and higher income as a result of higher production. The economic analysis becomes rather more complex than for blending and much more

difficult to apply since factors such as added value of the product must be considered.

8.4 ECONOMIC BACKGROUND

The cases of blending and solids suspension provide a fairly simple economic problem. The outgoings resulting from running the process are composed of a single payment to cover the capital cost of the agitator, followed by a uniform series of payments to cover running costs.

One method of comparing the trade-off between running costs and capital costs is the present value method. This is used in the paper of Meyer and Kime (1976), and involves discounting payments of running costs in future years back to the present. A uniform rate of return on investment is assumed, such that the value of £Y after k years is given by:

$$Y_k = Y (1 + i)^k \quad (8.14)$$

Conversely, if the annual running costs are £R, the amount of money which would have to be invested now to realise £R after k years would be:

$$P_R = \frac{R}{(1 + i)^k} \quad (8.15)$$

Summing the discounted running costs over the total mixer lifetime of n years gives an expression for the present value of the capital and running costs of the agitator:

$$C_T = C_A + \sum_{k=1}^n \frac{R}{(1 + i)^k} \quad (8.16)$$

Summing the geometric series gives:

$$C_T = C_A + \frac{R (1+i)^n - R}{i (1+i)^n} \quad (8.17)$$

Inflation is not explicitly accounted for in the treatment, largely because of the difficulty of predicting inflationary trends over a period of ten to fifteen years. The rate of return on investment, i , will reflect inflationary trends at the time of making the comparison. Another factor which could be utilised is the increase in annual maintenance costs with increasing age. Maintenance costs are not likely to be a function of agitator diameter and do not therefore influence the comparison of agitators. Also, maintenance costs are also likely to be significant only at the end of the life-time of the mixer, so the present value will be relatively small.

The alternative to the present value approach is the equivalent annual cost method. Instead of converting running costs to a single current value, the capital cost is spread over the mixer lifetime to give an annual cost which is the sum of the annual running costs and the equivalent annual cost of the capital. The capital is effectively treated as a loan which is paid back in annual installments over the mixer life-time, using an appropriate rate of return on investment. The equivalent annual cost is therefore given by:

$$C_E = R + \frac{Ci(1+i)^n}{(1+i)^n - 1} \quad (8.18)$$

This provides a quantity which is more meaningful to some people than the present value method, but the two methods are clearly equivalent.

For simple cases like blending and solid suspension, the concept of payback may be applied. This is a frequently used criterion in engineering economic analysis which considers the time taken to repay a capital investment in terms of increased profits. The weakness of the pay-back criterion is that profits after the pay-back period are ignored. This is irrelevant for projects which involve only uniform series, such as the analysis of mixers. Pay-back is most useful in considering modifications to existing plant to improve operating profitability, and will be

used in this analysis to evaluate the feasibility of retrofitting.

8.5 COST ESTIMATION

The cost of a mixer is largely determined by the torque on the impeller shaft as seen in the shaft design method developed by Pollard (1983), which is similar in concept to that of Chemineer (Ramsey and Zoller, 1976). The shaft diameter is calculated from:

$$d^3 = \frac{16}{\pi \sigma_w} (\tau^2 + \mu^2)^{1/2} \quad (8.19)$$

$$\mu = \text{shaft bending moment} = \lambda \frac{L}{D} \tau$$

λ = empirical constant

Shaft diameter is thus, broadly speaking, a function of torque only. In fact, λ depends on the ratio of blade-passing frequency to natural frequency, but for most practical applications λ may be taken as being constant. The shaft diameter determines the shaft seal size and the gearbox size. The impeller cost, naturally, increases with increasing torque (the agitator becomes larger), as does the cost of the shaft itself. Table 8.2 gives some information on costs of the individual units of the drive train. The only item which falls in cost with increasing torque is the prime mover, since power required is actually falling.

Once again, valuable information on costs is provided by Chemineer (Mayer and Kime, 1976) which is illustrated graphically in Fig 8.1. Middleton (1978) produced an equation based on the Chemineer graph, which appeared to be adequate in describing the capital cost for motors larger than 7.5kW. A copy of the original paper by Meyer and Kime as submitted to Chemical Engineering has kindly been provided by DuPont Chemical Company, Wilmington, Delaware, USA which contains the original

data table from which Fig 8.1 was produced. Correlating the data using non-linear regression shows that, for 1976 dollar prices:

$$C = \$24000 \frac{(hP)^{0.7}}{(rpm)^{0.7}} \quad (8.20)$$

This can be updated using the Process Engineering capital cost index and then converted to sterling at 1987 exchange rates, (£1 = \$1.60):

$$C = £2200 \frac{P_K^{0.7}}{N^{0.7}} \quad (8.21)$$

The original data are correlated to $\pm 20\%$, which is sufficiently accurate for the purposes of this analysis.

The tendency for capital cost to increase with increasing torque is clearly seen from equation (8.21).

The relationship is only valid for single agitator systems in 316 stainless steel. The costs for dual agitator systems are from 20% to 50% higher, whilst the cost for single agitator, carbon steel systems are from 15% to 60% lower than the stainless steel costs. To simplify the analysis, only single impeller mixers with stainless steel wetted parts will be considered.

A factor which is neglected when using an overall capital cost relationship is the potential mechanical unsuitability of a design on the grounds that the agitator as designed would be running too close to its critical speed. On the whole, this will tend to favour large, low speed agitators, but the influence of mechanical unsuitability is difficult to account for explicitly in an economic analysis. Mechanical aspects of the mixer design have therefore been neglected; it should be remembered that this omission will tend to favour the small diameter, high speed agitator option.

The cost of support structure should also be considered, since the weight of the mixer and the forces transmitted to the supports depend on the shaft torque. For the example quoted in Table 8.1, the weight of the agitator would be 4100 lbs for the lightest option, and 4500 lbs for the heaviest. The forces transmitted for the lowest speed option would be about 15% higher than those for the highest speed option. For practical purposes, the additional cost of supporting the larger diameter impeller can be neglected, especially since the cost will be very closely related to the method of support.

There are practical constraints on the maximum size of the agitator which can be used in a given vessel. The large diameter, low-energy impellers will probably be limited by the size of the opening in a closed vessel through which the agitator must pass. Often, however, impellers may be bolted or even welded in place in the vessel, or the tank may be open-topped, in which case this consideration becomes insignificant. A worst case might be that the vessel must be constructed in two pieces to accommodate the larger agitator, which will increase the capital cost very significantly. The application of this analysis to practical process engineering is clearly to be done on a case-by-case basis since there are too many practical considerations to take into account in a general treatment.

Finally running costs may be approximated by considering only the cost of electrical energy. The question of maintenance costs has already been discussed in an earlier section. The annual running cost is given by:

$$R = U_T U_C P_K \quad (8.22)$$

The key parameter is the utilisation factor, U_T , which is the number of hours per year for which the mixer is operated. The influence of utilisation on total agitation cost will be examined in a later section.

8.6 ECONOMIC ASSESSMENT OF BLENDING: fixed vessel size

The first case to be considered will be that of a tank of fixed size, which must achieve a specified blend time, θ_m . The impeller diameter is to be allowed to vary, and the optimum impeller size on an overall economic basis is to be assessed.

Using equations (8.1) and the form of equation (8.21) proposed by Middleton (1978), the following expression for capital costs may be derived (see Appendix 8A):

$$C_A = \frac{\pounds 14600 T^{2.82} D^{0.68}}{\theta_m^{1.3}} \quad (8.23)$$

Using equations (8.3), (8.22), and the definition of power number gives the following expression for annual running costs:

$$R = \frac{\pounds 54.7}{\theta_m^3} \frac{T^{6.5}}{D^{1.5}} U_T U_c \quad (8.24)$$

In deriving equations (8.23) and (8.24) the power number has been taken as 1.8 (a pitched blade turbine), and the fluid density has been taken as 1000 kg/m^3 . The prime mover power has been taken as $1.1 \times$ (power transmitted to fluid).

All that remains now is to pick a tank diameter and a realistic estimate for mixing time, ensuring that the prime mover power exceeds 7.5 kW. The tank diameter will be taken as 2.7 m with a mixing time of 12 seconds.

Equation (8.23) becomes:

$$C_A = \pounds 9500 D^{0.68} \quad (8.25)$$

Equation (8.24) becomes

$$R = \frac{\pounds 20.15}{D^{1.5}} U_T U_c \quad (8.26)$$

Using the present value method, the total present value of the mixer as a function of impeller diameter is shown in Table 8.3. The mixer lifetime has been taken as 10 years with a 10% expected rate of return on capital per annum. The utilisation factor has been taken as 5000 hours per year. The results show a steady fall in total cost with increasing impeller size. Strictly speaking, the points at $D/T = 0.2$ and $D/T = 0.6$ should have been omitted, being outside the range of the correlation obtained from equation (8.1); they have been included to illustrate that often, no minimum in the total cost is found. The larger the impeller, the lower the total cost becomes.

It is instructive to investigate the effect of the assumptions about the rate of return on investment and the utilisation factor. Figure 8.2 shows the result of varying the utilisation factor. The plots show that the utilisation must become very low for the high speed, low D/T mixer design to be the most economic.

Fig 8.3 shows the influence of the rate of return. There is no impact on the overall conclusion for this case, which is that larger impellers are more economic than small ones.

Figures 8.4 and 8.5 show the influence of estimated mixer lifetime and mixing time respectively. The estimated mixer lifetime affects the present worth of the running costs, and would have to be very short to make the larger D/T impeller uneconomic for this example. Mixing time is a more significant parameter, since running costs are inversely proportional to θ_m^3 whilst capital costs are inversely proportional to $\theta_m^{1.3}$. Fig 8.5 shows the importance of the degree of agitation in determining relative costs. At very long mixing times, the low capital cost option becomes the most economic.

Table 8.4 gives the same assessment as Table 8.3, except on an equivalent annual cost basis. It would be useful to convert these figures to a cost per unit of production, but too many

assumptions would need to be made about other processing and raw material costs. It is clear that savings can be substantial, and if any payback time criterion is applied, as is common in the process industry, the repayment period for the higher capital investment option will be months rather than years for this example.

8.7 ECONOMIC ASSESSMENT OF BLENDING: optimising vessel size

This analysis can be extended to include the effects of vessel volume, when there is the freedom to select the tank size as well as the impeller size. Let us consider the hypothetical case of a continuous mixer whose duty is to neutralise an acidic waste stream, which has a fixed flow rate (Fig 8.6). There are no major constraints on residence time, and in designing the mixer, the mixing time required will be set to one quarter of the residence time, which should be sufficient to allow the neutralisation reaction to proceed to completion.

Expressions for capital and running costs have already been derived, and an approximate expression for vessel volume is available (Middleton, 1978):

$$V_C = \alpha V^{0.7} \quad (8.27)$$

Adding this cost to the capital and running costs to give an equation for total cost:

$$C_T = \alpha_1 T^{2.1} + \frac{\beta}{T^{0.4}} \left(\frac{D}{T}\right)^{0.68} + \frac{\gamma}{T^4} \left(\frac{T}{D}\right)^{1.5} \quad (8.28)$$

$$\alpha_1 = \alpha \left(\frac{\pi}{4}\right)^{0.7}$$

The constant α is dependent on the vessel materials of construction and geometry (open or closed, one or two piece, etc). β and γ depend strongly on the flow rate of the waste stream, and γ also depends on the unit power cost, utilisation factor and rate of return on investment (see Appendix 8A).

By using the following set of values, equation (8.28) has been evaluated as a function of tank diameter and D/T (Fig 8.7).

$$\begin{aligned}\alpha &= \text{£}2800 \\ Q &= 0.25 \text{ m}^3/\text{s} \\ U_T &= 5000 \text{ hours/year} \\ i &= 10\% \\ U_c &= 0.05 \text{ £/kWh}\end{aligned}$$

The plot shows that the larger diameter impellers are still more economic than the smaller diameter agitators, and there is also a small effect of (D/T) on the optimum vessel volume. A larger (D/T) produces an optimum vessel of slightly less volume than a small (D/T), which may provide control advantages as well as cost advantages.

This example is, of course, idealised since it is rare for this type of freedom of choice to appear in practice, but it does show that the benefits of using larger diameter, lower energy impellers are not necessarily swamped by other process and economic considerations, and in this example the potential savings are considerable, with a 20% difference in total cost between the optimum solutions at D/T = 0.5 and D/T = 0.3.

Overall, the low energy input options are clearly worthy of consideration on economic grounds. It would be worthwhile to extend this analysis to different impeller types performing the same duty. For instance, one could compare a propeller with a pitched blade turbine for blending, and investigate the potential cost advantages of low torque agitators.

8.8 GAS LIQUID MIXING

In applying the concepts developed in Section 8.3 to gas-liquid mixing, the costs and income associated with altering the impeller geometry must be clearly identified.

8.8.1 Gas Compression Cost

There are two ways of assessing the cost of supplying additional gas to the agitator. If sufficient capacity is available from existing supply, no additional capital expenditure is required, and the additional gas supply cost is approximately that of the extra electrical energy required to compress the gas.

The power required to compress the gas is given by:-

$$P_G = \dot{m} C_p (T_2 - T_1) \quad (8.29)$$

If the compression process is isentropic:

$$\frac{T_2}{T_1} = \left(\frac{p_2}{p_1}\right)^{\frac{r-1}{r}} \quad (8.30)$$

r = ratio of specific heats of gas = C_p/C_v

Allowing for the non-ideality of the compressor, the actual power requirement is given by:

$$P_G = \frac{\dot{m} C_p T_1}{\eta} \left[\left(\frac{p_2}{p_1}\right)^{\frac{r-1}{r}} - 1 \right] \quad (8.31)$$

η = isentropic efficiency of compressor

If the vessel is supplied by a dedicated compressor, increasing the gas flow rate may well increase the capital cost since a larger compressor may need to be purchased. There may also be a requirement for larger diameter pipework to reduce friction losses and to ease the required delivery pressure from the compressor, which has a direct bearing on running costs. Accounting for these factors could prove to be extremely complex, since the costs will be very dependent on the type of compressor used, which will be a function of the operating conditions.

Equation (8.31) should be rewritten in terms of superficial gas velocity to use in a cost predictor equation:

$$\dot{m} = \rho_g \frac{\pi T^2}{4} v_{SG} \quad (8.32)$$

$$\therefore P_G = \frac{\pi T^2}{4} \frac{\rho_g C_p T_1}{\eta} \left[\left(\frac{P_2}{P_1} \right)^{\frac{r-1}{r}} - 1 \right] v_{SG} \quad (8.33)$$

8.8.2 Capital Cost

The capital cost of the agitator will increase as diameter increases since the torque increases. Capital cost is given by:

$$C_A = £2200 \frac{P_K^{0.7}}{N^{0.7}} \quad (8.34)$$

If P_g/P_u is approximately constant, then at constant power input:

$$N^3 = \frac{1000 P_K}{\rho_L P_o D^5} \left(\frac{P_u}{P_g} \right) \quad (8.35)$$

The capital cost may be expressed as:

$$C_A = £450 P_K^{0.47} \rho_L^{0.23} P_o^{0.23} D^{1.15} \left(\frac{P_g}{P_u} \right)^{0.23} \quad (8.36)$$

Increasing the gassing rate also increases the gas hold-up since:

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma \quad (8.37)$$

The volume of the dispersion, V_D , which will be assumed to be equal to the tank volume must therefore be increased with increasing gassing rate since:

$$V_D = \frac{V_L}{1 - \epsilon_G} \quad (8.38)$$

For simplicity, it will be assumed that the vessel volume will be increased by increasing the height of the vessel, not by increasing the diameter, which would produce complex

interactions with other equations. The vessel volume is thus given by:

$$V = \frac{V_L}{1 - \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma} \quad (8.39)$$

The cost of a stainless steel vessel, operating at 7 bar is given by Middleton (1978) as:

$$V_C = £2800 V^{0.7} \quad (8.40)$$

$$\text{or } V_C = £2800 \left[\frac{V_L}{1 - \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma} \right]^{0.7} \quad (8.41)$$

Note that this has been updated to 1987 prices.

8.8.3 Production

The variable costs associated with production which will be considered are as follows:

1. Raw material costs. Each tonne of finished product consumes a fixed quantity of raw material with a fixed unit cost.
2. Processing costs. There are downstream costs such as separation, drying, packaging, storage etc which will be directly related to the rate of production. Also, energy input apart from the mixer energy input must be considered. If the reaction is carried out at elevated temperature, the feedstock must be heated to start the reaction which probably requires far more energy than the mixer energy input.

Another aspect to consider is the effect of increasing production on product price in the market place. Increasing production may cause the price to fall, or the market may already be saturated and the surplus product may not be sold.

Such commercial considerations are too complex to include in this simple model. It is proposed that the net income generated by manufacture and sale of the product be described as the product of the production rate and the added value of the product per tonne:

$$I = A_p Z \quad (8.42)$$

Z = added value per tonne
 = difference between selling price and variable running costs (raw materials, energy, labour etc)

The selling price will be assumed to be independent of production rate.

For a batch production process, the rate of production is given by:

$$A_p = N_B V_L \quad (8.43)$$

If it is assumed that the process is entirely mass transfer-controlled, then the batch time, t_B , is related to $k_L a$:

$$t_B = \frac{A}{k_L a} \quad (8.44)$$

A is a constant of proportionality and is very system dependent.

The vessel will be in operation for a fixed number of hours per year, U_T , which enables one to express the production rate in terms of $k_L a$:

$$N_B = \frac{U_T}{t_B} \quad (8.45)$$

$$\therefore A_p = \frac{U_T k_L a V_L}{A} \quad (8.46)$$

Since $k_L a \propto v_{SG}^{0.5}$, this may be re-written as:

$$A_p = A' U_T V_L v_{SG}^{0.5} \quad (8.47)$$

8.8.4 Mixer Running Cost

Since it has been assumed that power input per unit liquid volume is fixed, the mixer running cost is also fixed:

$$R = U_T U_C P_K \quad (8.48)$$

8.8.5 Cost Equation

It is simplest to use the equivalent annual cost method to account for capital costs.

The annual net income from the project on this basis may be expressed as:-

$$I_N = I - R - C_G - f (V_C + C_A) \quad (8.49)$$

$$\begin{aligned} I &= A_p Z && = \text{net income from sales} \\ R &= U_T U_C P_K && = \text{annual cost of electrical power} \\ C_G &= U_T U_C P_G && = \text{annual gas compression cost} \\ f &= \text{equivalent annual cost factor} = \frac{i (1 + i)^n}{(1 + i)^n - 1} \end{aligned}$$

All the terms in this equation may be expressed in terms of superficial gas velocity, giving a complex function, remembering that there is a direct relationship between D and v_{SG} , from the flooding correlation using constant power input per unit volume.

$$v_{SG} = \frac{38.2 P}{\rho_L g P_o T^{5.5}} D^{2.5} \left(\frac{P_u}{P_g} \right) \quad (8.50)$$

The final cost equation is:

$$\begin{aligned}
 I_N = & A' U_T V_L Z v_{SG}^{0.5} - P_K U_C U_T - \\
 & \frac{\pi T^2}{4} \frac{P_G C_P T_1}{\eta} \left[\left(\frac{P_2}{P_1} \right)^{\frac{\gamma-1}{\gamma}} - 1 \right] U_C U_T v_{SG} - \\
 & f \left[2800 \left(\frac{V_L}{1 - \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma} \right)^{0.7} + \right. \\
 & \left. 3.51 \left(\frac{P}{P_u} \right)^{0.7} \rho_L^{0.7} P_o^{0.7} g^{0.46} T^{2.53} v_{SG}^{0.46} \right] \quad (8.51)
 \end{aligned}$$

The form of this equation is such that a clear trade-off exists between increased production and increased capital and running costs as a result of increasing production. The question of whether the net income will be increased by increasing production can be resolved only by referring to a specific example. Clearly, the major factor influencing the outcome is the added value of the product.

8.8.6 Major Assumptions

Before looking at a specific example, it would be wise to consider the limitations of the analysis:

1. The capacity of the plant is not infinitely expandable, some other processing step is likely to become limiting, and only small increases in production are likely to be achievable in practice.
2. The capital costs do not reflect the need for increased heat transfer within the vessel. As the reaction rate increases, the rate of heat evolution increases, which requires a higher specific heat transfer rate if the temperature of the batch is to remain constant. The heat transfer area for cooling would therefore need to be

increased, which influences capital cost, and could eventually become the limiting factor.

3. The chemistry of the process is assumed to be simple, so that the product yield and quality is not a function of batch time.

8.8.7 Worked Example

For this example, the process will be a hydrogenation, operating at 7 bar pressure. For a $k_L a$ value of 0.1 s^{-1} under virtually flooded conditions, it is known that it is possible to make 500 batches per year, and that the gas hold-up under these conditions is 10%. The agitator diameter on which these calculations were based was $T/3$.

Rate of Production

Batch volume	=	10 m ³
Operating time/year	=	6000 hours
Superficial gas velocity for D	=	$T/3 = 0.072 \text{ m/s}$
Number of batches/year	=	500

Using this information, the constant of proportionality, A' , in Equation (8.47) = 0.31.

$$A_p = 18633.9 v_{SG}^{0.5} \text{ [tonne/year]} \quad (8.52)$$

Gas Compression Cost

The gas is hydrogen, which will be compressed from atmospheric pressure to 7 bar. This is a rather unrealistic scenario, but will overestimate the gas compression costs, tending to cancel out the possible additional capital expenditure required to produce the higher gas flow rates.

Specific heat of hydrogen = 14700 J/kg K

Density of hydrogen at sparger conditions

(reaction temperature = 100°C) = 0.45 kg/m³

U_T = 6000 hrs/year

Energy cost = £0.05/kWh

η = efficiency of compression = 0.6

$$C_G = £2.1 \times 10^6 v_{SG} \quad (8.53)$$

Capital Cost

Using the following values, the mixer and vessel capital cost equations can be obtained:

V_L = 10 m³

(P_g/P_u) = 0.5

ρ_L = 1000 kg/m³

P_o = 5.5

g = 9.81 m/s²

T = 2.33 m

$$C_A = £2.2 \times 10^4 v_{SG}^{0.46} \quad (8.54)$$

$$C_V = £ \frac{14 \times 10^3}{(1 - 0.48 v_{SG}^{0.6})^{0.7}} \quad (8.55)$$

The equivalent capital cost factor, for $i = 0.1$ and a lifetime of 10 years, is 0.16.

Mixer Running Cost

For a power demand of 10 kW, the annual mixer running cost is £3000.

Final Equation

$$I_N = 1.9 \times 10^4 v_{SG}^{0.5} Z - 2.1 \times 10^6 v_{SG} - 3000 \\ - 3500 v_{SG}^{0.46} - \frac{2240}{(1 - 0.48 v_{SG}^{0.6})^{0.7}} \quad (8.56)$$

For this particular example, the mixer running cost and capital costs can be neglected. Any increase in production increases the net annual income for all except very low values of Z . The net income is positive whilst:

$$Z > 10^2 v_{SG}^{0.5} \quad (8.57)$$

There also exists an optimum operating superficial velocity at which the net income is a maximum. This is obtained by differentiating the cost equation neglecting the mixer costs:

$$v_{SG} = 2.5 \times 10^{-5} Z^2 \quad (8.58)$$

For all except low values of Z , this optimum value of v_{SG} is well beyond that normally found in practice. The conclusion is that almost any increase in throughput produces economic benefits. This is not surprising, since if it were otherwise, the economics of running the process at all would be marginal. The example is rather unrealistic, since in a real process plant other factors would limit production but it does show how a knowledge of gas-liquid dispersion processes could be used to improve the economics of a manufacturing process.

8.9 The Economics of Retro-Fitting

The arguments developed in the previous section can be applied to retro-fitting, with the same assumptions and qualifications. In the case of retro-fitting a vessel with an impeller designed to draw higher gassed power than the original,

with the same motor and gearbox, the improvement in $k_L a$ can be expressed, at constant gassing rate, at:

$$k_L a = A (\bar{\epsilon}_m)^{0.3} \quad (8.59)$$

There is an additional running cost, since electrical energy consumption is increased, and there is a capital expenditure to replace the original impeller. It is assumed that the retro-fit takes place during a planned shut-down so that no loss of production occurs.

Using a modified version of Equation (8.55) the additional income generated per year by the retro-fit is:

$$\Delta I_n = (\Delta A_p) Z - U_T U_C (\Delta P_K) - f C_I \quad (8.60)$$

C_I = capital cost of the agitator and the replacement exercise

ΔP_K = change in energy consumption as a result of retro-fit (kW)

ΔA_p = change in annual production as result of retro-fit

Once again, the added value per tonne of the product is the crucial variable. Referring to the example of section 8.8, with a cost of £1000 for the new impeller and a labour cost of £1000 to replace the old agitator, and assuming that production is increased by 5% by the retro-fit (250 tonne/year), the additional income becomes

$$\Delta I_n = 250 Z - 300 \Delta P_K - 320 \quad (8.61)$$

This is positive for

$$Z > 1.2 \Delta P_K + 1.3 \quad (8.62)$$

Since $\Delta P_K \approx 10$ kW at the highest for this example, the retro-fit generates additional income for all except very low values of Z .

The retro-fitting process can also be analysed in terms of pay-back time. Neglecting the discounting of future expenditure to the present, the pay-back time, T_p , is given by:

$$T_p = \frac{fC_I}{(\Delta A_p Z - U_T U_C \Delta P_K)} \quad (8.63)$$

The pay-back time for $Z = £50/\text{tonne}$ is of the order of two weeks for this example.

8.10 CONCLUSIONS

This economic analysis has demonstrated the potential value of retro-fitting and the application of academic research into energy efficiency to industrial situations. The examples discussed in this chapter, although rather artificial, have all shown that energy efficiency is economically highly desirable, as are measures taken to increase throughput. It is recognised that reality is very different from the cases discussed here, but the findings of this analysis show that, at least from a theoretical stand-point, academic research can lead to major improvements in profitability.

9 DISCUSSION

The object of this chapter is to set the experimental data in the context of scale-up rules, and to compare the data with published literature data, where this has not already been done.

9.1 GASED POWER : RUSHTON TURBINE

Many correlations have been produced to predict the gassed power consumption of Rushton turbines in air-water systems. The variety of the types of correlation available probably reflects the lack of coherence between the work of different authors.

The most recent correlation to appear in the literature was that of Warmoeskerken (1986) who attempted to correlate gassed power curves by linearisation. The resulting equations were :

$$Fl > 0.1 ; \quad \frac{P_g}{P_u} = 0.27 + \frac{0.022}{Fr} \quad (9.1)$$

$$Fl < Fl_C ; \quad \frac{P_g}{P_u} = 1 - 16.7 Fl Fr^{0.35} \quad (9.2)$$

$$Fl_C < Fl < 0.1 ; \quad \frac{P_g}{P_u} = B - 0.1 \frac{A - B}{Fl_C - 0.1} + \frac{(A - B) Fl}{Fl_C - 0.1} \quad (9.3)$$

$$A = 1 - 17 Fl_C Fr^{0.35} \quad (9.4)$$

$$B = 0.27 + \frac{0.022}{Fr} \quad (9.5)$$

$$Fl_C = 3.8 \times 10^{-3} \left(\frac{Re}{Fr} \right)^{0.067} \left(\frac{T}{D} \right)^{0.5} \quad (9.6)$$

Warmoeskerken pointed out that these equations were fairly crude approximations to the data, but it is clear that Froude number was a controlling factor on P_g/P_u for Warmoeskerken's data.

Furthermore, at high flow numbers, P_g/P_u is related to $1/Fr$ which is the tendency shown by all the data presented here. The application of these correlations to the current data is shown in Fig 9.1 at constant Froude number for the 0.61m and 2.67m tanks. There are clear differences between the predictions of the correlation and the large scale data, especially at low flow

numbers. It should be emphasised that this is not an accurate comparison with Warmoeskerken's data since there are many factors, such as D/T , impeller clearance and liquid type, which complicate the interpretation of the power curves, especially at low flow numbers. This uncertainty is caused by the effect of gas re-circulation, which is geometry dependent.

The predictions of Warmoeskerken's correlation for a Froude number of 0.2 are shown in Fig 9.2. The predicted trend is the opposite of the experimental results of this study when plotted at constant Froude number at different scales (Fig 4.9). The correlation is, of course, a simplification, but the scale effect predicted by the correlation is for the power curves to become steeper with increasing scale.

There is a clear scale effect in Fig 4.9 which does not appear in Warmoeskerken's correlations. At low flow numbers, the gassed power ratio is higher at large scale than at small scale although the Froude number was constant. The data of Davies (1986), show roughly the same effect, although the data cannot be plotted at constant Froude number. The only current explanation for this phenomenon is that re-circulation becomes less significant at large scale. The re-circulation correlation of Nienow et al (1977), predicts a maximum flow number of 0.046 for re-circulation to take place at a Froude number of 0.2, so re-circulation effects should certainly be present even though the Froude number is rather low. The extent of re-circulation is not predicted by the correlation. Scale effects on the extent of re-circulation at low flow number could be the cause of the change of the initial slope of P_g/P_u as a function of flow number with scale.

Nienow et al also produced a plot of gassed power curves at scales ranging from $T = 0.29\text{m}$ to $T = 1.83\text{m}$ (Fig 9.3). This was a composite set of curves with the condition that the flow number lay above that required for re-circulation and below that required for complete dispersion. The curves should thus be

free of re-circulation effects. A clear scale effect can be seen, with P_g/P_u tending to increase with increasing scale at constant flow number, especially at low flow numbers. Provided the prediction of the correlation for the onset of re-circulation was correct, this result implied that there is a scale effect independent of gas re-circulation on the gassed power demand.

Van't Riet (1975) also considered the effects of scale on gassed power curves, and predicted that the curves would show the same trends as Fig 9.3. The prediction was made on the basis of observations of cavity formation. Van't Riet observed large cavities to form one by one in sequence with increasing gassing rate, rather than the sudden transition to 3 large cavities described by Warmoeskerken and Smith (1982). This has been attributed to the method of observation, and Warmoeskerken and Smith maintained that van't Riet's original result was incorrect. Van't Riet took the gassed power ratio as being characteristic of the number of large cavities present (in the absence of re-circulation), and from observations of scale effects on the cavity structure rationalised the tendency for power curves to become flatter with increasing scale. This prediction fits the trends shown by the data of Nienow et al (1977), and the trends shown by the data reported here and, once again, scale effects on re-circulation have not been invoked to explain the phenomenon. The question is difficult to resolve because of the doubt cast over van't Riet's description of the sequence of cavity formation as a result of later work in the same laboratory. It seems clear that there are scale effects on gassed power curves which cannot be eliminated by plotting data from different scales at constant Froude number. Re-circulation effects are not the only cause of the tendency for gassed power ratio to increase with increasing scale at constant flow number, but an adequate explanation is currently lacking.

The various correlations for gassed power published in the literature have been compared to data obtained in water whilst

carrying out the 2.67m mass transfer experiments (Fig 9.4). This was to give the closest possible comparison with the operating conditions of other workers. It is immediately apparent that the correlations relating gassed power ratio to Weber number and flow number produce very poor predictions at large scale. This is probably because the data on which the correlations were based were taken at very small scale and the introduction of surface tension has forced the inclusion of the Weber number. Hughmark (1980) showed that the influence of surface tension on gassed power ratio is not statistically significant. The inclusion of Weber number is not therefore justified. Weber numbers tend to increase with scale, which is the cause of the very low predictions from the equations of Hassan and Robinson (1977) and Luong and Volesky (1979). Correlations involving Weber number for gassed power ratios are not recommended.

The best approximation to the high flow number data at all scales is the correlation of Michel and Miller (1962). The later modifications of Loiseau et al (1977) and Yung et al (1979) tend to overestimate the power demand, whilst the correlation of Nagata (1975) shows too great a dependency on gassing rate. The Michel-Miller correlation, for reasons which are not apparent from their paper, breaks down at even moderately low flow numbers, which would suggest that the original data was largely concentrated in the "3-3" regime. There are other features of the correlation which do not fit the trends evident in the data. At constant gas flow rate, the gassed power ratio decreased with increasing speed. The Michel-Miller correlation predicts the opposite trend, since by rearrangement of the correlation it may be shown that

$$P_g/P_u \propto N^{0.15} \quad (9.7)$$

Further manipulation of the equation shows that at constant Froude number and flow number,

$$P_g/P_u \propto D^{0.15} \quad (9.8)$$

Again, this prediction is not fulfilled by the experimental data (Fig 4.9 and 4.10).

If constant flow number and constant ungassed power input per unit volume are considered, then

$$P_g/P_u \propto D^{1/6} \quad (9.9)$$

The agreement between this prediction and the data shown in Fig 4.11 is probably coincidental. The choice of scale-up criterion has very little influence on the exponent on scale, which must be seen as a weakness in view of the experimental evidence to the contrary.

None of the published correlations for gassed power consumption are really satisfactory. Those involving the use of Weber number may be rejected because of spurious scale effects, whilst the fairly complex correlation of Warmoeskerken (1986) does not represent a significant advance over the correlation of Michel and Miller (1962) in terms of predictions for the experimental data of this project. There are too many effects, such as recirculation, which are adequately explained qualitatively, but do not provide quantitative information to allow the development of a true predictive equation, in spite of many papers being written on the subject of gassed power consumption in the relatively ideal model system of air and water.

9.2 GASED POWER: CONCAVE TURBINE

Apart from the work of van't Riet, Bruijn and Smith (1976), there is little published data on the gassed power consumption of the concave turbine. Any comparison should be made with great care since there were differences of scale and D/T between van't Riet et al and the results discussed here. Fig 9.5 shows the power curves obtained with $Fr = 0.2$ compared with van't Riet

et al's results at their minimum impeller speed, which corresponded to a Froude number of 0.36. The main feature of the curves is that at no time did P_g exceed P_u , in disagreement with van't Riet et al. No satisfactory explanation has been put forward for van't Riet's result. It does not appear to be affected by re-circulation as all van't Riet's curves coincided at low flow numbers. The 1.83m and 2.67m data produced lower values of P_g and P_u than van't Riet or the 0.61m data.

The discrepancies between the data reported here and van't Riet et al are in practical terms of little importance. Industrial equipment would not operate in the region of low flow number for which van't Riet et al found P_g to exceed P_u . The problems of reconciling power data and correlations for the Rushton turbine have already been discussed. Understanding of the fluid mechanics of the flow around Rushton turbine blades is relatively advanced, but no unified power correlation can be found. The concave turbine is much more poorly understood, so the detailed comparison of results between two sets of researchers using different geometries would appear to be fruitless. The important point is that the concave turbine gives flatter gassed power curves than the Rushton turbine. A worked example comparing the two agitators will be given in a subsequent section.

9.3 GASED POWER : UPWARDS-PUMPING PBT

The gassed power requirements of the downwards-pumping PBT have already been discussed in some detail in the chapter on power. Both Chapman, Nienow, Cooke and Middleton (1983) and Bujalski (1987) produced gassed power measurements for upwards-pumping PBT's of various geometries. Unfortunately, these are presented as lines of constant flow rate at varying impeller speed, rather than lines of constant speed at varying flow rate, which is the strategy adopted in this thesis. Bujalski, in a private communication, has very kindly made available some data from a vessel of 0.61m diameter plotted as lines of constant impeller

speed. The detailed impeller geometry was rather different to the 6PBT used in this study. The blade width used by Bujalski was over 20% wider than the 6PBT, whilst Bujalski also used blades with edges machined horizontally (to give a trapezoidal section through the blade), rather than simple angled plates which were used in this study. The data are compared in Fig 9.6 for the closest match of impeller speed between the two studies. The most probable explanation for the fact that the curve from this study lies below that of Bujalski is the difference in blade widths between the two studies. There were other differences, such as liquid type and vessel bottom geometry, which are complicating factors. Frijlink (1986) also reported very flat gassed power curves for upwards-pumping agitators.

The overall result is that the gassed power curves of this study are comparable with those of Bujalski, but it should be noted that the fall in P_g/P_u with increasing gas flow rate is a fairly strong function of impeller speed: this point does not emerge from Bujalski's work which covered fairly low ungassed power inputs per unit volume. In terms of producing flat gassed power curves, the upwards-pumping pitched blade turbine looks very promising, and is compared with the more traditional disc turbines in a later section.

9.4 GAS HOLD-UP

The main conclusions that may be drawn from the gas hold-up experiments are in broad agreement with the published data on hold-up.

To summarise the correlation methods used once more:

$$\epsilon_G = \alpha (\epsilon_m)^{\beta} v_{SG}^{\gamma} \quad (5.7)$$

$$\epsilon_G = \alpha' N^{\beta'} Q^{\gamma'} \quad (5.30)$$

Rushton and Bimbinet (1968), Davies (1986) and Bujalski (1987) all found that hold-up data over a range of tank sizes was correlated by equation (5.7), which was the conclusion drawn in this study for vessels up to $T = 2.67\text{m}$ for both coalescing and non-coalescing liquid. Bujalski (1987) found that the correlations for different vessels were statistically significantly different from each other, again in agreement with the results of this study. Equation (5.7) has little theoretical basis, but gives a reasonable fit to the data. No systematic scale effect has been found to suggest that significant problems would occur as a result of extrapolating equation (5.7) beyond the range of scales for which it has already been tested.

The difficulty which arises from the use of equation (5.7) in practice is the selection of the appropriate regression coefficients. These have shown considerable variation from vessel to vessel (Table 5.21). The literature correlations given in Table 2.1 show that, at least for water, the values of α , β , γ are in approximate agreement with established results. The comparison must be made carefully since some authors have included the isothermal expansion power of the gas in the power term. This tends to give higher values of β and lower values of γ than using the impeller energy dissipation rate alone. It is more instructive to look at the absolute values of hold-up predicted by the correlations for $\bar{\epsilon}_m < 2 \text{ W/kg}$ (the range of the current data) for constant superficial gas velocities $v_{SG} = 0.01 \text{ m/s}$ and $v_{SG} = 0.02 \text{ m/s}$ (Fig 9.7). The references from which the correlations were taken were:

- 1 Figueredo and Calderbank (1979)
- 2 Chapman et al (1983)
- 3 Bujalski (1987)
- 4 This study
- 5 Davies (1986)
- 6 Loiseau et al (1977)

At low superficial gas velocity, with the exception of the correlation of Chapman et al (1983), the agreement between the correlation predictions is extremely good. There is more divergence at high superficial gas velocity, but the agreement between this study and the data of Davies (1986) is remarkable. On the basis of these figures, it could be a worthwhile exercise to assemble the tap water hold-up data from references 1,3 4 and 5 to produce a general hold-up correlation. This would serve as a useful reference point both for other experimenters wishing to check their data and for the industrial designer.

To compare the alternative correlation method (equation (5.30)) with the literature data, the 0.61m tank will be considered for the T/3 Rushton turbine in tap water. The published correlations show that $0.8 < \beta' < 1.2$ and that $0.35 < \gamma' < 0.60$ (from Table 5.24). The data from this project are correlated by:

$$\epsilon_G = 0.46 N^{0.79} Q^{0.58} \quad (9.10)$$

Fig 9.8 compares this correlation with those of:

- 1 Warmoeskerken (1986) (for large cavities)
- 2 Hassan and Robinson (1979)
- 3 Yung et al (1979).

The gas flow rate has been taken as $2.75 \times 10^{-3} \text{ m}^3/\text{s}$ (1.0 VVM). The agreement between the four sets of data was extremely good, but a more testing comparison of the literature correlations is extrapolation to the 2.67m tank, for which the correlation produced in this project was:

$$\epsilon_G = 0.27 N^{0.55} Q^{0.7} \quad (9.11)$$

A flow rate of $0.1 \text{ m}^3/\text{s}$ has been adopted for comparative purposes (Fig 9.9). The errors in the correlations of Hassan

and Robinson and Yung et al are apparent at once. This probably resulted from the small scale at which the experimental work was carried out, and in the case of Yung et al, the inclusion of the Weber number. Since for constant physical properties:

$$We \propto N^2 D^3 \quad (9.12)$$

Weber number almost always increases with scale. Any correlation including We will have strong scale effects which, from the power data and the hold-up data, are completely spurious. Surface tension is not generally regarded as a suitable correlating parameter for either gassed power or hold-up in gas-liquid mixing. If it is to be included in correlations in a dimensionless form, it would be far more satisfactory from the point of view of scale-up to use the group (σ/σ_w) to account for surface tension. (σ_w = surface tension of water)

It is interesting to note that whilst Warmoeskerken's correlation was in good agreement with the current data at small scale (in spite of differences in (D/T) between the experiments), the hold-up at large scale was rather overestimated by Warmoeskerken's equation. No explanation for this is offered, since the effects of impeller diameter and (D/T) have already been shown to be complex. It is notable, however, that Warmoeskerken's equation was by far the best of the literature correlations, probably because it was at least partly based on data from a vessel with $T = 1.2m$.

Although the N, Q correlation method has the advantage of avoiding the necessity to predict the gassed power demand, some quite alarming errors can be made unless the correlations are applied very carefully, especially when considering scale-up. The $\bar{\epsilon}_m, v_{SG}$ method has the advantage of being relatively robust and the effects of (D/T) and changes of impeller type are accounted for in the correlation. Whilst agreeing with the philosophy behind the N, Q approach, the author recommends the

use of the $\bar{\epsilon}_m, v_{SG}$ method until much more work has been done on the influence of scale and geometry on the N,Q method.

If, by use of one of the correlations discussed in the previous section, the gassed power demand can be predicted to $\pm 20\%$, the error in the predicted hold-up is determined by the value of β , the exponent on power input:

$$(\Delta \epsilon_G) = \beta (\Delta \bar{\epsilon}_m) \quad (9.13)$$

Since β is typically in the range 0.2 - 0.3, the error in predicted hold-up is of the order of 5%. The correlation only fits the data to about $\pm 20\%$ in any case. The worst case is for this fit of $\pm 20\%$ to be converted to an error of $\pm 25\%$ as a result of a $\pm 20\%$ error in power estimation. Whilst not a desirable state of affairs, the situation is not quite as hopeless as it may seem from consideration of the problems of predicting the gassed power demand.

9.5 MASS TRANSFER

Comparison of mass transfer data with results of other researchers is a fairly difficult exercise. One of the most remarkable things about the published correlations on mass transfer is the variability of the predictions of $k_L a$ for a particular set of operating conditions. Fig 9.10 shows the predictions of correlations by the following authors, for a superficial gas velocity of 0.01 m/s:

- 1 This work
- 2 Smith, van't Riet and Middleton (1977)
- 3 Van't Riet (1975)
- 4 Davies (1986)
- 5 Van't Riet (1979)

The correlation for this data agrees exactly with van't Riet's correlation based on literature data. This should not

necessarily be regarded as a validation of the current work. Van't Riet used data which was obtained using various methods including the no gas phase depletion assumption. At such low values of $k_L a$, however, the assumed gas phase RTD should be of little significance, which probably explains the fact that there is fair agreement between correlations 1, 2, 3 and 5. Davies (1986) found much higher values of $k_L a$ using the double response method, but it should be noted that Davies' plug flow data showed fairly good agreement with the double response data for $k_L a < 0.1 \text{ s}^{-1}$. The data of Figueiredo and Calderbank (1978) are similar in magnitude to the results of Davies for these operating conditions, again exceeding the $k_L a$ values measured in this work by a considerable margin.

In the face of such problems, one can only conclude that the data fall within the established range found in the literature, if one may call anything to do with $k_L a$ established. No explanation appears to be available for the variation in results other than experimental problems. However, the range of $k_L a$ values covered is such that those problems (of probe response and assumed gas phase RTD) should be minimised. It should also be pointed out that no temperature correction has been applied to the plots of Fig 9.9. The correlation from this study would give 12% higher values of $k_L a$ if corrected to 20°C , which was the operating temperature used, for instance, by Davies. This does not account for the difference between this data and that of Davies, which is all the more puzzling because Davies used exactly the same type of oxygen probe as has been used here.

In a sense, the absolute value of $k_L a$ is not important from the point of view of scale-up. It is more important to establish whether or not the measured values of $k_L a$ are credible when compared with literature data. The general conclusion that $k_L a$ can be correlated by power input and superficial gas velocity up to industrial scales is the significant finding.

The influence of electrolyte on $k_L a$ is of some industrial

significance, and it is of interest to establish whether the very low dependency of $k_L a$ on v_{SG} reported by van't Riet (1979) is valid:

$$k_L a = 0.50 \bar{\epsilon}_m^{0.8} v_{SG}^{0.2} \quad (9.14)$$

If true, this would influence any proposed method of increasing $k_L a$: it would be more worthwhile to increase power input than v_{SG} . The correlation of Smith et al (1977) for non-coalescing liquids showed that there was a simple factor of 2 relating the non-coalescing value of $k_L a$ to the coalescing value. This is borne out by the experimental data of this study when comparing $k_L a$ values from electrolyte solution to those from water at identical operating conditions. There is no visible difference in the overall trends in the $k_L a$ data between coalescing and non-coalescing systems. Whilst not a new conclusion, this is a potentially useful result for the chemical engineer in industry and it is useful to produce further evidence in support of the rule.

9.6 SCALE-UP

From the finding that both gas hold-up and mass transfer coefficient, $k_L a$, are both correlated in terms of $\bar{\epsilon}_m$ and v_{SG} over a wide range of scales, one can consider the effect of applying particular scale-up rules, including those often used in industry. For scale-up, the case of constant mass transfer rate per unit volume will be taken as the reference point. The process result is assumed to be determined only by the specific mass transfer rate and not to be affected by other scale effects such as liquid phase mixing. The gas will be assumed to be in plug flow to enable the concentration driving force to be assessed, and the concepts developed in section 6.1 concerning depth effects will be used. Initially, however, the criterion for constant process result on scale-up will be taken as constant $k_L a$.

9.6.1 Constant Specific Power Input and VVM

Maintaining the gassed power input and flow expressed as vessel volumes per minute constant is a common industrial scale-up rule. VVM is the inverse of the gas residence time in minutes based on the reactor volume, so the reasoning behind the rule is to keep the liquid and gas residence times constant. This is sometimes referred to as constant stoichiometry.

The expression relating VVM to v_{SG} is:

$$\text{VVM} = \frac{60Q}{V} = \frac{60}{V} \frac{\pi T^2}{4} v_{SG} \quad (9.15)$$

Since, $V \propto T^3$,

$$\text{VVM} = \frac{k}{T} v_{SG} \quad (9.16)$$

$k = \text{constant}$

But,

$$k_L a = \alpha_1 (\bar{\epsilon}_m)^{\beta_1} (v_{SG})^{\gamma_1} \quad (9.17)$$

Since $\bar{\epsilon}_m$ is constant and v_{SG} is constant:

$$k_L a \propto T^{\gamma_1} \quad (9.18)$$

This represents considerable over-design, since γ_1 is approximately 0.5.

The true gas residence time is given by:

$$t_G = \frac{V_G}{Q} \quad (9.19)$$

Approximately,

$$V_G = \epsilon_G V_R \quad (9.20)$$

$V_R = \text{reactor volume.}$

Hold-up is given by:

$$\epsilon_G = \alpha_2 (\bar{\epsilon}_m)^{\beta_2} (v_{SG})^{\gamma_2} \quad (9.21)$$

For constant (VVM), Q/V_R is constant, so the gas residence time, t_G , becomes proportional to hold-up:

$$t_G \propto \epsilon_G \propto T^{\gamma_2} \quad (9.22)$$

The true gas residence time increases with scale, again showing the conservatism of the design rule. To reduce the gas residence time and $k_L a$ to constant values with scale-up, the power input per unit volume must be reduced.

9.6.2 Constant tip speed and (VVM)

If the power input per unit volume when gassed is proportional to $N^3 D^2$, then at constant tip speed ($= \pi ND$) :

$$\bar{\epsilon}_m \propto \frac{1}{T} \quad (9.23)$$

At constant tip speed and VVM, it follows from equations (9.16) and (9.21):

$$k_L a \propto T^{\gamma_1 - \beta_1} \quad (9.24)$$

$$\epsilon_G \propto T^{\gamma_2 - \beta_2} \quad (9.25)$$

Since, for virtually all the data available, $\gamma > \beta$, $k_L a$ and ϵ_G will still increase with scale, but not as strongly as with constant power input. Superficially, this appears to be an attractive result since by adopting the constant tip speed scale-up criterion, considerable energy savings are possible, as well as reductions in capital cost over the traditional power input per unit volume method. However, two limiting factors would probably cause problems if the tip speed criterion were used:

- 1 The liquid phase mixing time would increase more rapidly with scale than for the specific power input criterion since:

$$N\theta_m = \text{constant, independent of scale} \quad (9.26)$$

Poor liquid mixing could easily give reductions in yield if competitive reactions are present.

- 2 Flooding of the agitator would occur more easily than if the specific power input scale rule were used since for the scaled-up reactor, the agitator speed would be lower. If one expresses the flooding correlation of Nienow et al (1985) as:

$$\frac{Q}{N_F D^3} = 30 \left(\frac{D}{T}\right)^{3.5} \frac{N_F^2 D}{g} \quad (9.27)$$

and since at constant VVM,

$$Q \propto T^3 \quad (9.28)$$

$$N_F \propto \frac{1}{D^{1/3}} \quad (9.29)$$

At constant tip speed:

$$N \propto \frac{1}{D} \quad (9.30)$$

The agitator speed falls more rapidly with scale than the predicted speed to prevent flooding. Even if $N > N_F$ at small scale, there will be a point at which the agitator floods as scale is increased.

9.6.3 Constant Specific Power Input and v_{SG}

According to the correlations, $k_L a$ and ϵ_G will be constant if this scale-up rule is adopted. Nienow and Wisdom (1978) pointed out the problem of depletion of the gas phase if scale-up is

attempted at constant v_{SG} . However, there is a counter-effect since the concentration driving force is increased with increasing scale as a result of compression of the gas.

If the gas is in plug flow, the log mean driving force can be used to give the overall transfer rate provided that Henry's law applies; and that the dissolved gas concentration in the bulk liquid is zero:

$$R_V = k_L a \frac{C_i^* - C_o^*}{\ln \left(\frac{C_i^*}{C_o^*} \right)} \quad (9.31)$$

$C^* = KC_G$, K = equilibrium constant.

The concentration of gas in the liquid phase has been taken as zero for simplification. From an overall mass balance on the gas:

$$Q_i C_i - Q_o C_o = V(k_L a) \frac{C_i^* - C_o^*}{\ln \left(\frac{C_i^*}{C_o^*} \right)} \quad (9.32)$$

If $Q_i = Q_o$, this reduces to

$$C_o = C_i \exp \left(- \frac{KV k_L a}{Q} \right) \quad (9.33)$$

If $k_L a$ is constant and v_{SG} is constant, then:

$$C_o = C_i \exp \left(- \frac{k T k_L a}{v_{SG}} \right) = C_i \exp (-k' T) \quad (9.34)$$

k = constant, depending on geometry

$$k' = \text{constant} = \frac{kT}{v_{SG}}$$

Depth effects on C_i are given by:

$$C_i = \frac{p_o + \rho_L gH}{RT_a} = \frac{p_o}{RT_a} \left(1 + \frac{H}{H_o} \right) \quad (9.35)$$

The volumetric transfer rate may now be expressed as:

$$R_v = k_L a \frac{C_i^* \left(1 - \frac{C_o}{C_i^*}\right)}{\ln \frac{C_i^*}{C_o}} \quad (9.36)$$

$$= k_L a K \frac{1 - e^{-k' T}}{k' T} \frac{p_o}{RT} \left(1 + \frac{T}{H_o}\right)$$

Since $k'T \ll 1$ for normal operating conditions; this is approximately given by:

$$R_v = k_L a \frac{p_o}{RT} \left(1 + \frac{T}{H_o}\right) \frac{[k'T - (\frac{k'T}{2})^2]}{k'T} \quad (9.37)$$

$$R_v = k_L a \frac{p_o}{RT} \left(1 + \frac{T}{H_o}\right) \left(1 - \frac{k'T}{2}\right) \quad (9.38)$$

For pressurised vessels (large H_o) depletion will tend to dominate and the volumetric transfer rate will fall on scale-up. For operation at atmospheric pressure, H_o is 10.3m, and a typical value for k' would be around 0.2. For this case ($1/H_o \approx k'/2$) depletion effects will still predominate over depth effects. The implication is that depletion effects will only become insignificant for very low values of $k_L a$, which was clear from the beginning.

An intermediate scale-up law for gas flow rate between constant v_{SG} and constant VVM can be selected according to the system parameters to account for depletion by increasing $k_L a$. This would have the advantage of reducing the risk of flooding compared to the (VVM) rule and also reduce the gas consumption, which can be a significant cost. Keeping the impeller power input per unit volume constant should help to minimise the potential loss of yield as a result of poor liquid phase mixing.

A possible objection to this strategy is that the gas residence time would no longer be constant on scale-up. Rather than reject the suggested scale-up method on this basis, it might be

better to question the requirement for constant gas residence time. The reactions take place in the liquid, not the gas, and the requirement is to keep the same mass transfer rate of gaseous reactant into unit volume of liquid. This is not directly related to the gas residence time; the success of the scale-up process should depend only on volumetric transfer rate.

9.7 COMPARISON OF AGITATOR TYPES

To make a correct comparison of agitator types, one should consider the situation where the ungassed power input is constant, unless the motor and drive train are sized to operate only under gassed conditions. In many instances, it is desirable for the agitator to be capable of operating under ungassed conditions.

Consider the 5 agitators tested in electrolyte solution. The ungassed power input has been taken as 1.0 W/kg (the T/3 Rushton turbine operating at 270 RPM in the 0.61m tank) and the gassing rate has been taken as $2.75 \times 10^{-3} \text{ m}^3/\text{s}$ (1.0 VVM). Table 9.1 shows the results of this exercise with the correlation predictions at the calculated power input being very close to each other. More divergence appeared in the actual hold-up values recorded at the nominal operating conditions, but the speeds tested did not correspond to exactly 1.0 W/kg ungassed power input.

The results show that there is in actual fact little to be gained from using agitators other than the Rushton turbine, unless one considers flooding. An impeller which can handle larger gas flow rates than the traditional impellers without flooding will probably have many advantages since it would be able to operate efficiently at higher gassing rates.

10 CONCLUSIONS AND RECOMMENDATIONS

The work documented in this thesis has largely focused on the extension of existing knowledge to tanks of industrial scale. The experiments carried out at large scale were, for the most part, as accurate and as reliable as the small-scale experiments. The conclusions are considered for power, hold-up, mass transfer and flooding in turn.

10.1 POWER

- 1 There was a definite scale effect on the unaerated power number of the Rushton turbine which agreed well with the recent correlation of Bujalski et al (1987).
- 2 Scale effects on the gassed power curves of the Rushton turbine were observed for scale-up at constant ungassed power input per unit volume and constant Froude number.
- 3 The concave turbine gave flatter power curves than the Rushton turbine at all scales. The results of van't Riet et al (1976) were, however, not exactly reproduced.
- 4 A 6-bladed pitched blade turbine, pumping upwards, gave very flat power curves with increasing gassing rate at constant impeller speed. A significant difference in power number between upwards and downwards-pumping was observed.

10.2 HOLD-UP

- 1 The use of the following correlation method for gas hold-up was found to give the best results:

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^{\beta} v_{SG}^{\gamma}$$

This accounted for scale and changes in (D/T) fairly successfully, although detailed analysis of the

correlations obtained in different vessels showed that there are weak scale effects which caused differences in the correlations obtained for each vessel. Very good agreement with published correlations was found.

- 2 The advantages of the alternative approach to correlation using speed and gas flow rate are recognised, but further work would appear to be necessary on scale effects in particular and geometric factors in general.
- 3 There are no advantages in terms of energy efficiency (energy required to produce given hold-up) to be gained from using agitators other than the Rushton turbine. This also applied if one considered constant ungassed power per unit volume.
- 4 Non-coalescing liquids produced significantly higher hold-up than coalescing liquids at the same operating conditions. The ratio of hold-ups in the two liquid types appeared to be systematically affected by the gas flow rate.

10.3 MASS TRANSFER

- 1 The same conclusions can be drawn on scale-up for mass transfer as for gas hold-up.
- 2 The Rushton and concave turbines have been compared at both large and small scale. At small scale, in non-coalescing liquid, the concave turbine tended to produce higher values of $k_L a$ than the Rushton turbine. At large scale, in water, there was a small but statistically significant difference between the agitators, with the concave turbine giving the higher values of $k_L a$.
- 3 The methods of analysing mass transfer experiments have been re-evaluated with particular attention to the effects

of scale on the concentration driving force available for mass transfer.

- 4 For each set of mass transfer tests, an approximately linear relationship between $k_L a$ and hold-up has been obtained. The data were always within $\pm 20\%$ of the regression line, but systematic effects of speed and gas flow rate were visible, which suggested that $k_L a$ is not truly proportional to hold-up.

10.4 FLOODING

- 1 For Rushton turbines, the flooding correlation of Nienow et al (1985), has been confirmed at scales up to $T = 2.67\text{m}$:

$$(Fl)_F = 30 \left(\frac{D}{T}\right)^{3.5} (Fr)_F$$

- 2 Concave turbines appear to flood at slightly higher gas flow rates at the same agitator speed as a Rushton turbine of identical diameter. At constant ungassed power input per unit volume, the concave turbine would be able to handle approximately 40% more gas than the Rushton turbine without flooding.
- 3 Flooding of the upwards-pumping pitched blade turbine has been briefly studied at one scale. This type of agitator is greatly superior to disc turbines in its ability to disperse gas without flooding. Some oscillations in flow pattern were observed in the region of the flooding point.
- 4 By considering the gassed power curves, the scale effects on the flooding/loading transition for downwards-pumping pitched blade turbines found by Chapman, Nienow, Cooke and Middleton (1983) have been confirmed up to a tank diameter of 2.67m.

10.5 PROCESS ECONOMICS

A method of making an objective economic comparison of different agitation options to achieve a particular process task has been developed. The results show clearly that research work can lead to potentially large savings in overall mixing costs. An investigation of the potential benefits of increasing throughput using knowledge of gas-liquid mixing showed further potential economic gains to be made by the application of information from research work.

10.6 SUGGESTIONS FOR FURTHER WORK

The work described in this thesis could usefully be extended to cover a wider range of agitator types. It would be useful to investigate hold-up and mass transfer at small scale and also at a second, larger scale. In particular, the performance of pitched blade turbines at large scale needs to be more fully investigated, especially for upwards-pumping operation. The differences between pitched blade turbines and disc turbines are worth investigating further, especially for mass transfer, which has not been covered in this work. The large scale mass transfer experiments showed some problems which could have been caused by the state of liquid mixing in the vessel. Again, further large scale work is needed to investigate these phenomena, especially if a wider range of impeller power input can be covered than was possible here. This is a line of enquiry which leads to the requirement for a full understanding of the hydrodynamics of a gassed stirred vessel, accounting for liquid behaviour as well as gas behaviour, which, of course, is the ultimate aim of research such as this.

As a subsidiary point, the influence of additives such as anti-foam should be investigated in a systematic way. This could lead to huge problems for the experimenter, since some way of classifying the anti-foams available would be needed to make a generally applicable correlation, but to start with a few

commercial anti foams and to carry out some hold-up and mass transfer measurements would be very valuable.

One could produce an enormous list of questions which still require answers in the field of gas-liquid mixing. It does not seem a worth-while exercise to do that here, so the comments of this section are intended only to highlight some of the ways in which the current work could be usefully extended in the relatively short term.

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Appendix 3A Gas Inlet Temperatures

At first sight, a gas inlet temperature of 60°C could cause a problem for mass transfer measurement since $k_L a$ is quite sensitive to liquid temperature. The rate of change of gas temperature with time must be considered. By assuming that all the resistance to heat transfer is in the liquid phase, the following expression may be written for the gas phase temperature:

$$\rho_g V_B C_p \frac{dT_G}{dt} = h A_B (T_G - T_w) \quad (3A.1)$$

T_G = gas temperature

T_w = water temperature

C_p = specific heat of gas = 1000 J/kgK

V_B = bubble volume

A_B = area of bubble

ρ_g = gas density = 1.27 kg/m³ at inlet.

The heat transfer coefficient, h , may be estimated by regarding the bubble as an isolated sphere. From standard theory, the Nusselt number of a stationary sphere is 2 (this is a conservative assumption).

$$Nu = \frac{hd_B}{k} = 2 \quad (3A.2)$$

For water, $k = 0.53$ W/mK, and the bubble size is of the order of 3×10^{-3} m, from which the heat transfer coefficient is approximately 350 W/m²K. Equation (3A.1) may be solved to give the gas temperature as a function of time:

$$T_G = T_w + (T_{Go} - T_w) e^{-\alpha t} \quad (3A.3)$$

$$\alpha = \frac{h A_B}{\rho_g V_B C_p}$$

α is the time constant of the system, and may be calculated from the estimated values already available. In this case, α is equal to about 140. The gas temperature is thus within 1% of the water temperature within about 40 milliseconds. The gas would be in thermal equilibrium with the liquid before it reached the agitator, and the effects of gas inlet temperature can be ignored, except for the flow rate measurement. The gas temperature at the

orifice plate was measured to enable the accurate calculation of the gas density at the measuring point.

APPENDIX 5A : Gas Hold-Up in Electrolyte Solutions

All the hold up data in the 0.61m diameter vessel were obtained using a solution of 2.5kg sodium chloride in tap water with a total volume of 0.165 m³. As has already been discussed, the ultra-sonic level detector is sensitive to the presence of foam on the dispersion surface, and it was for this reason that a relatively low electrolyte concentration was selected. During the tests, the surface was observed to check that no permanent foam layer was formed beneath the ultra-sonic level detector.

A complete set of hold-up data for the five agitator types had been acquired, and plotting the data out showed very large difference between the upwards-pumping pitched blade turbine and the 0.47T Rushton turbine. At high power input, the Rushton turbine produced almost twice as much hold-up as the PBT. However, the two sets of data were not taken using the same batch of electrolyte solution, so it was decided to repeat a few experiments using each agitator in a fresh batch of electrolyte. The differences between the pitched blade turbine data from the new batch of electrolyte and the original data were very large and, yet, visually, the solutions were similar in behaviour. The original data are given in Tables A1 and A2, the second set of data has been used for correlation and comparison purposes in the main body of the thesis.

The Rushton turbine experiments were repeated for a limited set of data. The results are given in Table A3, showing that the original data were conservative. The order to err on the side of caution, the original data were retained, and have been used to compare with the 6PBT pumping upwards.

Naturally, the question arises as to why these differences in results appeared. Some checking of the original PBT data was done using visual observation, which was in fair agreement with the ultra-sonic data. The highly disturbed nature of the surface made estimation of the level difficult, a problem which is also likely to

have affected the ultra-sonics, and may itself be part of the cause of the discrepancy.

The following experiments were carried out to try to isolate the source of the discrepancies:

(i) Repeatability

Some of the pitched blade turbine experiments were repeated several times over the course of a day. The first experiments tended to give lower values of hold-up, but after several hours of testing, the results were extremely reproducible, as shown in Table A4.

(ii) Sparger alignment

The sparge ring, which was originally placed 15mm below the lower edge of the agitator blades was lowered to 30mm clearance, and several tests repeated. The hold-up compared very closely with that obtained at the original sparger clearance (Table A5).

(iii) Contamination

It was suspected that a possible cause of the problem was contamination of the liquid with surfactant. A quantity of activated carbon was added to the solution to absorb soluble impurities such as organic compounds. It was reasoned that at a concentration of less than 0.1 wt % of activated carbon, the influence of the solids on gas hold-up would be negligible, as was found by Chapman, Nienow, Middleton and Cooke (1983).

A small reduction in hold-up occurred as a result of adding the solids (Table A6), but no visible reduction of the foam occurred. The addition of a second quantity of activated carbon actually increased the measured hold-up slightly, but this was probably because the activated carbon appeared to be stabilising the thin foam layer. Having obtained some evidence of the presence of impurities which had some influence on hold-up, it was decided to attempt to

remove the foam, which was causing a measurement problem. A commercial anti-foaming agent (Nalco 8639 manufactured by Nalco International Ltd, Oakbrook, Illinois, USA) was added to the electrolyte/activated carbon mixture, at a concentration of a few tens of ppm. The effect was a dramatic collapse in the gas hold-up (Table A6), and a radical change in the visual appearance of the dispersion. The bubbles in the region of the vessel wall became much larger as a result of adding the anti-foam, which appeared to work by promoting coalescence.

This set of experiments did not establish conclusively the cause of the differences between the two sets of PBT data. There may have been an instrumentation problem which can no longer be traced, or there may have been some contamination in the vessel which reduced the gas hold-up. The experience is a reminder of the care needed in setting up experiments in this field, and the result of the experiment with anti-foam shows how difficult it can be to translate laboratory data into industrial applications, many of which involve the use of additives of some kind and for which experimental data is almost certainly not available.

TABLE A1 : RAW HOLD-UP DATA : 6 BLADE, PITCHED BLADE TURBINE, 0.61 M TANK

UPWARDS PUMPING - ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP ϵ_G
2.0	0.000613	0.233	0.00210	0.0130
2.0	0.00145	0.226	0.00498	0.0287
2.0	0.00215	0.224	0.00736	0.0257
2.0	0.00288	0.218	0.00988	0.0349
2.0	0.00365	0.209	0.01249	0.0434
2.0	0.00449	0.205	0.0163	0.0528
2.0	0.00616	0.200	0.0211	0.0660
2.0	0.00770	0.199	0.0264	0.0790
2.0	0.01031	0.200	0.0353	0.0971
3.0	0.000613	0.746	0.00210	0.0182
3.0	0.00145	0.713	0.00498	0.0378
3.0	0.00215	0.690	0.00735	0.0391
3.0	0.00288	0.656	0.00988	0.0479
3.0	0.00365	0.610	0.01249	0.0551
3.0	0.00449	0.592	0.0163	0.0706
3.0	0.00616	0.558	0.0211	0.0842
3.0	0.00847	0.511	0.0290	0.0982
3.0	0.01031	0.510	0.0353	0.1108
3.0	0.01144	0.499	0.0392	0.1145
4.0	0.000613	1.745	0.0021	0.0269
4.0	0.00145	1.632	0.00498	0.0471
4.0	0.00215	1.570	0.00735	0.0498
4.0	0.00288	1.450	0.00988	0.0603
4.0	0.00365	1.364	0.01249	0.0717
4.0	0.00476	1.243	0.0163	0.0770
4.0	0.00770	1.073	0.0264	0.1027
4.0	0.01031	1.031	0.0353	0.1176
5.0	0.00145	3.142	0.00498	0.0537
5.0	0.00215	2.905	0.00735	0.0556
5.0	0.00288	2.770	0.00988	0.0612
5.0	0.00365	2.524	0.01249	0.0736
5.0	0.00476	2.284	0.0163	0.0791
5.0	0.00770	2.081	0.0211	0.0924
5.0	0.01031	1.850	0.0353	0.1158

TABLE A2 : RAW HOLD-UP DATA : 6 BLADE, PITCHED BLADE TURBINE, 0.61 M TANK

DOWNWARDS- PUMPING ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
2.0	0.000613	0.226	.00210	.0170
2.0	0.00145	0.168	.00498	.0216
2.0	0.00215	0.167	.00736	.0229
2.0	0.00288	0.169	.00988	.0268
2.0	0.00365	0.173	.01249	.0328
2.0	0.00449	0.163	.01537	.0389
3.0	0.000613	0.759	.00210	.0266
3.0	0.00145	0.408	.00498	.0416
3.0	0.00215	0.369	.00736	.0407
3.0	0.00288	0.369	.00988	.0463
3.0	0.00365	0.363	.01249	.0490
3.0	0.00449	0.348	.01537	.0550
3.0	0.00616	0.361	.02110	.0659
4.0	0.000613	1.828	.00210	.0322
4.0	0.00145	0.861	.00498	.0505
4.0	0.00215	0.760	.00736	.0516
4.0	0.00288	0.671	.00986	.0614
4.0	0.00365	0.632	.01249	.0686
4.0	0.00449	0.657	.01537	.0732
4.0	0.00616	0.650	.02110	.0854

Table A3 : Rushton turbine (D=0.47T) electrolyte data.

Impeller Speed (rev/s)	Gas Flow (m ³ /s)	Hold-up	
		Original Data	Repeated data
1.5	0.00275	0.0329	0.0396
2.0		0.0452	0.0567
2.5		0.0593	0.0721
3.0		0.0744	0.0843
3.5		0.0892	0.0969
4.0		0.1030	0.1094
4.5		0.1167	0.1233
5.0		0.1253	0.1372

Table A4 : Repeatability of PBT data in electrolyte

Impeller Speed (rev/s)	Gas Flow (m ³ /s)	Hold-up				
		Lowered Sparger				
		I	II	III	IV	V
2.0	.00275	.0419	.0421	.0424		
3.0		.0553	.0652	.0654	.0652	.0657
4.0		.0709	.0875	.0858		
5.0		.0943	.1014	.1069		

Table A5 : Effect of lowering sparge ring on gas hold-up.

Speed = 3.0 rev/sec

Gas Flow (m ³ /s)	Original Position	Lower Position	
		IV	V
.00215	.0531	.0546	.0531
.00288	.0645	.0652	.0657
.00365	.0755	.10760	.10778
.00449	.0863	.0858	.0884

Table A6 : Effect of contaminants on hold-up (using Rushton turbine).

Flow = 0.00275 m³/s

Speed	150g Activated Carbon rev/s) (0.09 wt%)	300g Activated Carbon (0.18 wt%)	Anti- Foam	Uncontaminated hold-up
2.5	0.0667	0.0672	0.0361	0.0721
3.0	0.0783	-	-	0.0843
3.5	0.0917	0.0934	0.0412	0.0969
4.0	0.1054	0.1068	0.0447	0.1094
4.5	0.1192	-	-	0.1233
2.5(repeat)		0.06650	-	-

APPENDIX 5B : Statistical analysis of data Correlation

In the absence of standard statistics packages, a set of computer programs was written to correlate the data using regression techniques. The first approach was to linearise the data:

$$Z = \alpha x^{\beta} y^{\gamma} \quad (A1)$$

$$\log Z = \log \alpha + \beta \log x + \gamma \log y \quad (A2)$$

Linear regression was then used to calculate α , β and γ . The alternative approach of using non-linear least squares regression (McCalla 1972) which was rather more complex to program, was also developed. A comparison between the linearised and non-linear methods showed that while the fit to the data was quite similar, the regression coefficients were apparently more sensitive to data points well away from the regression surface for the linearised technique. A comparison of the two methods is given in Table 5B1.

The correlation coefficient has been frequently quoted in the main body of the thesis as a measure of the fit of the regression to the experimental data. It is defined as:

$$R^2 = \frac{\sigma^2 - E^2}{\sigma^2} \quad (A3)$$

σ^2 = variance about the mean for the dependent variable

$$= \sum_{i=1}^n (y_i - \bar{y})^2 \quad (A4)$$

E^2 = error sum of squares about correlating function

$$= \sum_{i=1}^n (y_{pi} - y_i)^2 \quad (A5)$$

y_{pi} = predicted value of dependent variable for i^{th} data point.

R^2 has been quoted as the correlation coefficient because it represents the fraction of the total variance about the mean which is accounted for by the correlation.

Significance Testing

The analysis of variance approach has been used to determine whether the difference between two correlations is statistically significant.

For a given data set, the regression coefficients represent the best fit to the data using the correlating equation. If a second set of regression coefficients are then applied to the data set, the resulting error sum of squares must be greater than the error sum of squares obtained using the best fit regression coefficients. The F-statistic is calculated as :

$$F = \frac{\text{Error sum of squares from new coefficients}}{\text{Error sum of squares from best fit.}}$$

The value of F is compared with statistical tables as a test for significance, with the number of degrees of freedom being (n-2).

The only problem with this technique occurred when considering two sets of data (A and B) on the borderline of significance. It was possible to test regression coefficients B on data A and find no significant difference, and to test coefficients A on data B to find that there was a significant difference, but not generally at the 1% confidence level. It was therefore decided to consider as valid differences which were predicted as significant at the 1% confidence level.

TABLE 3B1 : COMPARISON OF REGRESSION TECHNIQUES

$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma$						
VESSEL DIAMETER (M)	REGRESSION METHOD	REGRESSION COEFFICIENTS		CORRELATION COEFFICIENT	MEAN ERROR (%)	MAXIMUM ERROR (%)
		α	β			
0.61	LINEAR	1.58	0.53	0.956	9.6	19.5
	NON LINEAR	1.94	0.58	0.961	10.0	28.4
1.83	LINEAR	1.65	0.21	0.936	4.8	18.9
	NON LINEAR	1.25	0.20	0.945	4.7	23.5
2.67	LINEAR	4.33	0.38	0.934	4.4	11.3
	NON LINEAR	3.04	0.34	0.942	4.5	15.1

APPENDIX 8A : Derivation of Cost Equations

For capital costs, we have (Middleton, 1978):

$$C = £2200 \frac{P_k^{0.7}}{N^{0.8}} \quad (A1)$$

Impeller speed is given by (Eq. 8.1):

$$N = \frac{b\left(\frac{T}{D}\right)^a}{\theta_m (Po)^{\frac{1}{3}}} \quad (A2)$$

Power is given by:

$$P = Po \cdot \rho N^3 D^5 \quad (A3)$$

Simplifying,

$$C = £2200 (Po\rho)^{0.7} N^{1.3} D^{3.5} \quad (A4)$$

Using appropriate values of a and b , we eventually arrive at Equation (9) by combining (A1) and (A3).

For running costs, we have:

$$R = U_T U_C P_k \quad (8.22)$$

It is then straightforward to obtain Equation (8.23) from Equations (A1) and (A3).

Continuous Mixers

For the case of a continuous mixer, the same equations apply, but an expression for mixing time must be included, since mixing time is a function of vessel volume. In this case,

$$\theta_m = \frac{V}{4Q} \quad (A5)$$

This simply states that the mixing time for the desired process result is one-quarter of the residence time in the tank. By using

$$V = \frac{\pi T^3}{4}, \text{ we obtain:}$$

$$\theta_m = \frac{\pi T^3}{16Q} \quad (A6)$$

Equation 8.23 for capital costs can therefore be written as:

$$C = \frac{\pounds 14600}{\pi^{1.3}} \frac{(16Q)^{1.3}}{T^{0.4}} \left(\frac{D}{T}\right)^{0.68} \quad (A7)$$

In the nomenclature of Equation 8.28:

$$\beta = \pounds 121,300 Q^{1.3}$$

Equation 8.24 for annual running costs can be written as, using Equation (A6),

$$R = \frac{\pounds 547}{\pi^{1.3}} \frac{(16Q)^3}{T^{2.5}} \frac{U_T U_c}{D^{1.5}} \quad (A8)$$

In the nomenclature of Equation 8.28:

$$\gamma = \pounds 7230 Q^3 U_T U_c . F$$

Where $F = \frac{(1+i)^n - 1}{i(1+i)^n}$, converting annual running costs to a present worth.

TABLE 2.1 : HOLD-UP CORRELATIONS - POWER PER UNIT VOLUME

REFERENCE	CORRELATION	TANK SIZE (H)	LIQUIDS / OTHER COMMENTS
Calderbank (1958)	$\epsilon_G = \left(\frac{v_{SG} \epsilon_G}{u_t} \right)^{0.5} + 2.16 \times 10^{-4} \left(\frac{P}{\rho} \right)^{0.2} \left(\frac{v_{SG}}{u_t} \right)^{0.05}$	0.5	Water, pure organic liquids
Rushton & Bimblinet (1968)	$\frac{\epsilon_G}{1-\epsilon_G} = 0.31 \left(\frac{P}{V} \right)^{0.31} v_{SG}^{0.60}$	0.22-0.91	Water 0.18 < D/T < 0.53
Loiseau et al (1977)	$\epsilon_G = \frac{0.011}{\rho} \frac{P}{0.36 \mu} \left(\frac{P}{V} \right)^{0.27} v_{SG}^{0.36}$	0.22	Water, glycol, solution, electrolyte solutions, organic acid solutions, water + surfactant
Figueriedo & Calderbank (1978)	$\epsilon_G = 0.23 \left(\frac{P}{V} \right)^{0.26} v_{SG}^{0.68}$	0.91	Water
Chapman et al (1983)	$\epsilon_G = 0.23 \left(\frac{P}{V} \right)^{0.31} v_{SG}^{0.67}$	0.56	Water
Davies (1986)	(1) $\epsilon_G = 0.02 \left(\frac{P}{V} \right)^{0.40} v_{SG}^{0.35}$ (2) $\epsilon_G = 0.14 \left(\frac{P}{V} \right)^{0.20} v_{SG}^{0.49}$	0.31-1.83	Water (1) = clinging cavities (2) = "3-3" structure
van't Riet (1975)	$\epsilon_G = 0.14 \left(\frac{P}{V} \right)^{1/3} v_{SG}^{2/3}$		Water

TABLE 2.2 : HOLD-UP CORRELATIONS - IMPELLER SPEED

REFERENCE	CORRELATION	TANK SIZE (M)	COMMENTS
Van Dierendonck (1968)	$\epsilon_G = 0.31 \left(\frac{\mu^v}{\sigma} \right)^{2/3} \left(\frac{\rho \sigma^3}{g \mu^4} \right)^{1/6} + \frac{0.45 (N - N^*) D^2}{\sqrt{g T^3}}$	0.45	N* = minimum speed for surface aeration
Hassan & Robinson (1977)	$\epsilon_G = 0.113 \left(\frac{Q N^2}{\sigma} \right)^{0.57}$	0.15, 0.29	Non-electrolyte solutions
Yung et al (1979)	$\epsilon_G = 0.0052 F l^{0.5} W e^{0.65} \left(\frac{D}{T} \right)^{1.4}$	0.4	Non-electrolyte solutions
Warmoschkerken (1986)	$\epsilon_G = 0.133 Q^{0.35} N^{0.16} D^{0.16} (1)$ $\epsilon_G = 0.177 Q^{0.4} N^{0.9} D^{0.1} (2)$	0.44, 1.2	Water (1) = clinging cavities (2) = "3-3" structure
Yoshida and Miura (1963)	$\epsilon_G \propto N^{0.8} D^{1.2} V_{SG}^{0.75}$	0.25-0.59	Alkali solutions

TABLE 2.3 : k_L A CORRELATIONS - COALESCING SYSTEMS

AUTHORS	CORRELATIONS	TANK DIAMETER (M)	COMMENTS
van't Riet (1975)	$k_L a = 0.05 \left(\frac{P}{V} \right)^{1/3} v_{SG}$	0.44	Dynamic response, no depletion $0.0025 < v_{SG} < 0.04 \text{ m/s}$ $600 < \frac{P}{V} < 7000 \text{ W/m}^3$
Smith et al (1977)	$k_L a = 0.01 \left(\frac{P}{V} \right)^{0.475} v_{SG}^{0.4}$	0.44-1.83	Dynamic response, no depletion $100 < \frac{P}{V} < 500 \text{ W/m}^3$ $0.004 < v_{SG} < 0.02 \text{ m/s}$
van't Riet (1979)	$k_L a = 0.1 \left(\frac{P}{V} \right)^{0.4} v_{SG}^{0.5}$	-	Based on literature data from several sources
Nishikawa et al (1981)	$k_L a = C v_{SG}^{1/3} \left(\frac{P}{V} \right)^{0.8}$	0.15-0.60	Dynamic response or chemical method. Different correlations for different flow regimes. This equation valid for agitator control
Figueredo & Calderbank (1979)	$k_L a = 0.001 (P_g)^{0.58} v_{SG}^{0.75} \frac{T}{V}$	0.38, 0.91	Dynamic response, well-mixed gas phase $400 < \frac{P}{V} < 4000 \text{ W/m}^3$ $0.0064 < v_{SG} < 0.013 \text{ m/s}$
Dralj & Sincic (1984)	$k_L a = 0.18 \left(\frac{P}{V} \right)^{0.38} v_{SG}^{0.38}$	0.1 non coalescing	Dynamic response, no depletion $0.001 < v_{SG} < 0.01 \text{ m/s}$ $100 < \frac{P}{V} < 10000 \text{ W/m}^3$ (P_g estimated from paper)
Warmskerken (1986)	$\frac{k_L a}{N} = 4.23 \times 10^{-17} F_l^{0.6} Re$ $\frac{k_L a}{N} = 1.89 \times 10^{-7} F_l^{0.4} Re$	0.44, 0.64	Clinging cavities "3-3" cavities Dynamic double response method

TABLE 2.4 : $k_L a$ CORRELATIONS - NON-COALESING SYSTEMS

AUTHORS	CORRELATION	TANK DIAMETER (M)	COMMENTS
Smith et al (1977)	$k_L a = 0.02 \left(\frac{P_g}{V} \right)^{0.475} v_{SG}^{0.4}$	0.44-1.83	0.11 m potassium sulphate
van't Riet (1979)	$k_L a = 2 \times 10^{-13} \left(\frac{P_g}{V} \right)^{0.8} v_{SG}^{0.2}$	-	Literature data
Davies (1986)	$k_L a = 1.05 \times 10^{-3} \left(\frac{P_t}{V} \right)^{0.73} v_{SG}^{0.11}$ $k_L a = 1.05 \times 10^{-3} \left(\frac{P_t}{V} \right)^{0.68} v_{SG}^{0.0}$	0.31-0.91	Clinging cavities "3-3" cavities 0.11 m potassium sulphate

TABLE 2.5 : WORKERS COMPARING DIFFERENT AGITATOR TYPES

AUTHORS	QUANTITY MEASURED	AGITATORS
Chapman Nienow Middleton Cooke (1983)	ϵ_G $k_L a$	Rushton Turbine Pitched blade (upwards and downwards-pumping) Axial flow (Downflow less efficient)
Hoecker Langer & Werner (1981)	$k_L a$	Rushton Mig Flat disc (Higher viscosity than water)
Warmoeskerken Speur & Smith (1984)	$k_L a$	Rushton Downwards-pumping PBT (different blade angles)
van't Riet (1975)	$k_L a$	Rushton turbine (6 and 12 bl) Convex, concave, perforated
Zlokarnik (1978)	$k_L a$	Rushton turbine Propeller

TABLE 2.6 : RESULTS OF HASSAN AND ROBINSON

$$\epsilon_G = C \left(\frac{QN^2}{\sigma} \right)^Z$$

IMPELLER TYPE	FLUID TYPE	C	Z
	Non-electrolyte	0.113	0.57
Rushton Turbine	0.4 kmol/m ³ Na ₂ SO ₄ solution	0.209	0.44
6 blade paddle	Non-electrolyte	0.102	0.65
4 blade paddle	Non-electrolyte	0.316	0.50

TABLE 3.1 : SPARGER DIMENSIONS

Tank diameter (m)	Sparger diameter (m)	Number of holes	Hole diameter (m)	Hole direction
0.61	0.122	30	0.002	DOWN
	0.210	30	0.002	DOWN
	0.519	30	0.002	UP
1.83	0.362	140	0.004	DOWN
2.67	0.533	270	0.004	DOWN
	0.945	270	0.004	DOWN

TABLE 4.1 : DISC TURBINE POWER NUMBERS (UNGASSED)

TANK DIAMETER (M)	POWER NUMBER
0.61	5.13
1.83	5.31
2.67 (i)	5.69
(ii)	5.71

TABLE 4.2 : CONCAVE TURBINE POWER NUMBERS (UNGASSED)

TANK DIAMETER (M)	POWER NUMBER
0.61 (i)	3.65
(ii)	3.88
1.83	4.45
2.67 (i)	4.43
(ii)	4.52

TABLE 4.3 : PITCHED BLADE TURBINE POWER NUMBERS (UNGASSED)

TANK DIAMETER (M)	NUMBER OF BLADES	PUMPING DIRECTION	POWER NUMBER
0.61	4	DOWN	1.43
2.67	4	DOWN	1.59
0.61	6	DOWN	1.96
0.61	6	UP	1.75

TABLE 5.1 : COMPARISON OF HOLD-UP METHODS

Ultra-Sonics Hold-Up	Spill-Over Hold-Up	% Difference
0.0563	0.0550	+ 2.4
0.0602	0.0662	- 9.6
0.0499	0.0509	- 2.0
0.0648	0.0638	+ 1.5
0.0757	0.0723	+ 4.5
0.0825	0.0792	+ 4.0
0.0984	0.0930	+ 5.5
0.1067	0.0983	+ 7.9

TABLE 5.2 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 0.61 M TANK, TAP WATER

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
3.50	0.001375	0.272	0.0047	0.0267
3.50	0.00208	0.235	0.0071	0.0358
3.50	0.00260	0.225	0.0089	0.0408
3.50	0.00336	0.209	0.0115	0.0466
3.50	0.00415	0.226	0.0142	0.0470
4.50	0.001375	0.634	0.0047	0.0318
4.50	0.00208	0.515	0.0071	0.0417
4.50	0.00260	0.475	0.0089	0.0480
4.50	0.00336	0.443	0.0115	0.0563
4.50	0.00415	0.412	0.0142	0.0641
4.50	0.00477	0.378	0.0163	0.0705
5.00	0.001375	0.917	0.0047	0.0355
5.00	0.00208	0.715	0.0071	0.0461
5.00	0.00260	0.661	0.0089	0.0513
5.00	0.00336	0.612	0.0115	0.0614
5.00	0.00415	0.571	0.0142	0.0697
5.00	0.00477	0.535	0.0163	0.0764
5.50	0.001375	1.207	0.0047	0.0376
5.50	0.00208	0.963	0.0071	0.0489
5.50	0.00260	0.880	0.0089	0.0553
5.50	0.00336	0.829	0.0115	0.0652
5.50	0.00415	0.784	0.0142	0.0740
5.50	0.00477	0.726	0.0163	0.0815
5.75	0.001375	1.347	0.0047	0.0384
5.75	0.00208	1.121	0.0071	0.0594
5.75	0.00260	1.024	0.0089	0.0570
5.75	0.00336	0.957	0.0115	0.0682
5.75	0.00415	0.908	0.0142	0.0761
5.75	0.00477	0.867	0.0163	0.0783

TABLE 5.3 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 0.61 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
4.5	0.0007	0.796	0.0023	0.044
4.5	0.0014	0.558	0.0047	0.048
4.5	0.0021	0.465	0.0071	0.060
4.5	0.0027	0.427	0.0094	0.068
4.5	0.0041	0.378	0.0141	0.083
4.5	0.0055	0.343	0.0188	0.095
4.5	0.0082	0.360	0.0282	0.112
2.58	0.0027	0.094	0.0094	0.029
3.62	0.0027	0.201	0.0094	0.035
4.12	0.0027	0.311	0.0094	0.047
4.65	0.0027	0.440	0.0094	0.057
5.00	0.0027	0.545	0.0094	0.069
3.1	0.0007	0.230	0.0023	0.017
3.1	0.0014	0.152	0.0047	0.026
3.1	0.0021	0.141	0.0071	0.033
3.1	0.0027	0.149	0.0094	0.033
3.1	0.0034	0.147	0.0117	0.034
2.58	0.0014	0.094	0.0047	0.022
3.62	0.0014	0.251	0.0047	0.031
4.12	0.0014	0.366	0.0047	0.035
4.65	0.0014	0.543	0.0047	0.044

TABLE 5.4 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 0.61 M TANK, ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
4.5	0.001375	0.551	0.000471	0.0445
4.5	0.00206	0.451	0.00706	0.0554
4.5	0.00275	0.412	0.00941	0.0680
4.5	0.00344	0.384	0.01176	0.0785
4.5	0.00413	0.347	0.01411	0.0864
4.5	0.00550	0.313	0.01882	0.1003
5.5	0.00138	1.212	0.00471	0.0554
5.5	0.00206	0.933	0.00706	0.0643
5.5	0.00275	0.835	0.00941	0.0761
5.5	0.00344	0.788	0.01176	0.0887
5.5	0.00413	0.746	0.01411	0.0983
5.75	0.001375	1.315	0.00471	0.0574
5.75	0.00206	1.066	0.00706	0.0667
5.75	0.00275	0.951	0.00941	0.0761
5.75	0.00344	0.891	0.01176	0.0851
5.75	0.00413	0.848	0.01411	0.0943
5.0	0.00275	0.556	0.00941	0.0730
4.65	0.00275	0.459	0.00941	0.0686
4.12	0.00275	0.316	0.00941	0.0591
5.5	0.00275	0.834	0.00941	0.0787
4.5	0.00275	0.412	0.00941	0.0658

TABLE 5.5 : RAW HOLD-UP DATA

RUSHTON TURBINE, 0.47T, 0.61 M TANK, ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
3.5	0.001375	1.732	0.00471	0.0715
3.5	0.00206	1.421	0.00706	0.0851
3.5	0.00275	1.246	0.00941	0.0892
3.5	0.00344	1.159	0.01176	0.1006
3.5	0.00413	1.072	0.01411	0.1075
3.0	0.00138	1.147	0.00471	0.0601
3.0	0.00206	0.910	0.00706	0.0709
3.0	0.00275	0.800	0.00941	0.0744
3.0	0.00344	0.786	0.01176	0.0832
3.0	0.00413	0.693	0.01411	0.0904
2.5	0.001375	0.681	0.00471	0.0489
2.5	0.00206	0.538	0.00706	0.0560
2.5	0.00275	0.480	0.00941	0.0593
2.5	0.00344	0.457	0.01176	0.0682
2.5	0.00413	0.418	0.01411	0.0726
1.5	0.00275	0.124	0.00941	0.0329
2.0	0.00275	0.260	0.00941	0.0452
3.25	0.00275	1.008	0.00941	0.0800
4.0	0.00275	1.812	0.00941	0.1030
4.5	0.00275	2.584	0.00941	0.1167
5.0	0.00275	3.392	0.00941	0.1253

TABLE 5.6 : RAW HOLD-UP DATA

CONCAVE TURBINE, T/3, 0.61 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
3.1	0.0027	0.156	0.0094	0.039
3.62	0.0027	0.232	0.0094	0.052
4.12	0.0027	0.341	0.0094	0.061
4.65	0.0027	0.495	0.0094	0.063
5.00	0.0027	0.617	0.0094	0.069
2.58	0.0014	0.097	0.0047	0.026
3.62	0.0014	0.273	0.0047	0.033
4.12	0.0014	0.418	0.0047	0.040
4.65	0.0014	0.586	0.0047	0.050
5.00	0.0014	0.780	0.0047	0.057
5.50	0.0014	0.958	0.0047	0.068
4.50	0.0007	0.672	0.0023	0.031
4.50	0.0014	0.565	0.0047	0.044
4.50	0.0021	0.487	0.0071	0.056
4.50	0.0027	0.448	0.0094	0.062
4.50	0.0041	0.418	0.0141	0.077
4.50	0.0055	0.373	0.0188	0.091
4.50	0.0082	0.411	0.0282	0.117
3.10	0.0007	0.218	0.0023	0.016
3.10	0.0014	0.170	0.0047	0.028
3.10	0.0021	0.152	0.0070	0.037
3.10	0.0027	0.157	0.0094	0.045
3.10	0.0034	0.159	0.0117	0.045
3.10	0.0041	0.155	0.0141	0.052

TABLE 5.7 : RAW HOLD-UP DATA

CONCAVE TURBINE, T/3, 0.61 M TANK, ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
4.5	0.001375	0.615	0.00471	0.0414
4.5	0.00206	0.490	0.00706	0.0492
4.5	0.00275	0.459	0.00941	0.0511
4.5	0.00344	0.408	0.01176	0.0700
4.5	0.00413	0.381	0.01411	0.0824
3.5	0.00138	0.273	0.00471	0.0280
3.5	0.00206	0.233	0.00706	0.0293
3.5	0.00275	0.222	0.00941	0.0367
3.5	0.00344	0.201	0.01176	0.0475
3.5	0.00413	0.199	0.01411	0.0536
5.75	0.001375	1.301	0.00471	0.0562
5.75	0.00206	1.024	0.00706	0.0665
5.75	0.00275	0.926	0.00941	0.0763
5.75	0.00344	0.841	0.01176	0.0869
5.75	0.00413	0.804	0.01411	0.1119
5.5	0.00138	1.145	0.00471	0.0531
5.5	0.00206	0.898	0.00706	0.0633
5.5	0.00275	0.807	0.00941	0.0691
5.5	0.00344	0.760	0.01176	0.0869
5.5	0.00413	0.711	0.01411	0.1060
5.0	0.001375	0.859	0.00471	0.0475
5.0	0.00206	0.674	0.00706	0.0562
5.0	0.00275	0.601	0.00941	0.0594
5.0	0.00344	0.562	0.01176	0.0808
5.0	0.00413	0.547	0.01411	0.0955
5.75	0.00550	0.788	0.01882	0.1263
5.75	0.00275	1.459	0.00941	0.0462
5.5	0.00275	1.287	0.00941	0.0411
5.0	0.00275	0.970	0.00941	0.0392

TABLE 5.8 : RAW HOLD-UP DATA

4 BLADE, PITCHED BLADE TURBINE, 0.61 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.03	0.00275	0.013	0.0094	0.0382
1.38	0.00275	0.023	0.0094	0.0414
1.85	0.00275	0.133	0.0094	0.0464
2.07	0.00275	0.166	0.0094	0.0494
2.75	0.00275	0.286	0.0094	0.0615
1.03	0.001375	0.038	0.0047	0.0251
1.38	0.001375	0.069	0.0047	0.0228
1.85	0.001375	0.122	0.0047	0.0288
2.07	0.001375	0.175	0.0047	0.0350
2.75	0.001375	0.360	0.0047	0.0536
2.07	0.00208	0.165	0.0071	0.0360
2.07	0.00275	0.165	0.0094	0.0444
2.07	0.00342	0.180	0.0117	0.0540
2.07	0.00413	0.161	0.0141	0.0620
2.07	0.00550	0.187	0.0188	0.0717
2.07	0.00825	0.201	0.0282	0.1090
4.13	0.00067	1.303	0.0023	0.0320
4.13	0.001375	0.950	0.0047	0.0432
4.13	0.00208	0.817	0.0071	0.0510
4.13	0.00275	0.678	0.0094	0.0486
4.13	0.00413	0.756	0.0141	0.0702
4.13	0.00550	0.782	0.0188	0.0823

TABLE 5.9 : RAW HOLD-UP DATA : 6 BLADE, PITCHED BLADE TURBINE, 0.61 M TANK

DOWNWARDS- PUMPING ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
2.0	0.00145	0.168	.00498	.0272
2.0	0.00215	0.167	.00736	.0304
2.0	0.00288	0.169	.00988	.0343
2.0	0.00365	0.173	.01249	.0427
2.0	0.00449	0.163	.01537	.0503
3.0	0.00145	0.408	.00498	.0452
3.0	0.00215	0.369	.00736	.0520
3.0	0.00288	0.369	.00988	.0578
3.0	0.00365	0.363	.01249	.0620
3.0	0.00449	0.348	.01537	.0699
4.0	0.00145	0.861	.00498	.0600
4.0	0.00215	0.760	.00736	.0662
4.0	0.00288	0.671	.00986	.0734
4.0	0.00365	0.632	.01249	.0821

TABLE 5.10 : RAW HOLD-UP DATA : 6 BLADE, PITCHED BLADE TURBINE, 0.61 M TANK

UPWARDS PUMPING - ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP ϵ_G
2.0	0.00145	0.226	0.00498	0.0225
2.0	0.00215	0.224	0.00736	0.0331
2.0	0.00288	0.218	0.00988	0.0421
2.0	0.00365	0.209	0.01249	0.0514
2.0	0.00449	0.205	0.0163	0.0599
2.0	0.00616	0.200	0.0211	0.0789
2.0	0.00770	0.199	0.0264	0.0910
2.0	0.01031	0.200	0.0353	0.1110
3.0	0.00145	0.713	0.00498	0.0415
3.0	0.00215	0.690	0.00735	0.0541
3.0	0.00288	0.656	0.00988	0.0645
3.0	0.00365	0.610	0.01249	0.0772
3.0	0.00449	0.592	0.0163	0.0865
3.0	0.00616	0.558	0.0211	0.1050
3.0	0.00847	0.511	0.0290	0.1170
4.0	0.00145	1.632	0.00498	0.0639
4.0	0.00215	1.570	0.00735	0.0766
4.0	0.00288	1.450	0.00988	0.0868
4.0	0.00365	1.364	0.01249	0.0975
4.0	0.00476	1.243	0.0163	0.1065
5.0	0.00145	3.142	0.00498	0.0848
5.0	0.00215	2.905	0.00735	0.0942
5.0	0.00365	2.524	0.01249	0.1040
5.0	0.00476	2.284	0.0163	0.1103
5.0	0.00770	2.081	0.0211	0.1348

TABLE 5.11 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 1.83 M TANK, TAP WATER

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER* INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.92	0.0536	0.319	0.0204	0.0717
2.17	0.0536	0.451	0.0204	0.0775
2.33	0.0536	0.571	0.0204	0.0806
2.83	0.0536	0.742	0.0204	0.0911
3.08	0.0536	1.035	0.0204	0.0957
3.33	0.0536	1.709	0.0204	0.1005
3.70	0.0536	2.330	0.0204	0.1067
3.33	0.0276	2.335	0.0105	0.061
3.33	0.0411	1.828	0.0156	0.082
3.33	0.0495	1.731	0.0188	0.0944
3.33	0.0595	1.657	0.0226	0.1040
3.33	0.0691	1.572	0.0263	0.1140
3.33	0.0776	1.433	0.0295	0.1198
3.33	0.0873	1.262	0.0332	0.1247
1.97	0.0205	0.628	0.0078	0.0257
1.97	0.0269	0.455	0.0102	0.0374
1.97	0.0352	0.365	0.0134	0.0509
1.97	0.0447	0.344	0.0170	0.0638
1.97	0.0534	0.331	0.0203	0.0723
1.97	0.0602	0.325	0.0229	0.0729
1.97	0.0638	0.328	0.0243	0.0814
1.97	0.0667	0.327	0.0254	0.0827
1.80	0.0214	0.410	0.0081	0.0250
1.80	0.0301	0.304	0.0114	0.0422
1.80	0.0396	0.281	0.0150	0.0550
1.80	0.0436	0.283	0.0166	0.0620
1.80	0.0476	0.276	0.0181	0.0650
1.80	0.0516	0.270	0.0196	0.0680
1.80	0.0563	0.269	0.0214	0.0716
2.6	0.0301	0.900	0.0114	0.0535
2.6	0.0389	0.820	0.0148	0.0663
2.6	0.0447	0.788	0.0170	0.0750
2.6	0.0541	0.737	0.0206	0.0854
2.6	0.0611	0.663	0.0232	0.0930
2.6	0.0682	0.629	0.0260	0.0982
2.6	0.0770	0.605	0.0293	0.1047
2.6	0.0854	0.594	0.0325	0.1123
3.33	0.132	1.209	0.0502	0.139
3.33	0.155	1.196	0.0589	0.147
3.33	0.184	1.126	0.0700	0.152
3.33	0.207	1.119	0.0787	0.159
3.0	0.0857	1.076	0.0326	0.1133
3.0	0.1182	0.897	0.0449	0.1242
3.0	0.1400	0.885	0.0532	0.1376
3.0	0.1648	0.858	0.0627	0.1469
3.0	0.1825	0.852	0.0694	0.1519
3.0	0.1291	0.885	0.0491	0.1290
3.0	0.1534	0.869	0.0583	0.1385

*Power assumed to be equal to that measured in octanol solution.

TABLE 5.12 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 1.83 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
2.17	0.0536	0.462	0.0204	0.0995
2.33	0.0536	0.586	0.0204	0.1036
2.83	0.0536	0.765	0.0204	0.1185
3.08	0.0536	1.070	0.0204	0.1255
3.33	0.0536	1.768	0.0204	0.1305
3.70	0.0536	2.417	0.0204	0.1389
1.97	0.0205	0.640	0.0078	0.0431
1.97	0.0269	0.464	0.0102	0.0570
1.97	0.0352	0.373	0.0134	0.0723
1.97	0.0447	0.352	0.0170	0.0841
1.97	0.0534	0.339	0.0203	0.0932
1.97	0.0602	0.334	0.0229	0.0990
1.97	0.638	0.335	0.0243	0.1015
1.97	0.0667	0.335	0.0254	0.1032
1.80	0.0207	0.418	0.0078	0.0431
1.80	0.0296	0.310	0.0113	0.0616
1.80	0.0396	0.287	0.0151	0.0736
1.80	0.0436	0.287	0.0166	0.0775
1.80	0.0476	0.281	0.0181	0.0827
1.80	0.0516	0.276	0.0196	0.0856
1.80	0.0563	0.275	0.0214	0.0900
2.6	0.0222	1.440	0.0084	0.0660
2.6	0.0296	0.926	0.0113	0.0801
2.6	0.0389	0.846	0.0148	0.0943
2.6	0.0447	0.811	0.0170	0.1007
2.6	0.0541	0.756	0.0206	0.1089
2.6	0.0611	0.677	0.0232	0.1121
2.6	0.0682	0.641	0.0259	0.1158
2.6	0.077	0.617	0.0293	0.122
2.6	0.0901	0.615	0.0343	0.130
2.6	0.0184	1.496	0.0070	0.058
2.6	0.0222	1.440	0.0084	0.066
2.6	0.0251	1.220	0.0095	0.073
3.0	0.0265	1.705	0.0101	0.0819
3.0	0.0331	1.393	0.0126	0.0906
3.0	0.0420	1.302	0.0160	0.0962
3.0	0.0477	1.269	0.0180	0.1054
3.0	0.0522	1.258	0.0198	0.1066
3.0	0.0709	1.062	0.0270	0.1230

TABLE 5.13 : RAW HOLD-UP DATA

CONCAVE TURBINE, T/3, 1.83 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.80	0.0245	0.435	0.0093	0.0505
	0.0312	0.332	0.0119	0.0631
	0.0391	0.302	0.0149	0.0730
	0.0440	0.297	0.0167	0.0796
	0.0480	0.289	0.0182	0.0821
	0.0526	0.283	0.0200	0.0846
	0.0572	0.279	0.0217	0.0864
	0.0218	0.462	0.0083	0.0418
1.97	0.0274	0.374	0.0104	0.0547
	0.0257	0.532	0.0098	0.0594
	0.0295	0.467	0.0112	0.0650
	0.0356	0.402	0.0135	0.0744
	0.0477	0.371	0.0181	0.0859
	0.0540	0.359	0.0205	0.0911
	0.0594	0.352	0.0226	0.0960
	0.0641	0.350	0.0244	0.0994
2.6	0.0657	0.343	0.0250	0.0986
	0.0702	0.344	0.0267	0.1036
	0.0276	1.162	0.0105	0.0799
	0.0344	0.961	0.0131	0.0908
	0.0426	0.871	0.0162	0.1012
	0.0501	0.834	0.0190	0.1072
	0.0577	0.810	0.0219	0.1099
	0.0616	0.794	0.0234	0.1193
3.0	0.0768	0.778	0.0292	0.1210
	0.0791	0.765	0.0301	0.1269
	0.0857	0.737	0.0326	0.1290
	0.0205	1.386	0.0078	0.0623
	0.0239	1.909	0.0091	0.0834
	0.0293	1.605	0.0111	0.0931
	0.0391	1.387	0.0149	0.1067
	0.0471	1.301	0.0179	0.1116
1.92	0.0548	1.251	0.0208	0.1180
	0.0622	1.216	0.0236	0.1243
	0.0695	1.206	0.0264	0.1299
	0.0764	1.165	0.0290	0.1322
	0.0828	1.157	0.0315	0.1326
	0.0901	1.143	0.0342	0.1381
	0.0536	0.342	0.0204	0.0883
	0.0536	0.476	0.0204	0.0967
2.17	0.0536	0.595	0.0204	0.1012
2.33	0.0536	1.006	0.0204	0.1161
2.83	0.0536	1.364	0.0204	0.1193
3.08	0.0536	1.745	0.0204	0.1242
3.33	0.0536			

TABLE 5.14 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 2.67 M TANK, TAP WATER

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.42	0.0664	0.347	0.0119	0.0474
1.42	0.0880	0.312	0.0158	0.0625
1.42	0.1110	0.299	0.0199	0.0700
1.58	0.0664	0.476	0.0119	0.0503
1.58	0.0880	0.424	0.0158	0.0659
1.58	0.1110	0.402	0.0199	0.0767
1.58	0.1338	0.387	0.0240	0.0868
1.58	0.1579	0.372	0.0283	0.0951
1.58	0.1852	0.403	0.0332	0.1003
1.67	0.0664	0.545	0.0119	0.0510
1.67	0.0880	0.494	0.0158	0.0680
1.67	0.1110	0.467	0.0199	0.0789
1.67	0.1338	0.446	0.0240	0.0879
1.67	0.1579	0.422	0.0283	0.0980
1.67	0.1852	0.413	0.0332	0.1092

TABLE 5.15 : RAW HOLD-UP DATA

RUSHTON TURBINE, T/3, 2.67 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.43	0.0553	0.574	0.01	0.063
1.30	0.0553	0.429	0.01	0.056
1.17	0.0553	0.311	0.01	0.053
1.15	0.0553	0.314	0.01	0.052
1.03	0.0553	0.203	0.01	0.045
0.90	0.0553	0.119	0.01	0.033
1.48	0.081	0.423	0.0146	0.083
1.42	0.081	0.365	0.0146	0.079
1.35	0.081	0.320	0.0146	0.077
1.28	0.081	0.276	0.0146	0.072
1.22	0.081	0.234	0.0146	0.067
1.50	0.0553	0.654	0.010	0.067
1.50	0.0665	0.518	0.0120	0.075
1.50	0.0776	0.437	0.0140	0.083
1.50	0.1000	0.395	0.0180	0.093
1.42	0.0441	0.575	0.008	0.055
1.42	0.0553	0.552	0.0100	0.063
1.42	0.0665	0.417	0.0120	0.073
1.42	0.0776	0.358	0.0140	0.078
1.42	0.1000	0.329	0.0180	0.090
1.27	0.0441	0.424	0.008	0.043
1.27	0.0553	0.366	0.010	0.053
1.27	0.0665	0.280	0.012	0.063
1.27	0.0776	0.257	0.014	0.069
1.27	0.0888	0.249	0.016	0.074
1.27	0.1000	0.256	0.018	0.076
1.27	0.1112	0.250	0.020	0.077
1.58	0.0553	0.665	0.010	0.068
1.58	0.0665	0.551	0.012	0.075
1.58	0.0776	0.472	0.014	0.080
1.58	0.1000	0.441	0.018	0.093
1.58	0.1120	0.448	0.020	0.098

TABLE 5.16 : RAW HOLD-UP DATA

CONCAVE TURBINE, T/3, 2.67 M TANK, TAP WATER

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.25	0.0664	0.251	0.0119	0.0410
1.25	0.0880	0.251	0.0158	0.0548
1.42	0.0664	0.368	0.0119	0.0445
1.42	0.0880	0.308	0.0158	0.0609
1.42	0.1110	0.294	0.0199	0.0699
1.58	0.0664	0.536	0.0119	0.0475
1.58	0.0880	0.421	0.0158	0.0649
1.58	0.1110	0.401	0.0199	0.0754
1.58	0.1338	0.387	0.2040	0.0851
1.67	0.0664	0.638	0.0119	0.0493
1.67	0.0880	0.490	0.0158	0.0671
1.67	0.1110	0.469	0.0199	0.0780
1.67	0.1338	0.445	0.0240	0.0878

TABLE 5.17 : RAW HOLD-UP DATA

CONCAVE TURBINE, T/3, 2.67 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
1.27	0.0445	0.325	0.0080	0.043
1.27	0.0557	0.293	0.0100	0.055
1.27	0.0667	0.250	0.0120	0.064
1.27	0.0780	0.237	0.0140	0.072
1.27	0.0890	0.225	0.0159	0.075
1.27	0.1002	0.225	0.0180	0.079
1.27	0.1112	0.240	0.0199	0.083
1.42	0.0445	0.438	0.0080	0.050
1.42	0.0557	0.392	0.0100	0.061
1.42	0.0667	0.390	0.0120	0.066
1.42	0.0780	0.352	0.0140	0.074
1.42	0.0890	0.320	0.0159	0.081
1.42	0.1002	0.312	0.0180	0.085
1.42	0.1112	0.304	0.0199	0.091
1.58	0.0445	0.631	0.0080	0.055
1.58	0.0557	0.593	0.0100	0.065
1.58	0.0667	0.540	0.0120	0.073
1.58	0.0780	0.505	0.0140	0.083
1.58	0.0890	0.479	0.0159	0.090
1.58	0.1002	0.457	0.0180	0.095
1.58	0.1112	0.442	0.0199	0.102
0.90	0.0557			0.037
1.03	0.0557			0.043
1.15	0.0557			0.047
1.17	0.0557			0.052
1.30	0.0557	Data not available		0.054
1.43	0.0557			0.062
1.53	0.0557			0.066
1.22	0.0812			0.072
1.28	0.0812			0.074
1.35	0.0812			0.077
1.42	0.0812			0.079
1.48	0.0812			0.083

TABLE 5.18 : RAW HOLD-UP DATA

4 BLADE, PITCHED BLADE TURBINE, 2.67 M TANK, OCTANOL SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	GAS HOLD-UP
0.75	0.0627	0.147	0.0112	0.0456
0.93	0.0627	0.301	0.0112	0.0564
1.00	0.0627	0.326	0.0112	0.0597
1.17	0.0627	0.637	0.0112	0.0651
1.25	0.0627	0.734	0.0112	0.0673
1.33	0.0627	0.868	0.0112	0.0673
1.30	0.0503	0.837	0.0090	0.0651
1.30	0.0749	0.553	0.0134	0.0760
1.30	0.1000	0.562	0.0179	0.0814
1.30	0.1251	0.585	0.0224	0.0933
1.30	0.1754	0.589	0.0314	0.0998
1.30	0.2251	0.596	0.0403	0.1107
1.30	0.2754	0.642	0.0493	0.1215
1.17	0.0749	0.450	0.0134	0.0664
1.17	0.1000	0.435	0.0179	0.0772
1.17	0.1251	0.451	0.0224	0.0838
1.17	0.1503	0.463	0.0269	0.0903
1.17	0.1754	0.462	0.0314	0.0968
1.17	0.2000	0.479	0.0358	0.1012
1.00	0.0503	0.412	0.0090	0.0467
1.00	0.0749	0.318	0.0134	0.0673
1.00	0.1000	0.327	0.0179	0.0738
1.00	0.1251	0.329	0.0224	0.0792
1.00	0.1503	0.331	0.0269	0.0835
1.00	0.1754	0.351	0.0314	0.0879
1.00	0.2000	0.340	0.0358	0.0966

TABLE 5.19 : RUSHTON TURBINE HOLD-UP CORRELATIONS

0.61 m & 2.67 m TANKS, OCTANOL SOLUTION

$$\epsilon_G = \alpha (\bar{\epsilon}_M)^\beta x^\gamma$$

x = variable representing gassing rate

x	α	β	γ	MEAN ERROR (%)	MAX ERROR (%)	CORRELATION COEFFICIENT
v_{SG}	2.02	0.41	0.68	8.4	46.6	0.938
VVM	0.11	0.39	0.15	26.7	64.6	0.289
Q	0.13	0.27	0.11	20.6	99.8	0.491

TABLE 5.20 : EFFECT OF INCLUDING GAS ISOTHERMAL EXPANSION POWER

$$\epsilon_G = \alpha (\bar{\epsilon}_m + C v_{SG})^\beta v_{SG}^\gamma$$

C	α	β	γ	MAX ERROR (%)	CORRELATION COEFFICIENT
0.0	1.25	0.20	0.62	23.5	0.945
0.2	1.21	0.21	0.61	23.4	0.945
0.4	1.18	0.21	0.61	23.3	0.945
0.6	1.14	0.22	0.60	23.2	0.944
0.8	1.11	0.23	0.60	23.1	0.944
1.0	1.07	0.23	0.59	23.0	0.944
2.0	0.92	0.27	0.56	22.6	0.942

TABLE 5.21 : RUSHTON TURBINE - SUMMARY OF HOLD-UP DATA

$$\epsilon_G = \alpha (\epsilon_m)^\beta v_{SG}^\gamma$$

LIQUID	TANK DIAMETER (M)	IMPELLER DIAMETER	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT	MEASURING METHOD
0	0.61	T/3	1.94	0.58	0.63	10.0	28.4	0.961	S
0	1.83	T/3	1.25	0.20	0.62	4.8	23.5	0.945	S
0	2.67	T/3	3.04	0.34	0.79	4.5	15.1	0.942	S
0	OCTANOL COMBINED	T/3	1.36	0.22	0.64	8.0	35.3	0.945	S
E	0.61	T/3	1.47	0.18	0.63	3.1	9.3	0.976	U
E	0.61	0.47T	0.97	0.38	0.53	2.7	11.1	0.989	U
E	0.61	COMBINED	1.44	0.33	0.61	5.8	21.4	0.939	U
W	0.61	T/3	1.48	0.24	0.69	2.2	14.0	0.983	U
W	1.83	T/3	0.59	0.19	0.49	7.9	47.9	0.947	S
W	2.67	T/3	1.41	0.18	0.71	3.3	6.9	0.978	U
W	COMBINED	T/3	0.63	0.17	0.51	7.5	47.6	0.954	-

TABLE 5.22 : CONCAVE TURBINE - SUMMARY OF HOLD-UP DATA

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma$$

LIQUID	TANK DIAMETER (M)	IMPELLER DIAMETER	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT	MEASURING METHOD
0	0.61	T/3	1.34	0.42	0.58	6.3	21.8	0.977	S
0	1.83	T/3	0.85	0.21	0.53	4.5	28.2	0.956	S
0	2.67	T/3	2.29	0.25	0.74	2.6	9.3	0.977	S
0	COMBINED	T/3	0.90	0.21	0.54	6.5	37.3	0.961	S
E	0.61	T/3	2.64	0.47	0.73	8.4	28.3	0.949	U
W	2.67	T/3	2.02	0.18	0.81	3.0	7.0	0.987	U

TABLE 5.23 : PITCHED BLADE TURBINE HOLD-UP CORRELATIONS

$$\epsilon_G = \alpha (\epsilon_m)^\beta v_{SG}^\gamma$$

LIQUID	TANK DIAMETER (M)	NO. OF BLADES	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT	MEASURING METHOD
0	0.61	4 (DOWN)	1.07	0.10	0.62	8.6	28.6	0.941	S
0	2.67	4 (DOWN)	0.48	0.20	0.42	3.4	15.5	0.972	S
0	COMBINED	4 (DOWN)	0.55	0.12	0.47	6.4	28.7	0.950	S
E	0.61	6 (UP)	1.55	0.32	0.66	6.0	24.3	0.964	U
E	0.61	6 (DOWN)	0.929	0.48	0.50	4.0	11.5	0.979	U
E	0.61	COMBINED	1.38	0.32	0.63	7.9	26.7	0.955	U

TABLE 5.24 : RANGE OF OPERATING CONDITIONS - RUSHTON TURBINE,
OCTANOL SOLUTION

VESSEL DIAMETER (M)	RANGE OF POWER INPUT (W/KG)	RANGE OF SUPERFICIAL GAS VELOCITY (M/S)
0.61	0.1 - 0.8	0.002 - 0.028
1.83	0.3 - 2.4	0.01 - 0.027
2.67	0.1 - 0.7	0.01 - 0.02

TABLE 5.25 : RUSHTON TURBINE - SUMMARY OF HOLD-UP DATA

$$\epsilon_G = \alpha N^{\beta} Q^{\gamma}$$

LIQUID	TANK DIAMETER (M)	IMPELLER DIAMETER	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT	MEASURING METHOD
O	0.61	T/3	0.14	1.61	0.54	9.8	23.7	0.961	S
O	1.83	T/3	0.27	0.60	0.52	4.5	21.1	0.955	S
O	2.67	T/3	0.21	1.02	0.55	3.7	17.2	0.963	S
O	COMBINED	T/3	0.14	0.84	0.37	9.4	33.1	0.909	S
E	0.61	T/3	0.68	0.63	0.55	3.5	8.8	0.975	U
E	0.61	0.47T	0.21	1.06	0.37	2.9	9.7	0.987	U
E	0.61	COMBINED	0.84	0.26	0.47	16.0	56.2	0.487	U
W	0.61	T/3	0.46	0.79	0.58	2.7	15.7	0.979	U
W	1.83	T/3	0.16	0.62	0.43	7.5	44.2	0.957	S
W	2.67	T/3	0.24	0.63	0.66	2.7	7.3	0.985	U
W	COMBINED	T/3	0.10	0.87	0.34	9.2	46.2	0.947	-

TABLE 5.26 : CONCAVE TURBINE - SUMMARY OF HOLD-UP DATA

$$\epsilon_G = \alpha N^8 Q^\gamma$$

LIQUID	TANK DIAMETER (M)	IMPELLER DIAMETER	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT	MEASURING METHOD
0	0.61	T/3	0.20	1.24	0.51	6.3	20.8	0.976	S
0	1.83	T/3	0.21	0.64	0.45	3.9	25.6	0.980	S
0	2.67	T/3	0.31	0.85	0.68	3.0	9.2	0.980	S
0	COMBINED	T/3	0.13	0.85	0.35	8.2	40.2	0.941	S

TABLE 5.27 : STATISTICAL SIGNIFICANCE TESTING

COMPARISON	BASIS	TANK DIAMETER (M)	LIQUID	F	DEGREES OF FREEDOM	SIGNIFICANCE	
						5%	1%
Rushton v Concave	Power	0.61	0	1.57	19	N	N
		1.83	0	1.62	37	N	N
		2.67	0	1.56	30	N	N
		COMBINED	0	1.27	89	N	N

TABLE 5.28 : COMBINED DISC TURBINE AND PITCHED BLADE

TURBINE HOLD-UP CORRELATIONS - OCTANOL SOLUTIONS

$$\epsilon_G = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma$$

TANK DIAMETER (M)	α	β	γ	MEAN (%)	MAXIMUM (%)	CORRELATION COEFFICIENT
0.61	1.19	0.18	0.62	14.8	64.1	0.852
2.67	0.46	0.18	0.40	8.5	34.1	0.829

Table 6.1(a) Results of simulation of bubble motion Full model

Water depth = 2.67 m

Temperature = 20°C

Initial bubble diameter (mm)	Bubble rise velocity (m/s)	Residence time (s)	Mean mass transfer rate [kmol/s x 10 ¹²]	Mean mass transfer flux [kmol/m ₂ s x 10 ⁸]
5	0.001	2670	0.265	1.37
	0.005	534	1.165	5.75
	0.01	267	1.1730	8.40
	0.05	53.4	2.52	12.00
	0.1	26.7	2.65	12.60
	1.0	2.67	2.77	13.13
3	0.001	2670	0.057	0.83
	0.01	267	0.474	6.47
	0.1	26.7	0.922	12.19
	1.0	2.67	0.995	13.09
1	0.001	2670	0.0042	0.54
	0.01	267	0.0211	2.67
	0.1	26.7	0.0870	10.45
	1.0	2.67	0.109	12.88
1	0.1	26.7	0.087	10.45
2			0.393	11.72
3			0.922	12.19
4			1.674	12.40
5			2.650	12.60
6			3.850	12.68

Table 6.1(b) Simulation results : scale-up using full model

Temperature = 20°C

$d_{B0} = 0.005\text{m}$

Water depth (m)	Rise velocity (m/s)	Residence time (s)	Mean Tansfer rate [kmol/s x 10 ²]
0.61	0.1	6.1	2.37
1.00	10.0		2.43
1.50	15.0		2.50
2.00	20.0		2.56
2.67	26.7		2.65
5.00	50.0		2.94
0.61	0.0229	26.7	2.28
1.00	0.0375		2.35
1.50	0.0562		2.44
2.0	0.0749		2.53
2.67	0.1		2.65
5.00	0.187		3.06

Table 6.1(c) Bubble Simulation : neglecting nitrogen transfer

Initial Bubble Size (mm)	Rise Velocity (m/s)	Mean Transfer rate [kmol/s x 10 ¹²]	Mean Transfer flux [kmol/m ² s x 10 ⁸]
5	0.001	0.265	1.43
	0.01	1.751	8.80
	0.1	2.660	12.67
	1.0	2.775	13.14
1	0.001	0.0037	0.51
	0.01	0.0211	2.82
	0.1	0.0878	10.75
	1.0	0.109	12.93
1	0.1	0.0878	10.75
2		0.395	11.91
3		0.926	12.32
4		1.680	12.54
5		2.66	12.67
6		3.86	12.75

Table 6.1(d) Bubble simulation : no nitrogen transfer, no depth effects : high inlet concentration

Water depth (m)	Rise velocity (m/s)	Mean Transfer rate [kmol/s x 10 ¹²]	Mean Transfer flux [kmol/m ² s x 10 ⁸]
0.61	0.1	2.38	11.90
1.00		2.47	12.61
1.50		2.56	13.07
2.00		2.64	13.52
2.67		2.75	14.12
5.00		3.12	16.08
0.61	0.0229	2.31	11.83
1.00	0.0375	2.39	12.26
1.50	0.0562	2.50	12.82
2.00	0.0749	2.61	13.38
2.67	0.100	2.75	14.12
5.00	0.1873	3.26	16.70

Note :

Constant residence time simulations carried out for residence time of 26.7 seconds

Table 6.1(e) Bubble simulation : no transfer of nitrogen, no
depth effects : low inlet concentration

Water depth (m)	Rise velocity (m/s)	Mean Transfer rate [kmol/s x 10 ¹²]	Mean Transfer flux [kmol/m ² s x 10 ⁸]
0.61	0.1	2.26	
1.00		2.25	11.46
1.50		2.22	11.36
2.00		2.20	11.27
2.67		2.17	11.13
5.00		2.08	10.67

For all simulations with residence time = 26.7 seconds at
different scales, the result was constant:

Mean transfer rate = 2.17×10^{12} kmol/s
Mean transfer flux = 11.14×10^8 kmol/m²s

TABLE 6.2 : RESULTS OF MATHEMATICAL MODELLING SATURATED OXYGEN CONCENTRATION

Flow rate (m ³ /s)	Speed (RPM)	Estimated k _L a (s ⁻¹)	Estimated hold-up	Calculated oxygen conc (H = 2.67m)	Calculated oxygen conc (H = 2m)	Measured concentration
0.041	85.3	0.026	0.0334	12.14	11.77	11.8
0.0728	85.3	0.035	0.0535	12.16	-	11.7
0.1422	85.3	0.048	0.0914	12.18	11.81	11.6
0.0643	64.9	0.0244	0.0412	12.16	11.77	11.5
0.0643	100.0	0.0303	0.0522	12.16	11.77	11.8

All oxygen concentrations are expressed in mg/litre

TABLE 6.3 : DISSOLVED OXYGEN MEASUREMENTS IN 2.7M TANK

IMPELLER SPEED	GAS FLOW (m ³ /s)	TEMPERATURE (°C)	D.O (ppm)	PROBE POSITION
85.3	0.0410	12.1	11.8	
	0.0656		11.8	
	0.0728		11.7	Near
	0.0938		11.6	Surface
	0.1261		11.5	
	0.1422		11.6	
64.9	0.0643	12.3	11.5	
74.6			11.6	
80.0			11.6	Near
85.3			11.7	Surface
96.3			11.8	
100			11.8	
Calibration		14.5	10.05	
Initial reading (no gas)		12.0	10.40	Near surface
With gas		12.0	11.75	Near surface
		12.0	12.0	Bottom tan line
No gas (agitator on)		12.0	11.8	Bottom tan line
		12.0	11.8	Near surface

TABLE 6.4

0.61M DATA RUSHTON TURBINE : TAP WATER
MASS TRANSFER DATA

IMPELLER SPEED (REV/S)	GAS FLOW RATE (m ³ /s)	POWER INPUT W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	k _L a (s ⁻¹)
5.50	0.0015	1.207	0.0051	0.0280
	0.0021	0.963	0.0072	0.0322
				0.0342
	0.0029	0.880	0.0099	0.0356
	0.0035	0.829	0.0120	0.0422
	0.0043	0.784	0.0147	0.0430
	0.0049	0.726	0.0168	0.0473
5.75				0.0451
	0.0015	1.347	0.0051	0.0307
				0.0309
	0.0021	1.121	0.0072	0.0358
				0.0359
	0.0029	1.024	0.0099	0.0398
				0.0388
3.5	0.0035	0.957	0.0120	0.0430
				0.0438
	0.0043	0.908	0.0147	0.0476
				0.0493
	0.0049	0.867	0.0168	0.0515
				0.0498
	0.0015	0.272	0.0051	0.0155
4.5				0.0151
	0.0021	0.235	0.0072	0.0184
	0.0029	0.225	0.0099	0.0214
	0.0035	0.209	0.0120	0.0212
				0.0225
	0.0043	0.226	0.0147	0.0229
				0.0240
5.00	0.0015	0.634	0.0051	0.0210
				0.0223
	0.0021	0.515	0.0072	0.0239
				0.0253
	0.0029	0.475	0.0099	0.0276
				0.0278
	0.0035	0.443	0.0120	0.0310
5.00				0.0306
	0.0043	0.412	0.0147	0.0327
				0.0327
	0.0049	0.378	0.0168	0.0343
				0.0350
	0.0015	0.917	0.0051	0.0239
				0.0254
5.00	0.0021	0.715	0.0072	0.0297
				0.0270
	0.0029	0.661	0.0099	0.0321
	0.0035	0.612	0.0120	0.0374
				0.0349
	0.0043	0.571	0.0147	0.0395
				0.0415
	0.0049	0.535	0.0168	0.0449

TABLE 6.5 : MASS TRANSFER DATA

RUSHTON TURBINE, T/3, 0.61 M TANK, ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	k _L a (s ⁻¹)
4.5	0.001375	0.551	0.00471	0.0436
4.5	0.00206	0.451	0.00706	0.0507
4.5	0.00275	0.412	0.00941	0.0549
4.5	0.00344	0.384	0.01176	0.0611
4.5	0.00413	0.347	0.01411	0.0605
5.5	0.00138	1.212	0.00411	0.0595
5.5	0.00206	0.933	0.00706	0.0686
5.5	0.00275	0.835	0.00941	0.0774
5.5	0.00344	0.788	0.01176	0.0854
5.5	0.00413	0.746	0.01411	0.0874
5.75	0.001375	1.315	0.00471	0.0633
5.75	0.00206	1.066	0.00706	0.0706
5.75	0.00275	0.951	0.00941	0.0768
5.75	0.00344	0.891	0.01176	0.0858
5.75	0.00413	0.848	0.01411	0.0909
4.65	0.00138	0.639	0.00471	0.0489
4.12	0.00138	0.417	0.00471	0.0402

TABLE 6.6 : MASS TRANSFER DATA

CONCAVE TURBINE, 0.61 M TANK, ELECTROLYTE SOLUTION

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	k _L a (s ⁻¹)
5.75	0.001375	1.301	0.00471	0.0707
5.75	0.00206	1.024	0.00706	0.0786
5.75	0.00275	0.926	0.00941	0.0818
5.75	0.00344	0.841	0.01176	0.0899
5.5	0.00138	1.145	0.00471	0.0578
5.5	0.00206	0.898	0.00706	0.0721
5.5	0.00275	0.807	0.00941	0.0800
5.5	0.00344	0.760	0.01176	0.0904
5.5	0.00413	0.711	0.01411	0.0887
5.0	0.001375	0.859	0.00471	0.0565
5.0	0.00206	0.674	0.00706	0.0630
5.0	0.00275	0.601	0.00941	0.0689
5.0	0.00344	0.562	0.01176	0.0762
5.0	0.00413	0.547	0.01411	0.0814
4.5	0.001375	0.615	0.00471	0.0492
4.5	0.00206	0.490	0.00706	0.0576
4.5	0.00275	0.459	0.00941	0.0599
4.5	0.00344	0.408	0.01176	0.0663
4.5	0.00413	0.381	0.01411	0.0682
3.5	0.00138	0.273	0.00471	0.0317
3.5	0.00206	0.233	0.00706	0.0332
3.5	0.00275	0.222	0.00941	0.0354
3.5	0.00344	0.201	0.01176	0.0380

TABLE 6.7

2.67M TANK : RUSHTON TURBINE : TAP WATER
MASS TRANSFER DATA

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	k _L a (s ⁻¹)
1.25	0.0664	0.241	0.0119	0.0302 0.0295 0.0310 0.0277 0.0314
1.42	0.0664	0.343	0.0119	0.0294
1.58	0.0664	0.476	0.0119	0.0328 0.0330 0.0341 0.0327
1.58	0.0880	0.402	0.0153	0.0406 0.0375
1.67	0.0664	0.545	0.0119	0.0329 0.0342
	0.0880	0.494	0.0153	0.0410 0.0405
	0.1110	0.466	0.0199	0.0410 0.0412 0.0424
1.67	0.1338	0.446	0.0240	0.0414 0.0462

TABLE 6.8

2.67M TANK CONCAVE TURBINE : TAP WATER
MASS TRANSFER DATA

IMPELLER SPEED (REV/SEC)	GAS FLOW RATE (m ³ /s)	POWER INPUT (W/kg)	SUPERFICIAL GAS VELOCITY (m/s)	k _L a (s ⁻¹)
1.25	0.0664	0.251	0.0119	0.0272
1.42	0.0664	0.368	0.0119	0.0298
				0.0279
				0.0281
1.58	0.0664	0.536	0.0119	0.0311
				0.0317
1.58	0.1110	0.421	0.0199	0.0370
				0.0425
1.67	0.0664	0.638	0.0119	0.0344
				0.0321
				0.0339
	0.1110	0.490	0.0199	0.0434
				0.0454
	0.1339	0.469	0.0240	0.0452
				0.0461
	0.1580	0.445	0.0283	0.0515
	0.1855	0.445	0.0332	0.0537

TABLE 6.9 : MASS TRANSFER CORRELATIONS - 0.61 M TANK, ELECTROLYTE SOLUTION

$$k_L a = \alpha (\bar{\epsilon}_m)^\beta v_{SG}^\gamma$$

IMPELLER	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT
Rushton (T/3)	0.83	0.42	0.50	2.7	7.6	0.979
Concave (T/3)	1.21	0.54	0.56	4.5	11.7	0.978
Combined Data	1.01	0.48	0.53	5.1	26.1	0.949

TABLE 6.10

$k_L a$ correlations based on:

$$k_L a = \alpha (\bar{\epsilon}_m)^\beta (v_{SG})^\gamma$$

TAP WATER

IMPELLER	TANK DIAMETER (m)	α	β	γ	CORRELATION COEFFICIENT	MEAN ERROR	MAX ERROR
	0.61	0.61	0.46	0.59	0.982	3.2	10.6
RUSHTON TURBINE	2.67	0.26	0.18	0.43	0.897	3.8	9.1
	COMBINED	0.502	0.39	0.55	0.934	5.3	25.6
CONCAVE TURBINE	2.67	0.49	0.27	0.58	0.966	3.4	8.5
RUSHTON & CONCAVE	2.67	0.351	0.17	0.51	0.919	4.4	10.6

**TABLE 6.11 : MASS TRANSFER CORRELATIONS - 0.61m TANK,
ELECTROLYTE SOLUTION**

$$k_L a = \alpha N^\beta Q^\gamma$$

IMPELLER	α	β	γ	MEAN ERROR (%)	MAXIMUM ERROR (%)	CORRELATION COEFFICIENT
Rushton (T/3)	0.042	1.47	0.33	2.6	5.7	0.979
Concave (T/3)	0.028	1.61	0.36	4.3	10.2	0.971
Combined data	0.061	1.15	0.42	5.3	24.3	0.935

TABLE 6.12

$k_L a$ correlations based on:

$$k_L a = \alpha N^\beta Q^\gamma$$

TAP WATER

IMPELLER	TANK DIAMETER (m)	α	β	γ	CORRELATION COEFFICIENT	MEAN ERROR	MAXIMUM ERROR
RUSHTON	0.61	0.042	1.41	0.43	0.985	2.8	8.8
CONCAVE	2.67	0.076	0.89	0.47	0.969	3.2	8.8
RUSHTON	2.67	0.073	0.52	0.37	0.901	3.7	8.8
RUSHTON	COMBINED	0.061	1.15	0.42	0.935	5.3	24.3

TABLE 6.13

PATTERN OF PROBLEMS WITH MASS TRANSFER DATA : 2.67 M TANK : RUSHTON TURBINE

FLOW RATE (m^3/s)	0.0664	0.0880	0.110	0.1338	0.1579	0.1854
1.25	YYYYYN	-	-	-	-	-
1.42	Y	NNN	-	Y	-	-
1.58	YYYY	YY	-	N	-	N
1.67	YY	Y	YYY	YY	NNNY	NNY

Y = Run accepted
N = Run rejected due to noise problems

TABLE 6.14

PATTERN OF PROBLEMS WITH MASS TRANSFER DATA : 2.67M TANK : CONCAVE TURBINE

FLOW RATE (m ³ /s)		0.664	0.088	0.111	0.1338	0.1529	0.1854
SPEED (RPS)		1.25	Y?N	NN	-	-	-
1.42	YY	-	?N	NN	-	-	-
1.58	YY	-	YY	??	?N	-	-
1.67	YY	-	YY	YY	Y	Y	Y

Y = Run accepted
 ? = Run accepted (with reservations)
 N = Run rejected

TABLE 7.1 : FLOODING DATA - RUSHTON TURBINE

TANK DIAMETER (M)	FROUDE NUMBER	PREDICTED FLOW NUMBER AT FLOODING	ESTIMATED FLOW NUMBER AT FLOODING
1.83	0.20	0.128	0.148 - 0.163
	0.24	0.154	0.165 - 0.173
	0.42	0.268	0.318 - 0.326
2.67	0.145	0.093	0.100 - 0.113
	0.20	0.128	0.111 - 0.122
	0.178	0.114	0.111 - 0.122

TABLE 7.2 : FLOODING DATA : 2.67m TANK
(Separate Experiments from 7.1)

IMPELLER	FROUDE NUMBER	PREDICTED FLOW NUMBER AT FLOODING	ESTIMATED FLOW NUMBER AT FLOODING
Rushton Turbine	0.182	0.117	0.11 - 0.12
	0.227	0.145	0.14 - 0.17
	0.252	0.162	0.14 - 0.15
Concave turbine	0.182	-	0.14 - 0.16
	0.227	-	0.17 - 0.19
	0.252	-	0.16

TABLE 7.3 : Flooding Data for Pitched Blade Turbines

Pumping Upwards

N_F (RPM)	$Q(m^3/s) \times 10^{-3}$
84	1.455
91	2.885
105	4.488
114	6.150
120	7.700

Table 8.1 : Example of mixer option (Chemineer)

Viscosity = 5000cP
Batch volume = 30000 gallons
Chemscale = 4
Fluid SG = 1
Pitched blade turbine

Prime Mover Power (hp)	Shaft Speed (rpm)	Impeller Diameter (m)
40	84	1.47
30	68	1.57
25	56	1.70
20	37	2.09

TABLE 8.2 : 1983-84 PRICES OF THE MAIN ITEMS IN THE DRIVE TRAIN

Shaft diameter (in)	Gearbox cost (pounds sterling)	Most used motor cost (pounds sterling)	Seal cost (pounds sterling)	Shaft cost per unit length (pounds sterling /ft)
2	1631	74	543	42
2½	2471	98	823	74
3	2793	155	931	105
3½	3381	167	1127	133
4	4046	526	1348	168
4½	4725	711	1575	200
5	8491	711	2831	231
5½	10710	1117	3570	255
6	12936	1117	4312	294

TABLE 8.3 : PRESENT VALUES OF MIXER OPTIONS

i = 0.1
N = 10 years
u_c = 0.05 £/kWh
U_T = 5000 hours/year

D/T	Capital cost (£)	Running cost (£)	Total cost (£)
0.2	6300	78000	84300
0.3	8200	42500	50700
0.4	10000	27600	37600
0.5	11700	19700	31400
0.6	13200	15000	28200

TABLE 8.4 : EQUIVALENT ANNUAL COSTS OF MIXER OPTIONS

$i = 0.1$
 $N = 10 \text{ years}$
 $u_c = 0.05 \text{ £/kWh}$
 $U_T = 5000 \text{ hr/yr}$

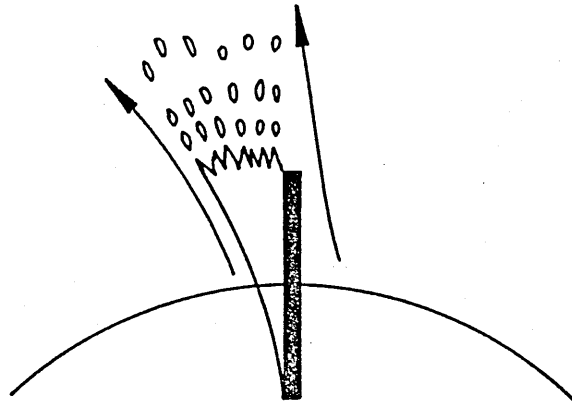
D/T	Equivalent annual capital cost (£)	Annual running cost (£)	Total cost £/year
0.2	1020	12700	13720
0.3	1340	6900	8240
0.4	1630	4500	6130
0.5	1910	3200	5110
0.6	2150	2400	4550

TABLE 9.1 : Comparison of Agitator Types

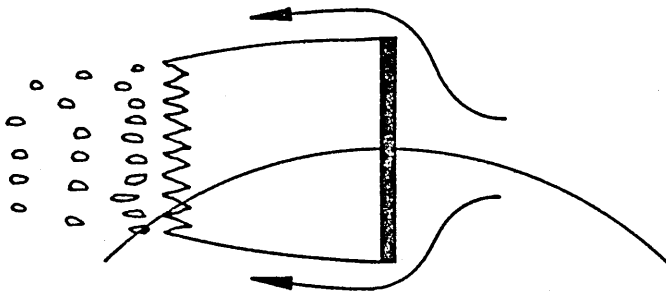
IMPELLER	SPEED* (RPM)	P_g/P_u	ϵ_G (calc)	ϵ_G **
Rushton T/3	270	0.425	0.0666	0.0680
Rushton .47T	150	0.568	0.0659	0.0744
Concave T/3	300	0.621	0.0699	0.0594
6PBT DOWN	180	0.436	0.0605	0.0578
6PBT UP	180	0.867	0.0680	0.0645

*Closest speed to that for $P_u = 1.0$ W/kg selected

**Actual measured hold-up



Clinging Cavity

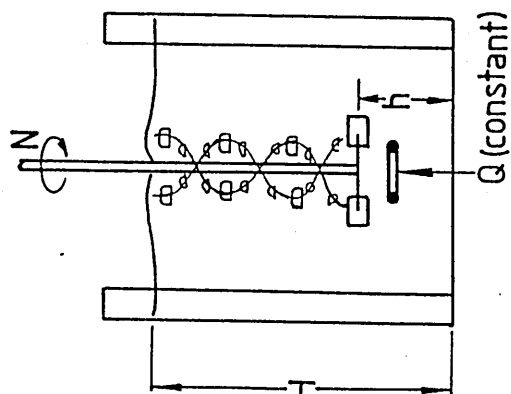


Large Cavity

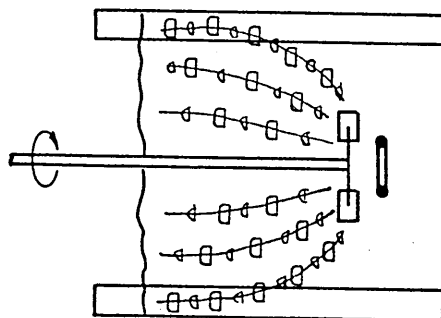
Increasing N →

FLOODED

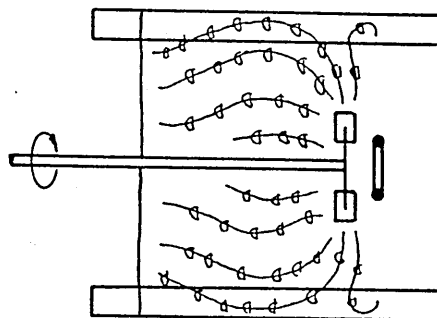
RECIRCULATING



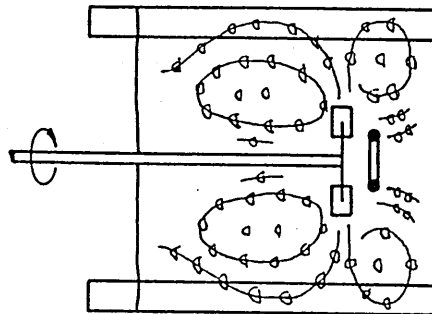
(a)



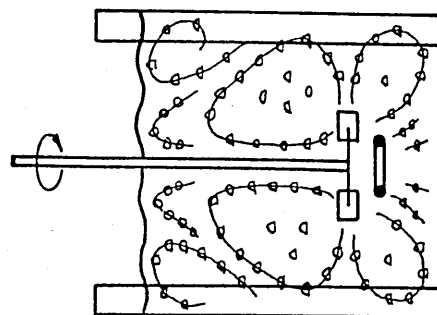
(b)



(c)

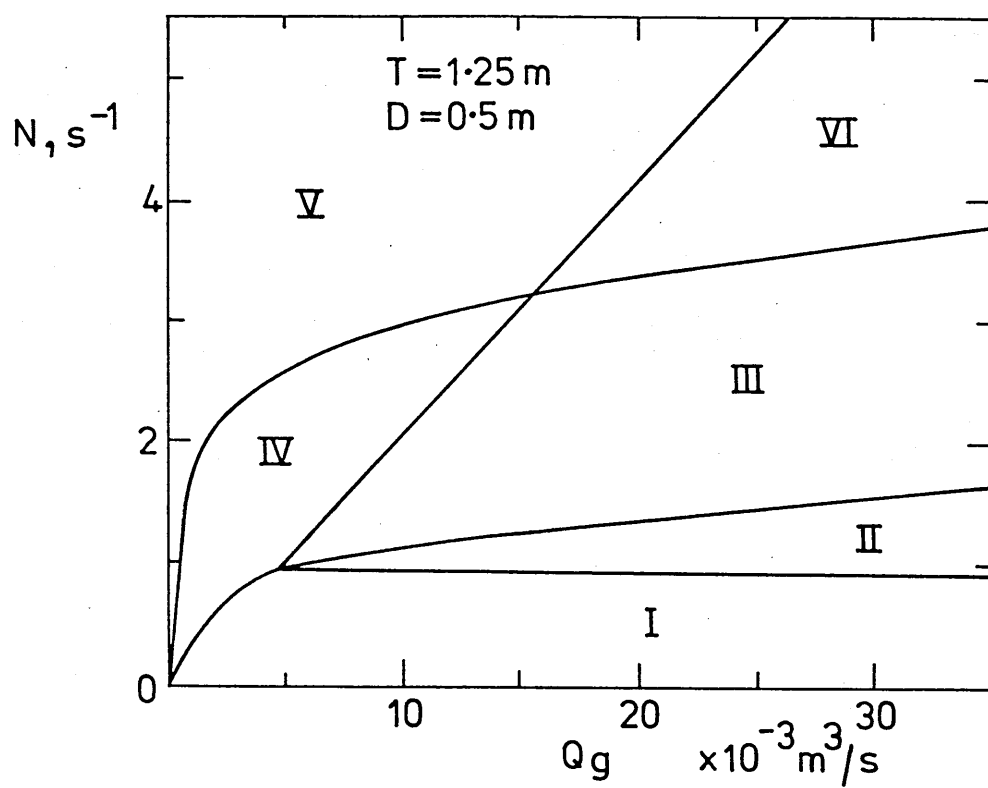


(d)



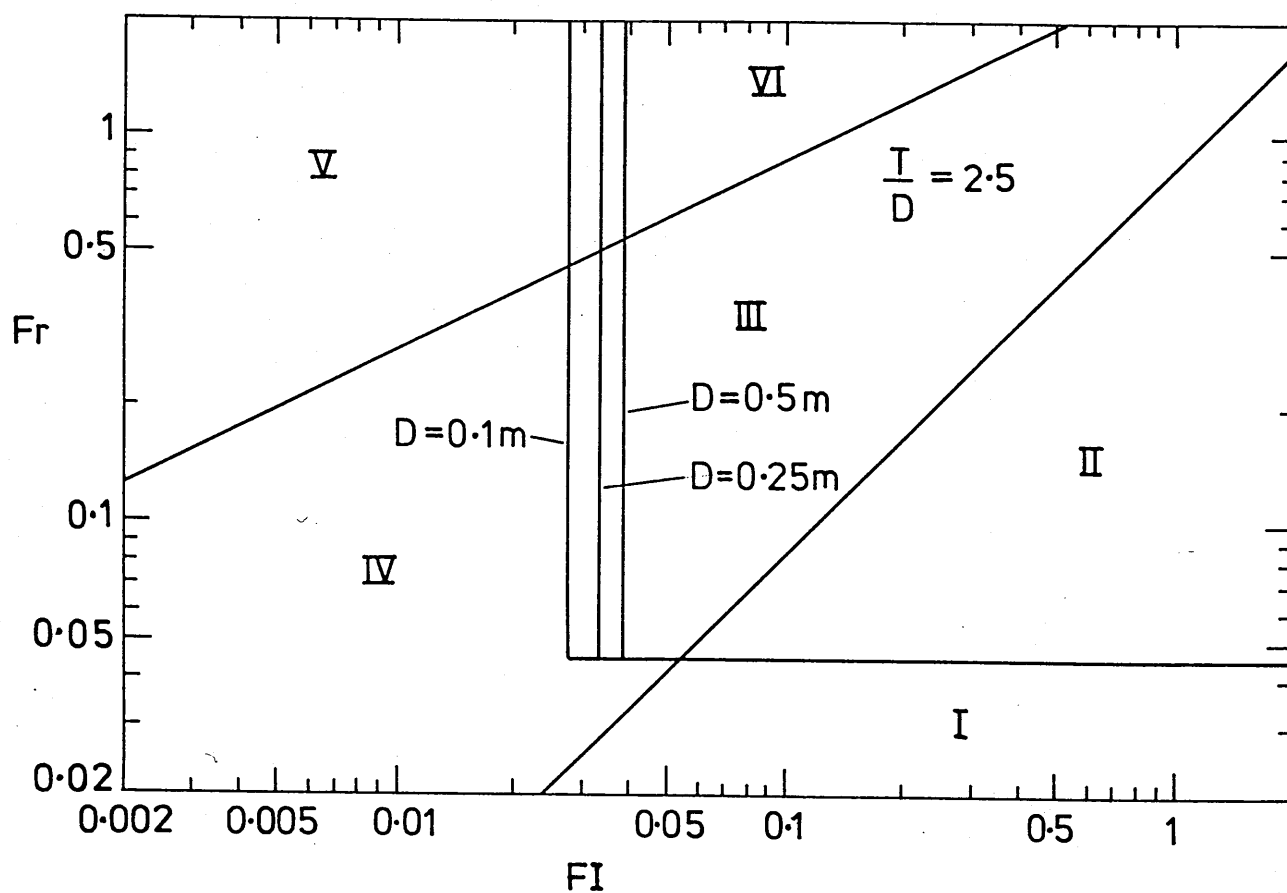
(e)

2.2 Gas liquid flow patterns: disc turbine



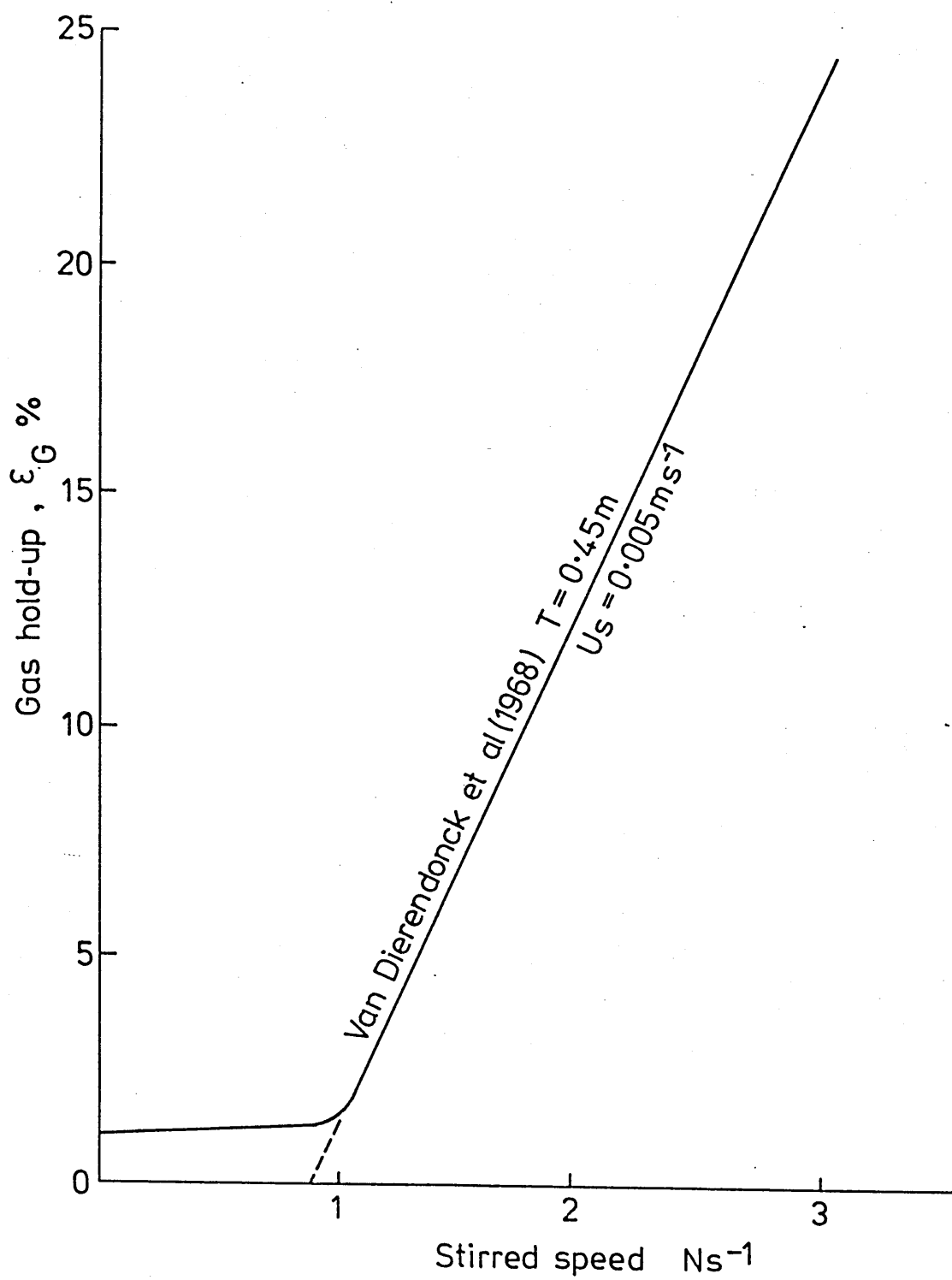
2.3

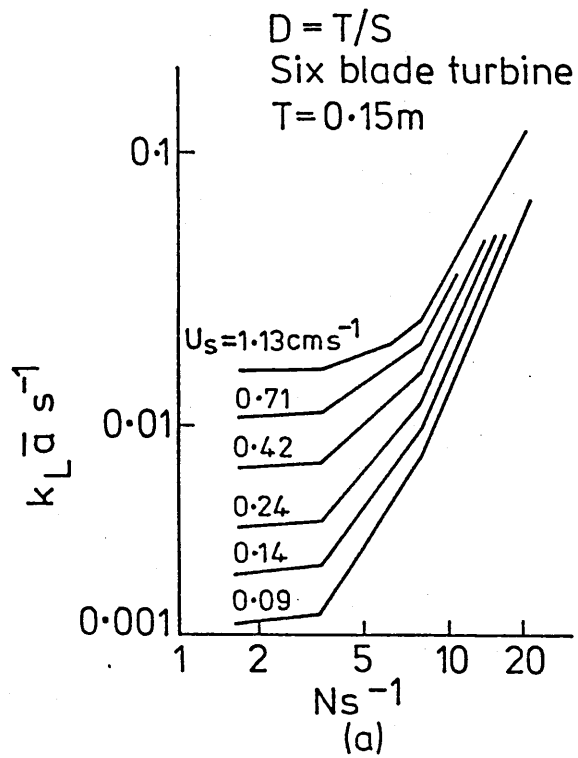
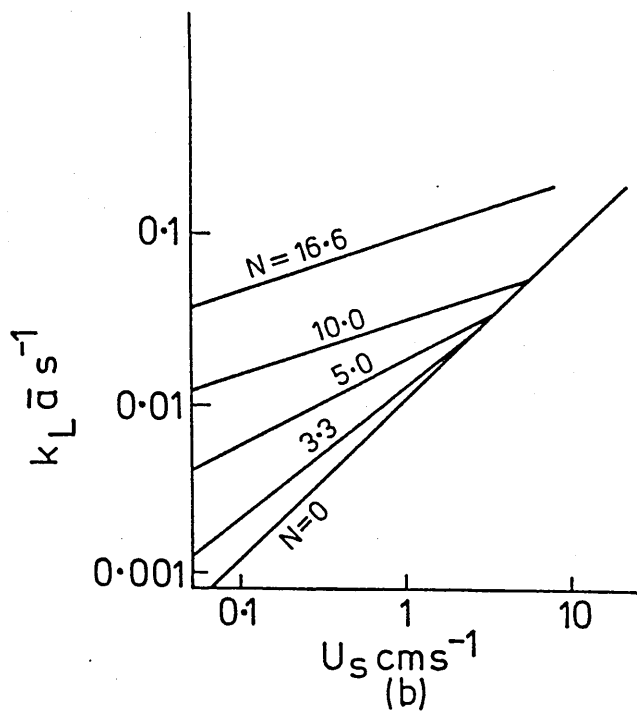
Dimensional flow map for disc turbine



2.4

Dimensionless flow map for disc turbine



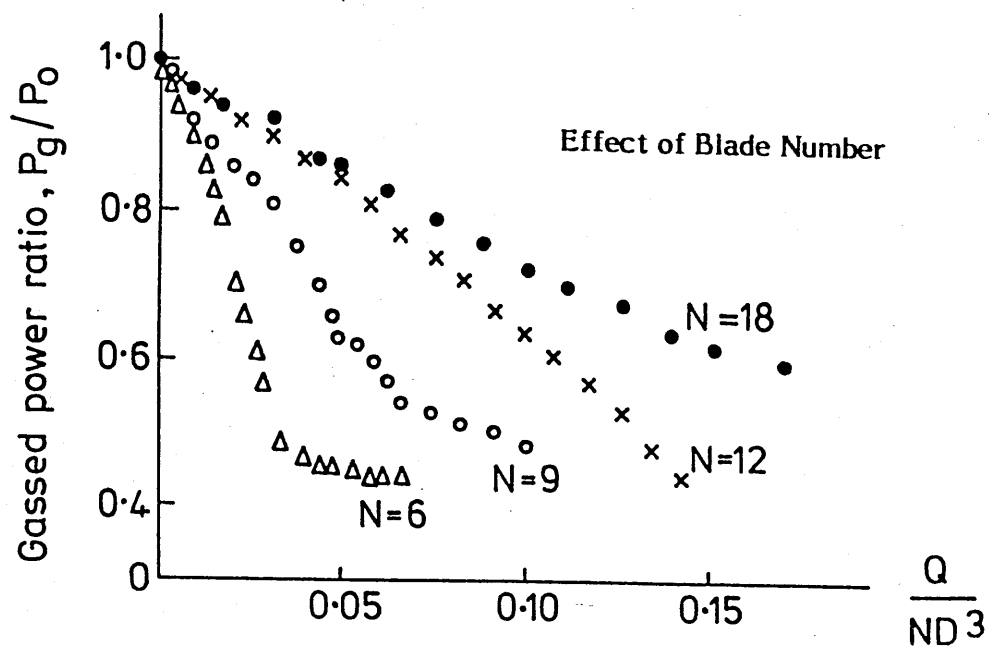
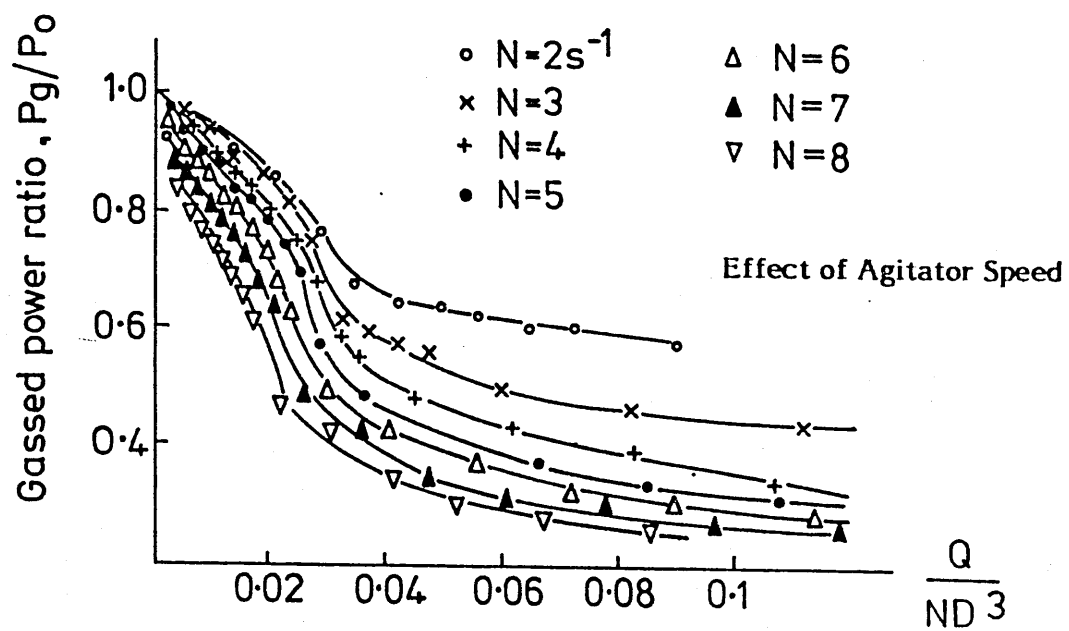


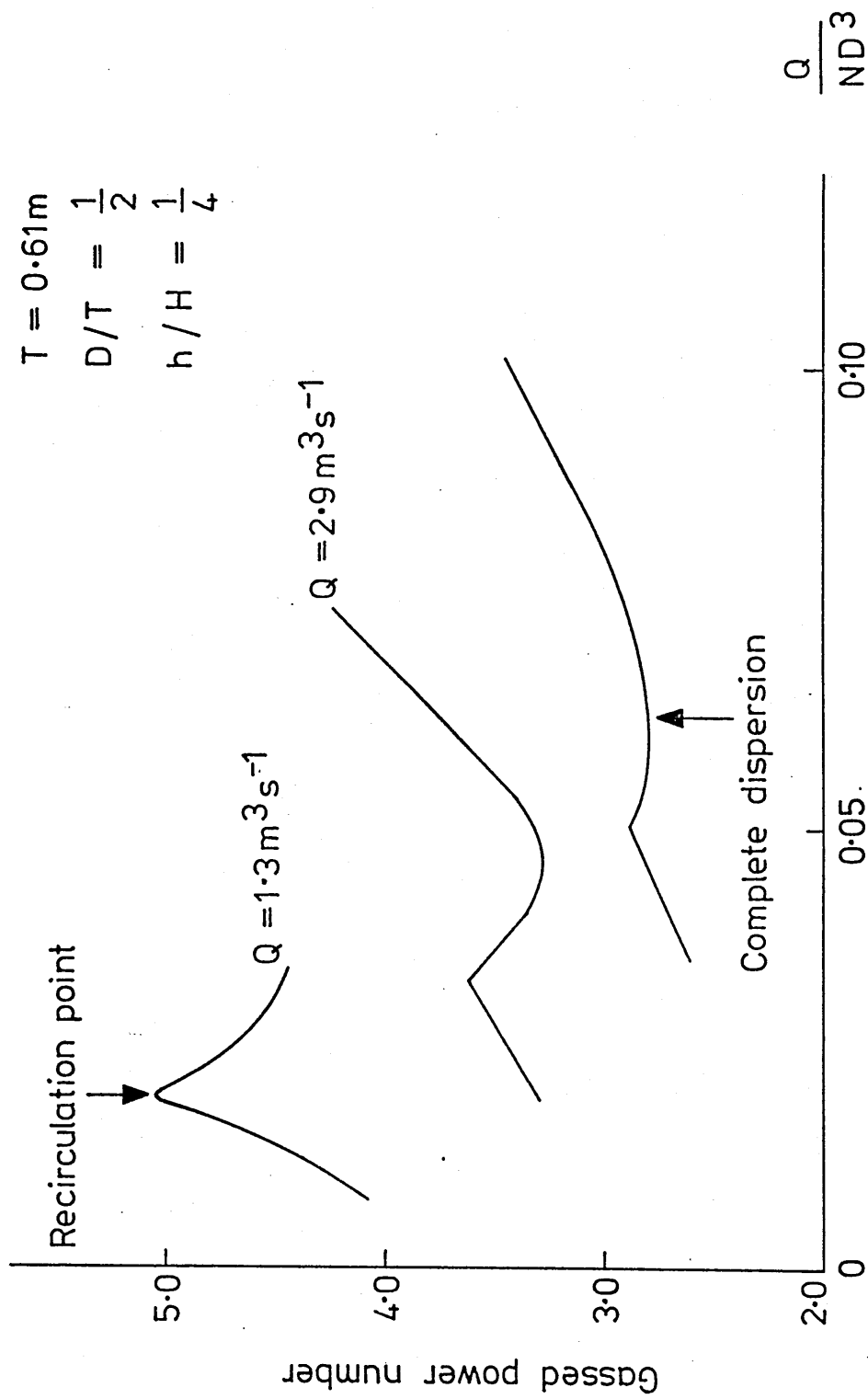
2.6

Mass transfer characteristics of a stirred vessel Nishikawa et al)

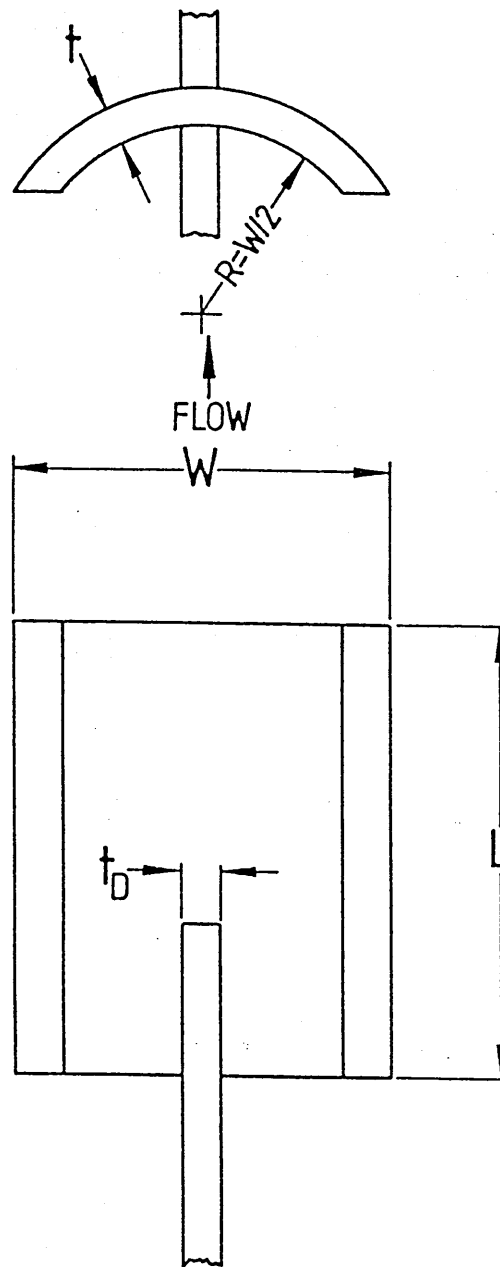
(a) Stirrer speed effect

(b) Superficial gas velocity effect



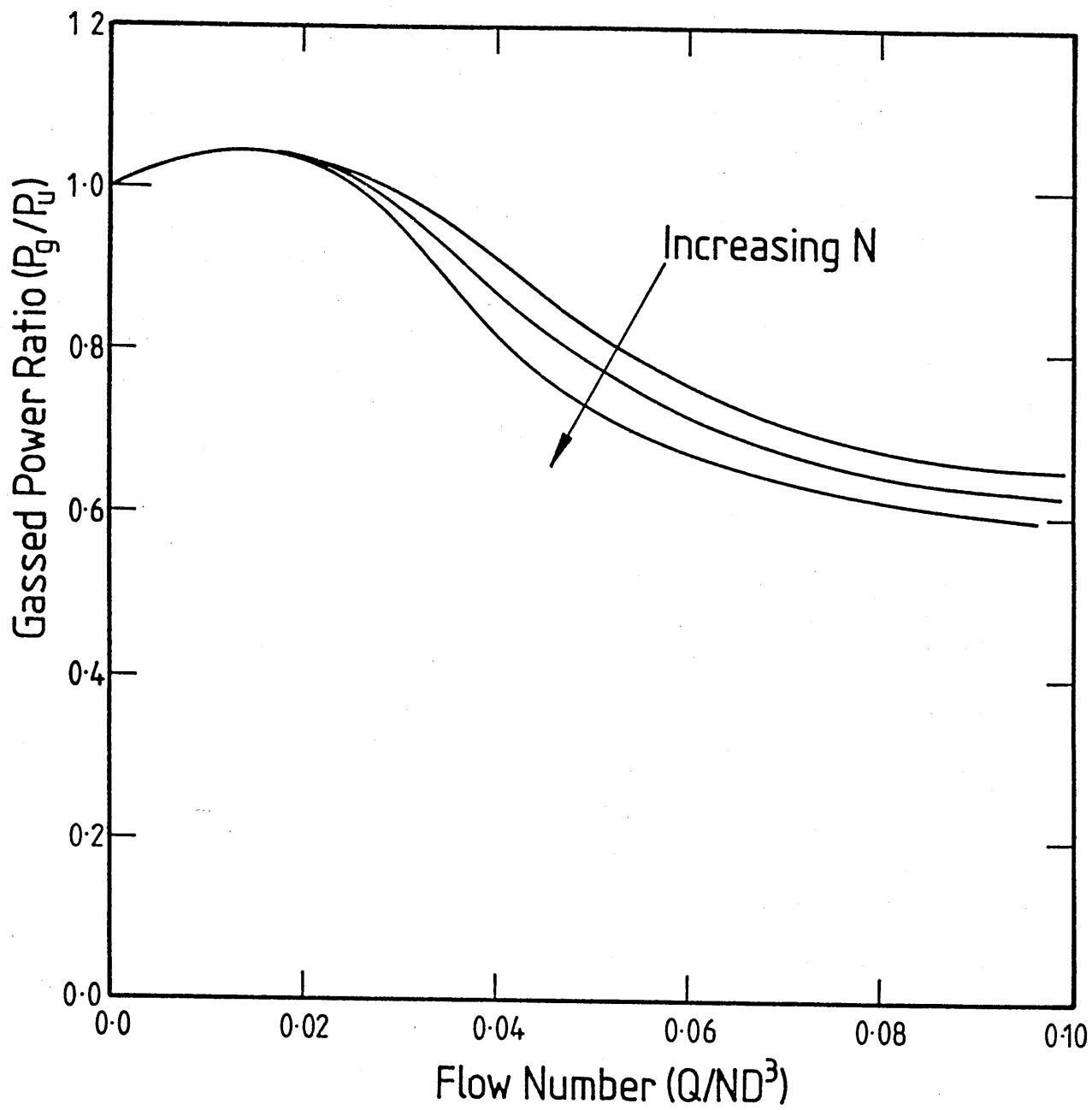


2.8 Gassed power curve at constant gas flow rate



2.9

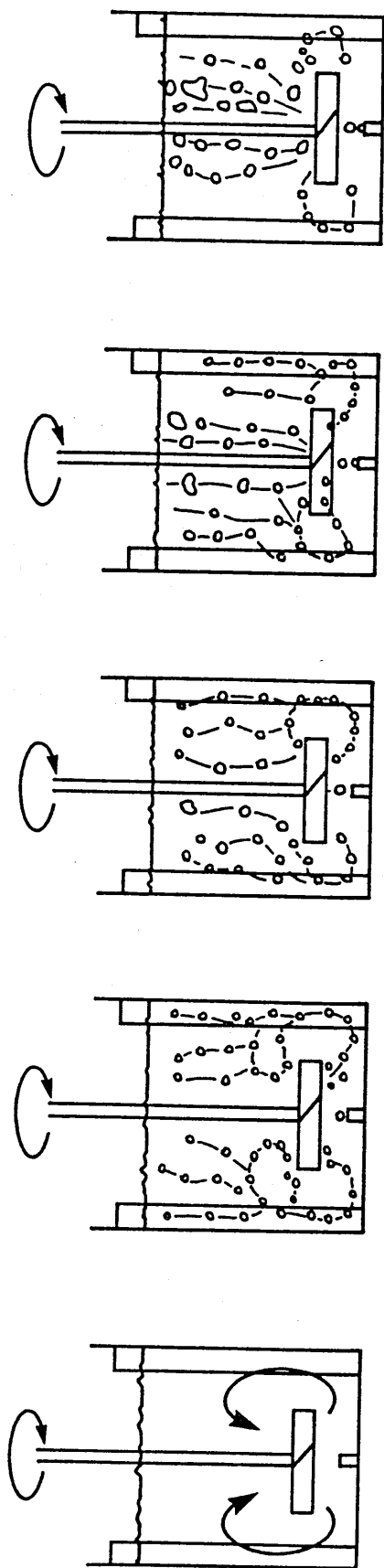
Concave turbine geometry



2.10

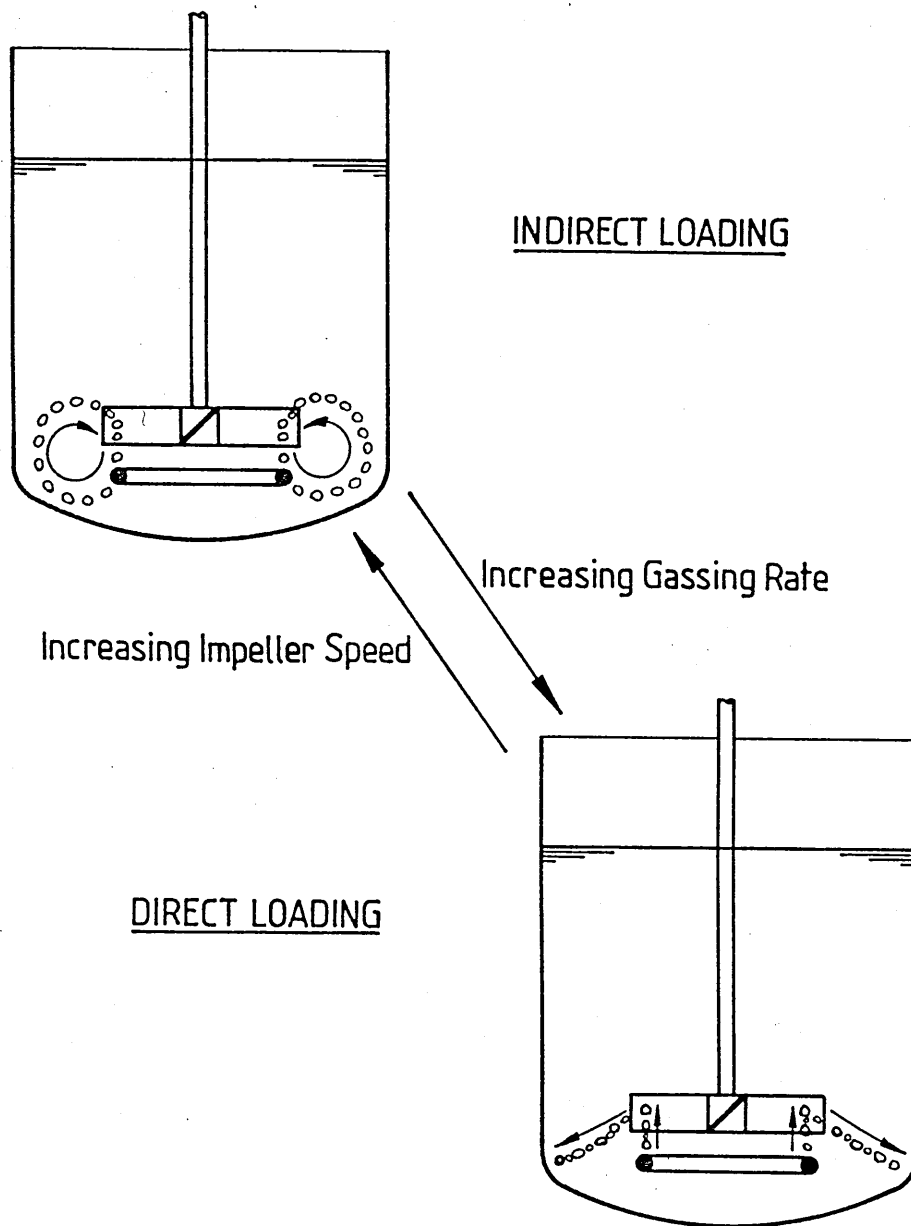
Concave turbine: power curves of Van't Riet, Boom and Smith

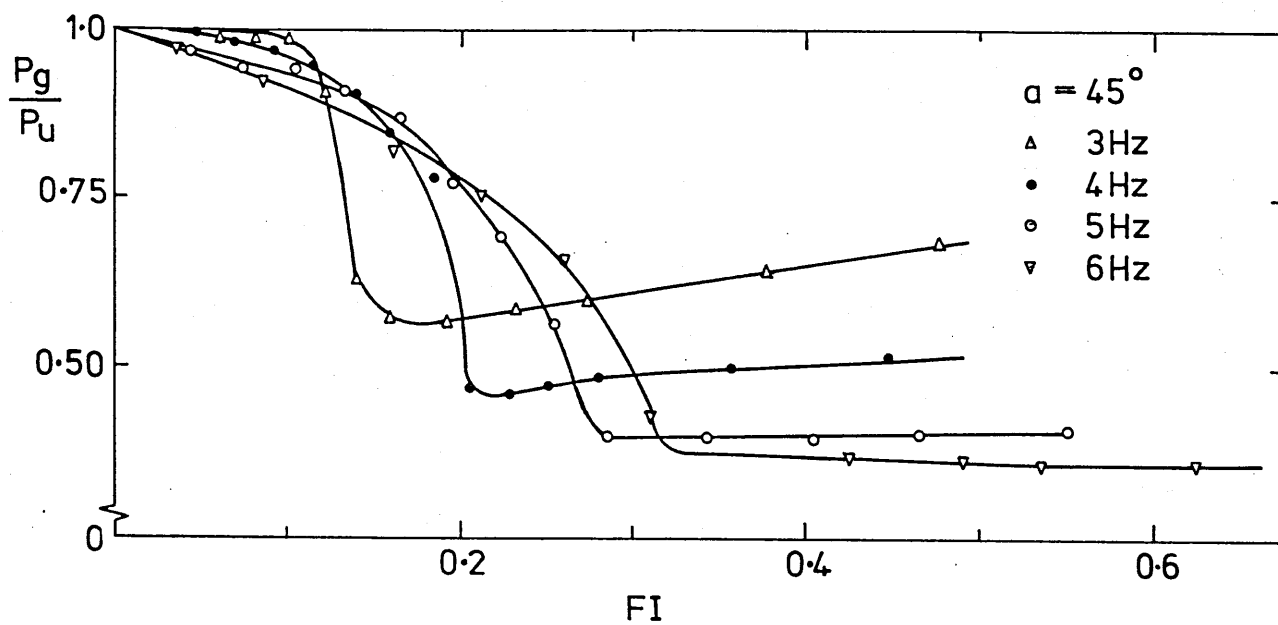
(1976)



Increasing Gas Flow

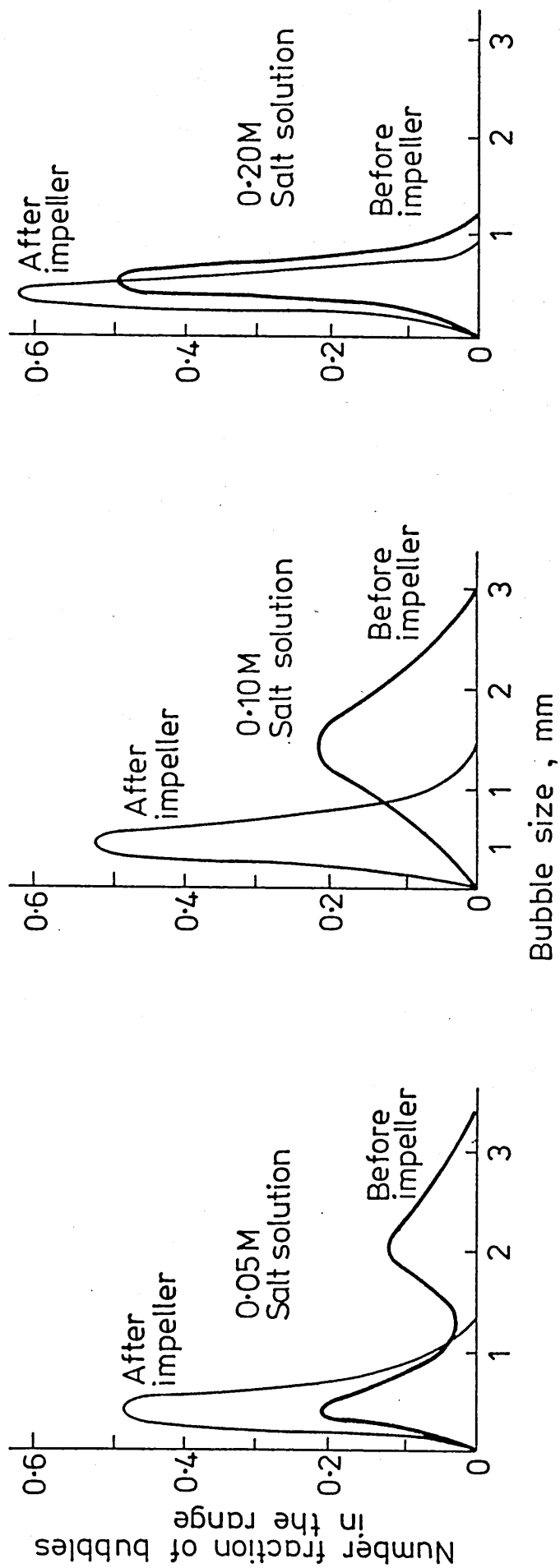
2.11 Gas liquid flow patterns: downwards pumping PBT

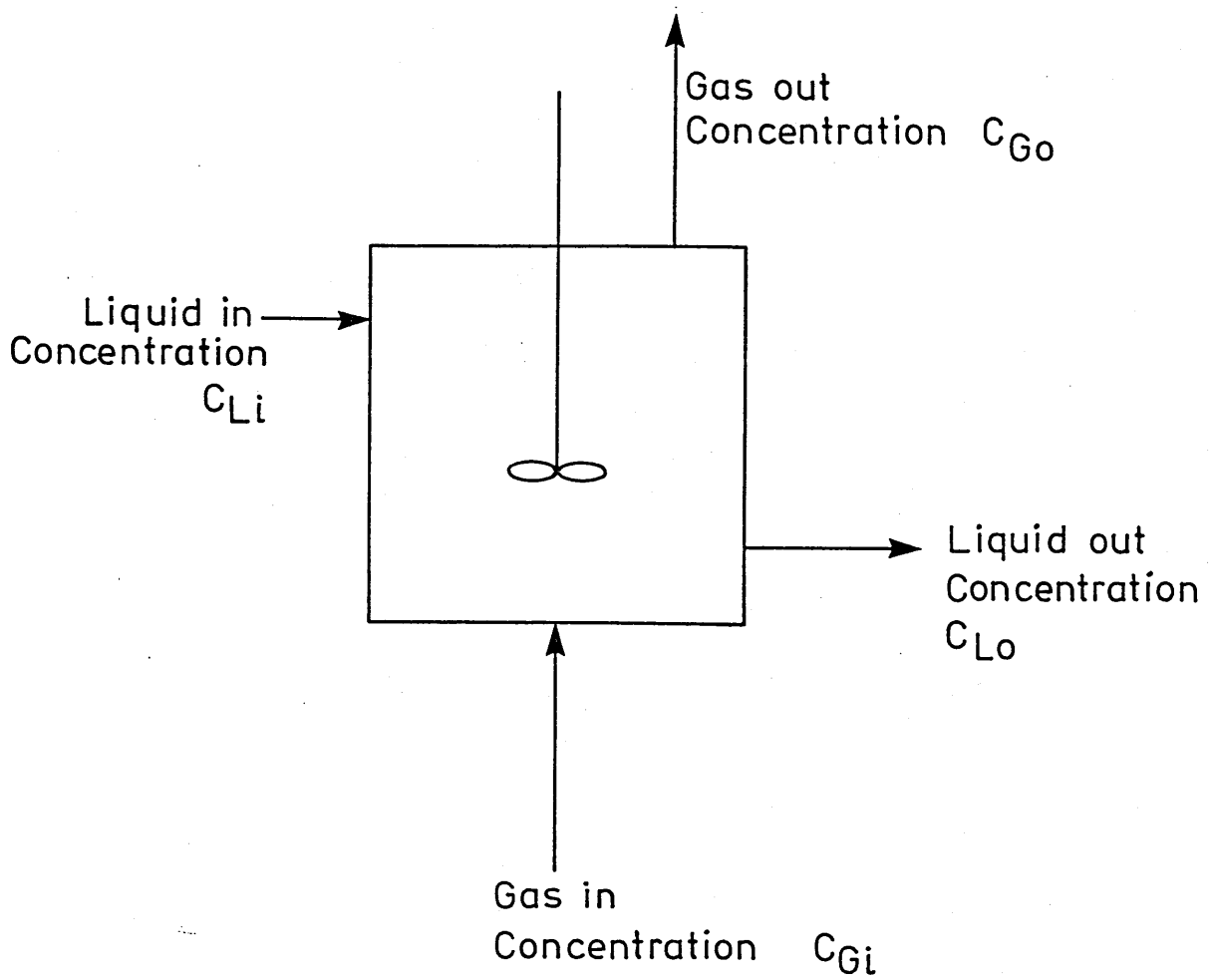


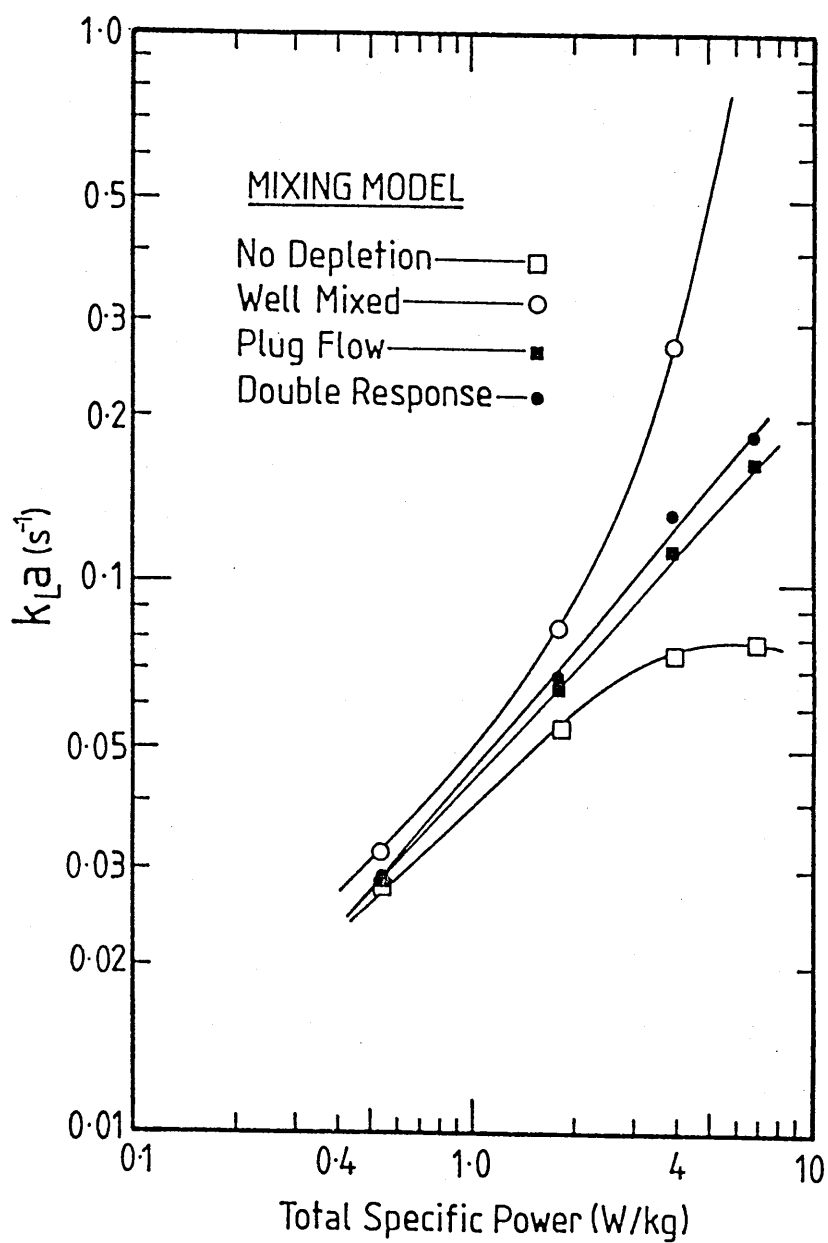


2.13

Gassed power curves: downwards pumping PBT (Warmoeskerken, 1986)

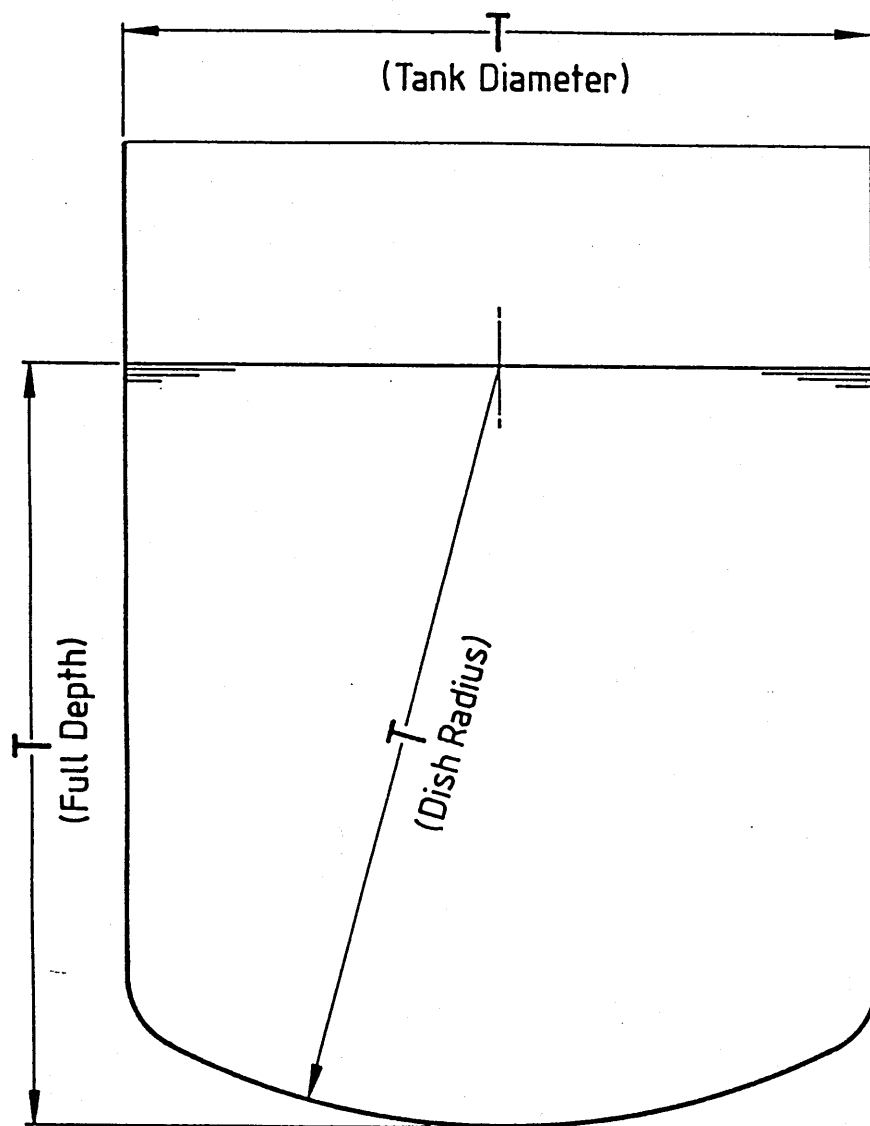






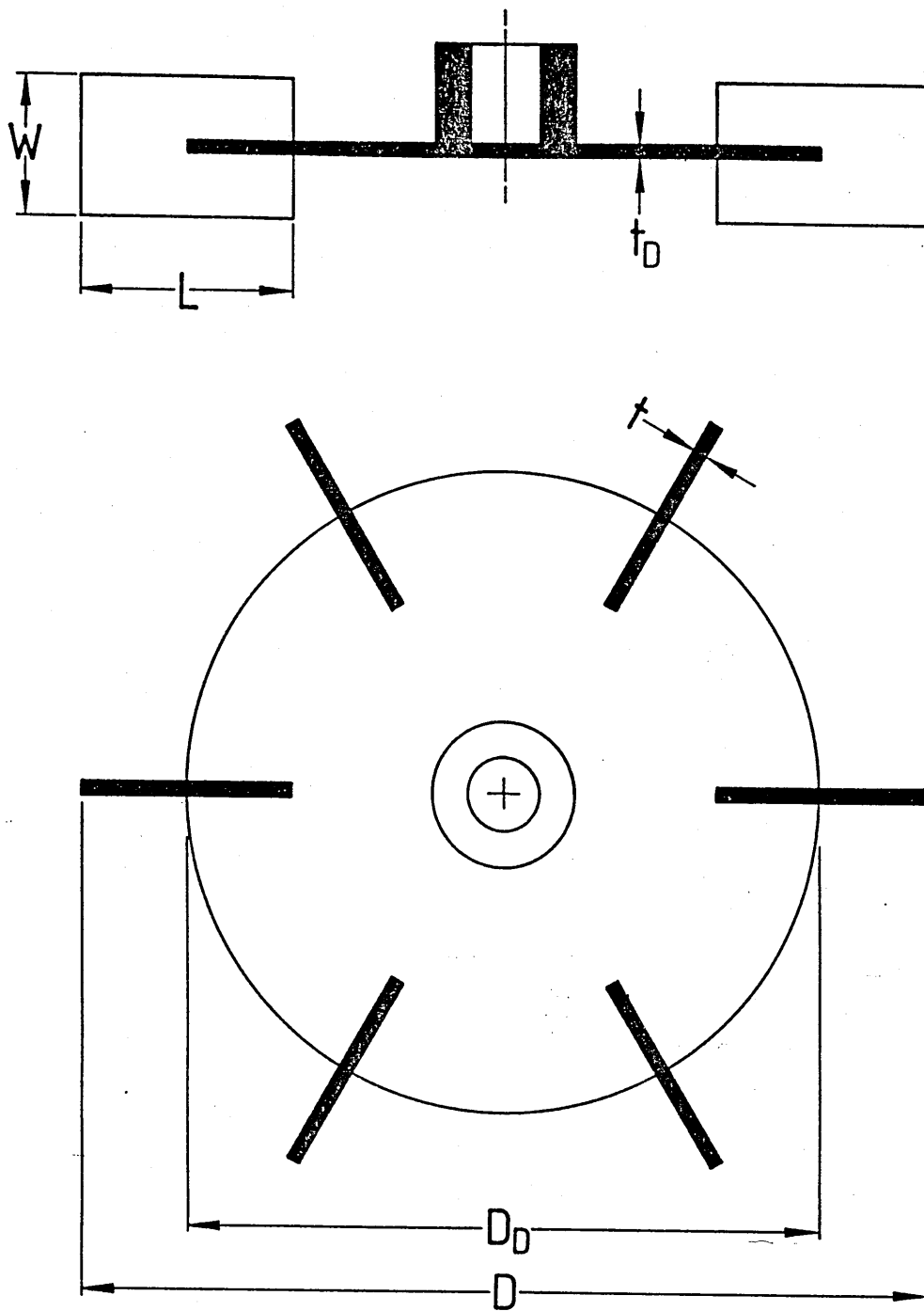
2.16

Effect of assumed gas phase residence time distribution on experimental value of k_La (Davies, 1986)



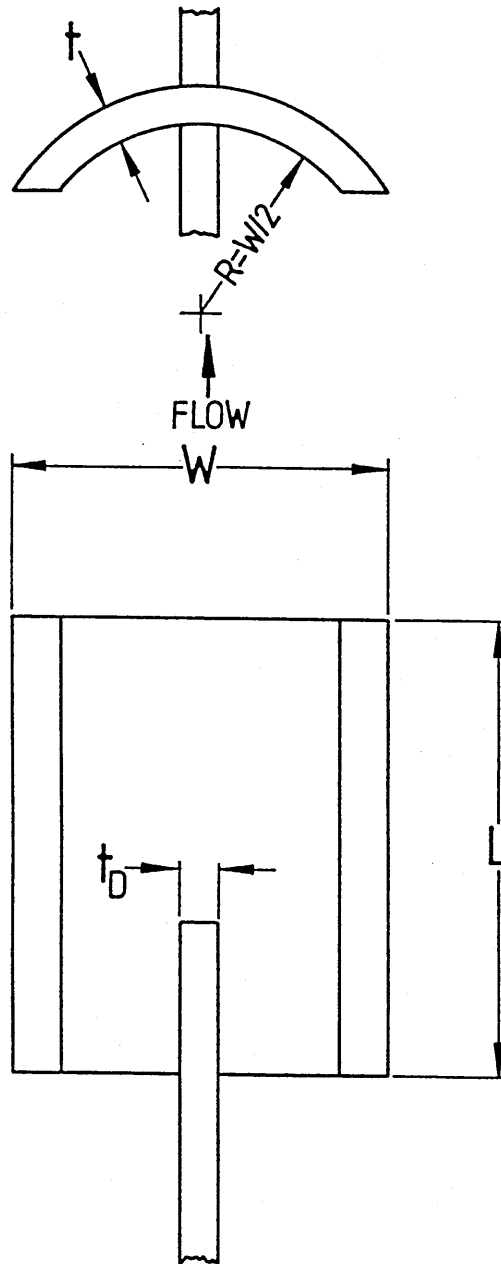
3.1

Standard vessel geometry

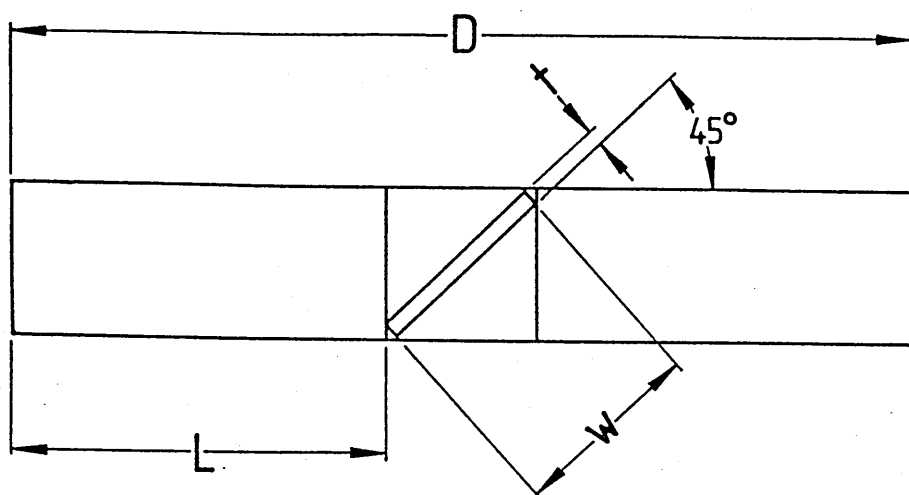


3.2 Rushton turbine geometry

Vessel Diameter (m)	Impeller Diameter, D (m)	Blade Width w (m)	Blade Length, L (m)	Disc Diameter D_d (m)	Blade Thickness, t (m)
0.61	0.2018	0.041	0.0508	0.134	0.0034
0.61	0.282	0.0574	0.0716	0.191	0.003
1.83	0.608	0.124	0.153	0.404	0.0103
2.67	0.885	0.180	0.222	0.599	0.0137

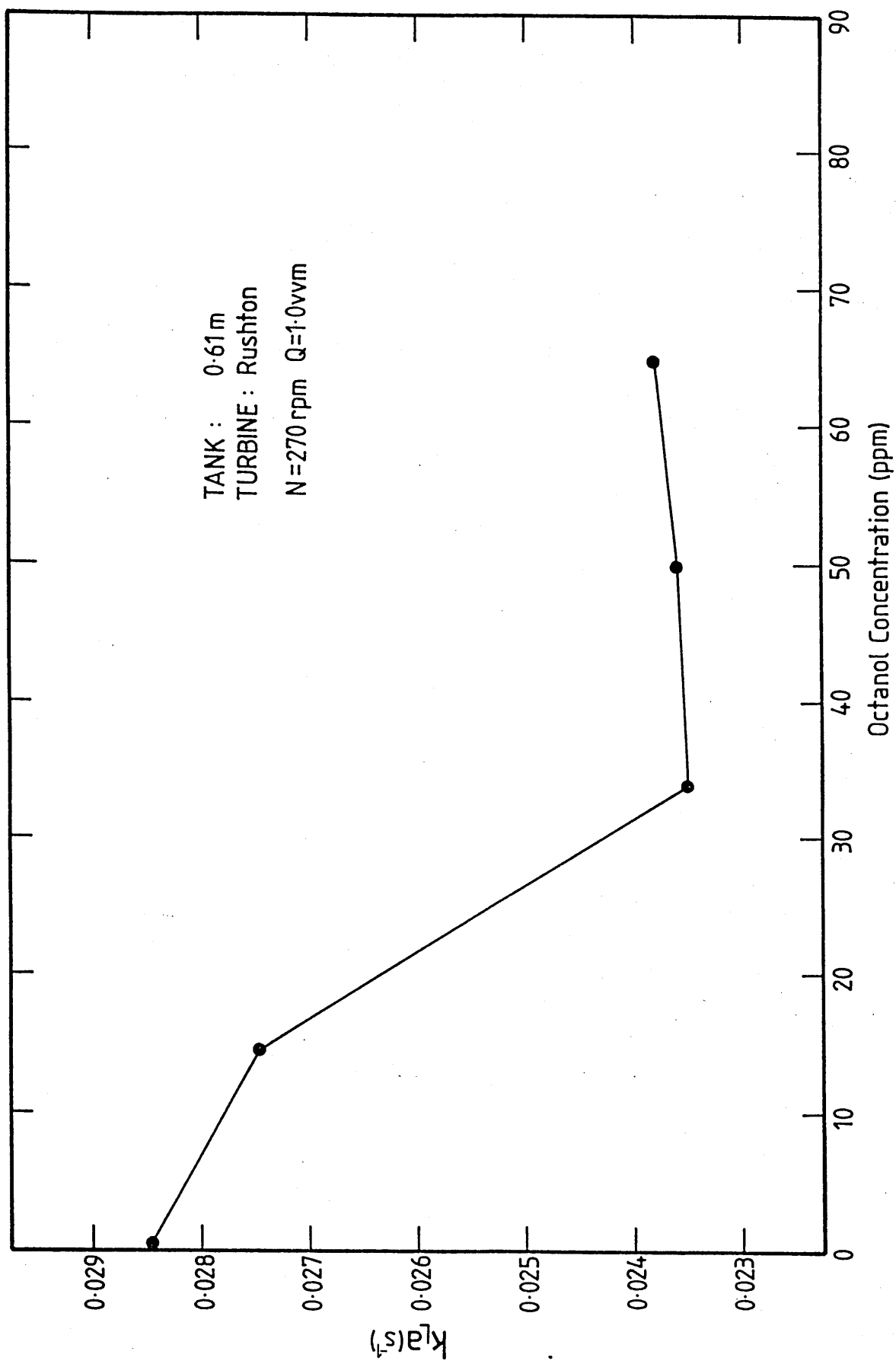


Tank Diameter (m)	Impeller Diameter (m) [D]	Disc Thickness (m) [t_D]	Blade Length (m) [L]	Blade Width (m) [W]	Disc Diameter (m) [D_d]	Blade Thickness (m) [t]
0.61	0.202	0.0032	0.0508	0.0415	0.134	0.0032
1.83	0.6085	0.0095	0.152	0.122	0.404	0.0095
2.67	0.889	0.0128	0.223	0.179	0.595	0.00135

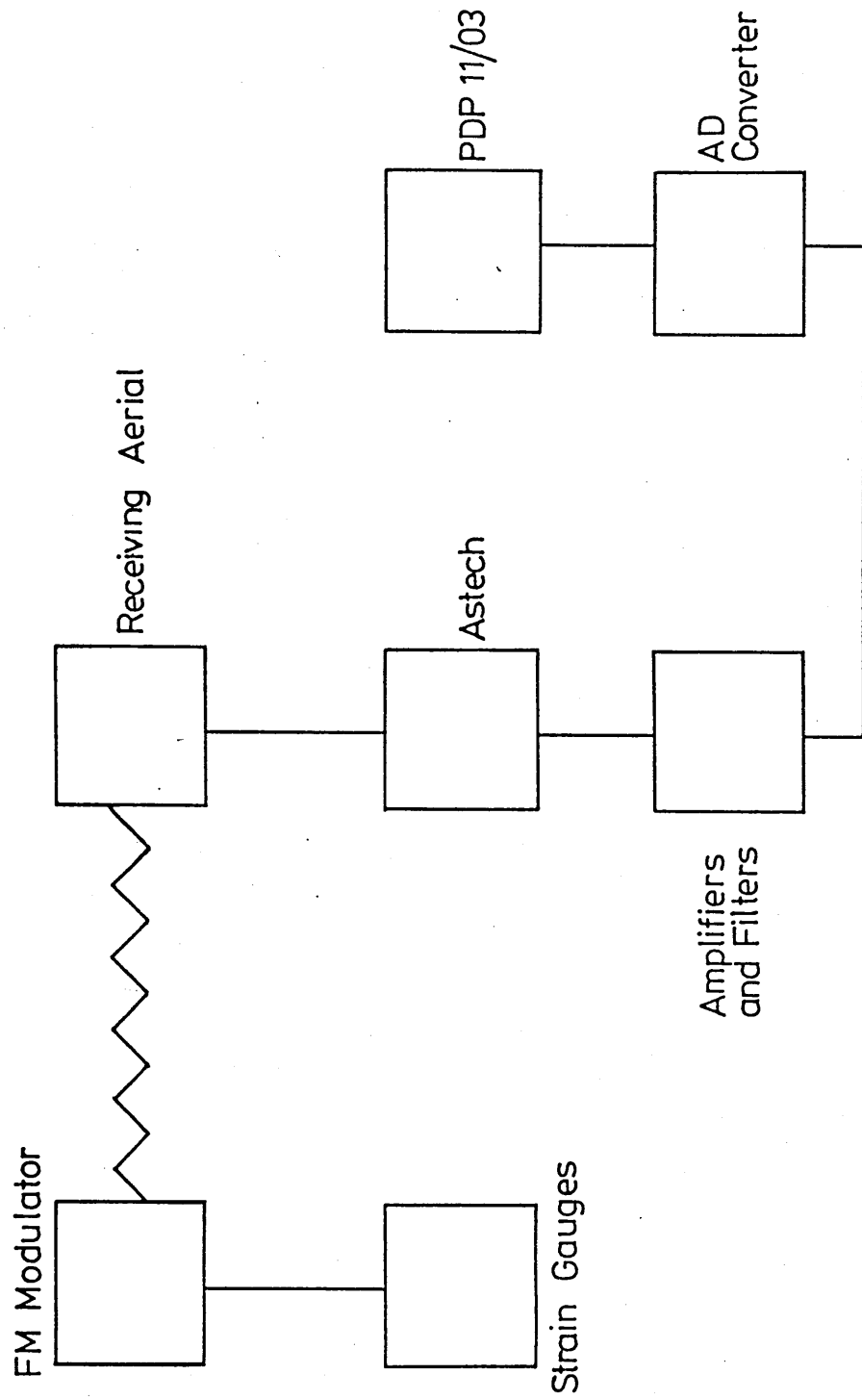


3.4 Pitched blade turbine geometry

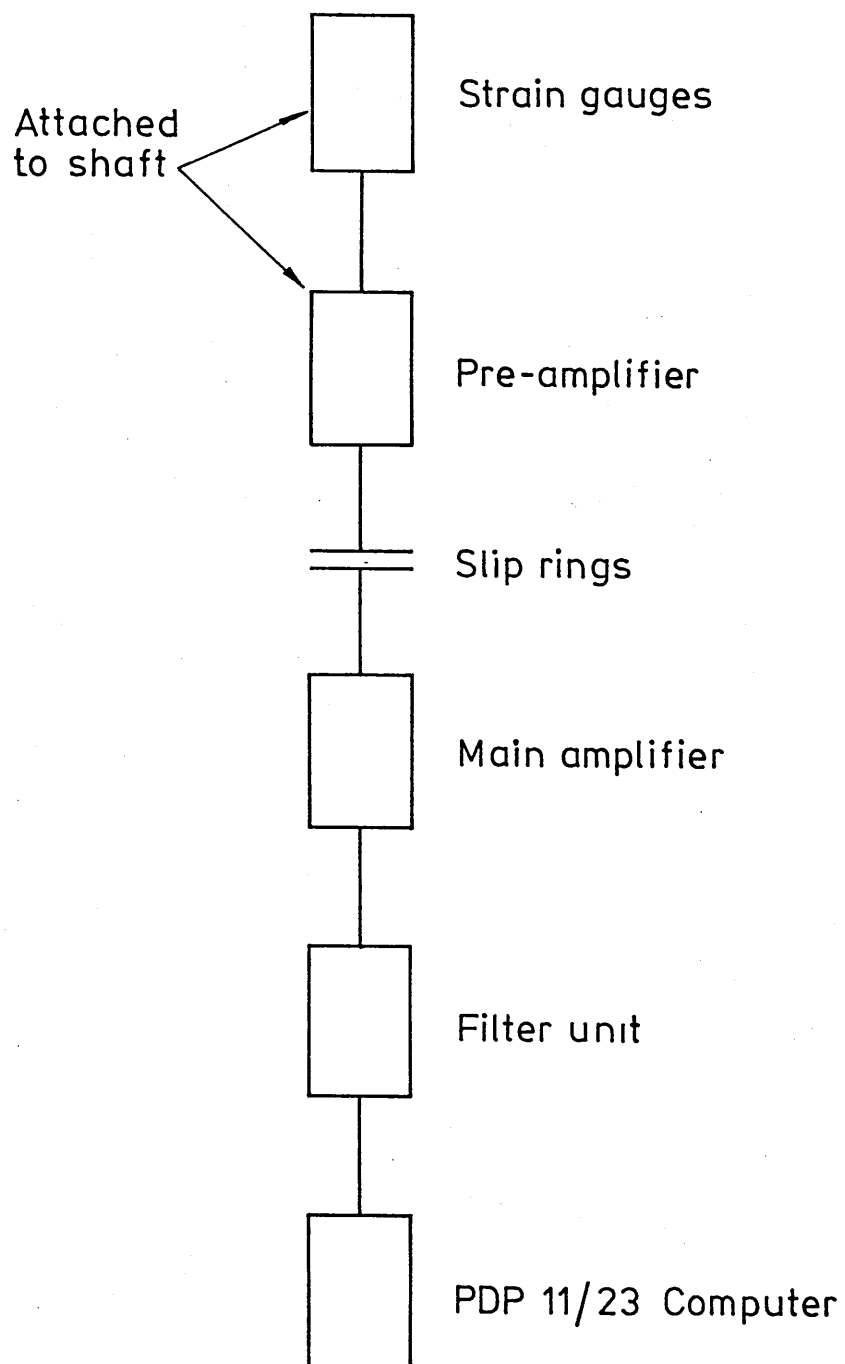
Tank Diameter (m)	Impeller Diameter, D (m)	Blade Length, L (m)	Blade Thickness, t (m)	Blade Width, W (m)	Number of Blades
0.61	0.300	N/A	0.0033	0.086	4
0.61	0.305	N/A	0.0016	0.071	6
2.67	1.33	0.544	0.0174	0.378	4

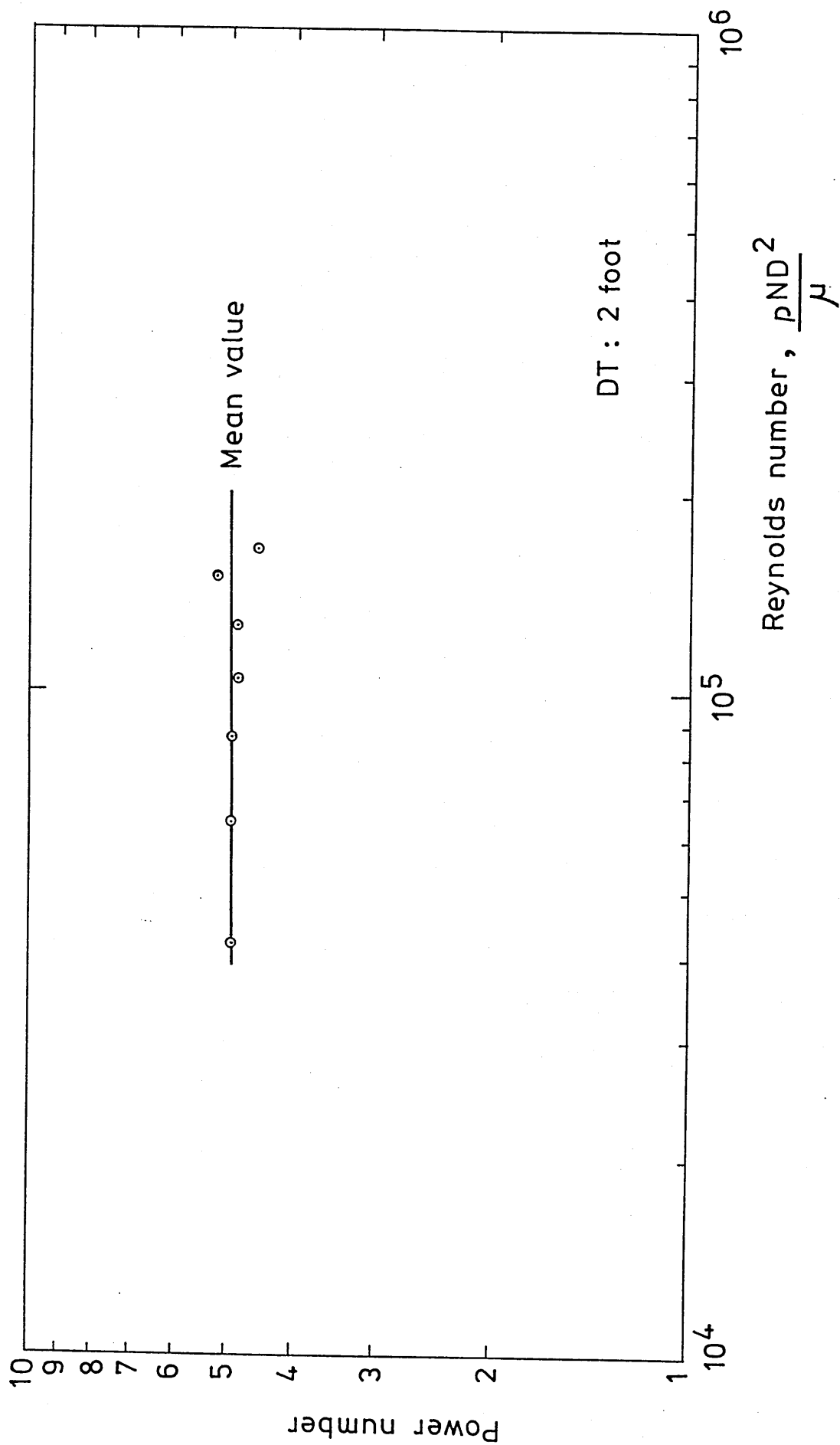


3.5 Effect of octanol concentration on $k_L a$



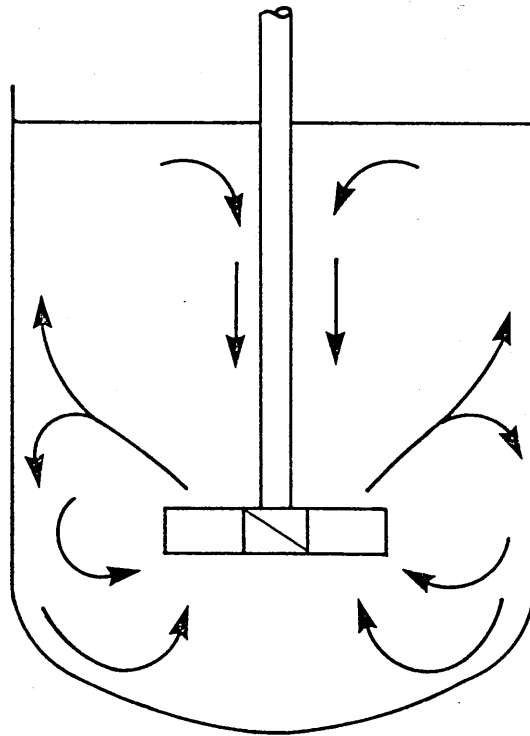
4.1 Instrument chain for strain gauge/telemetry system





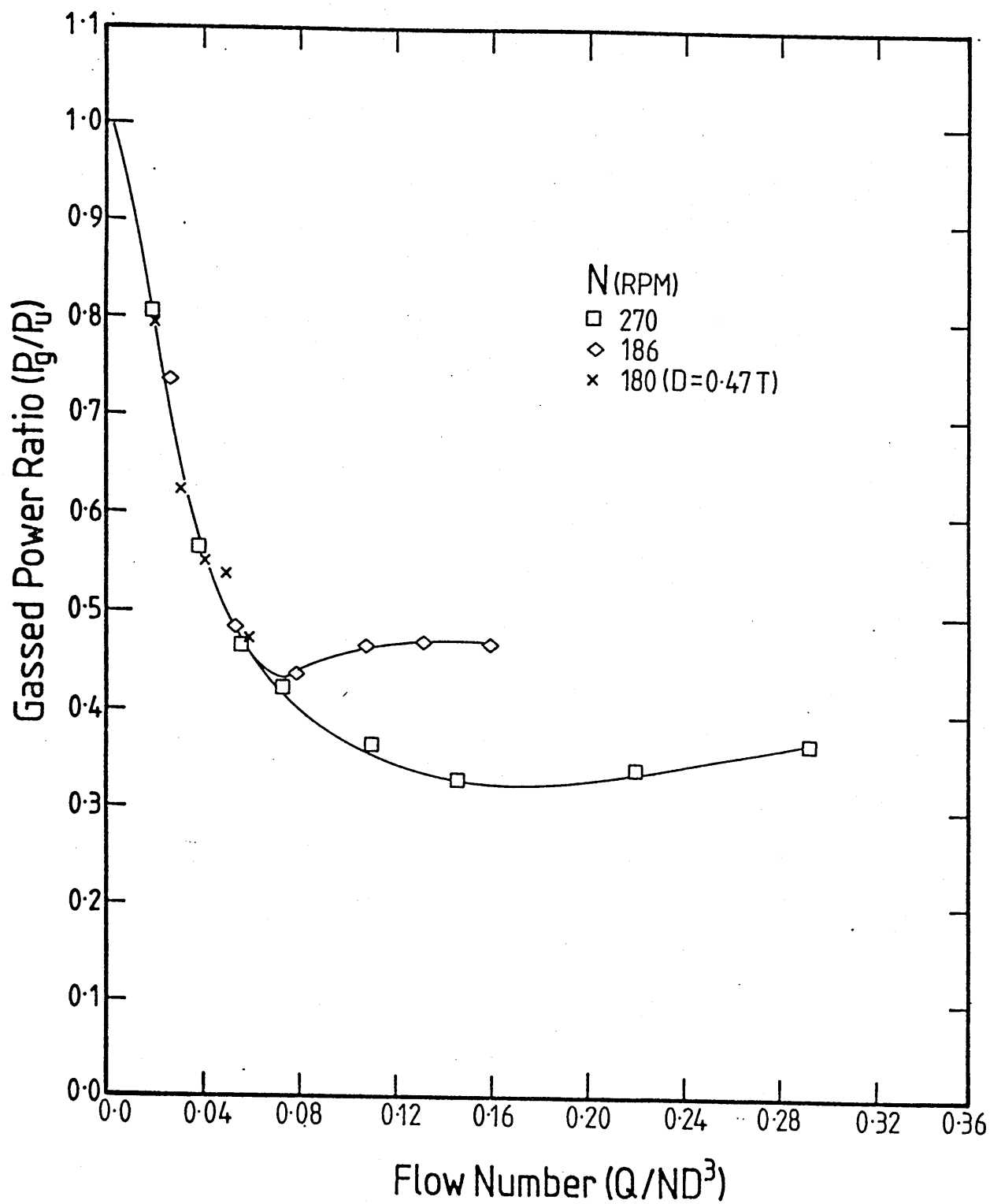
4.3

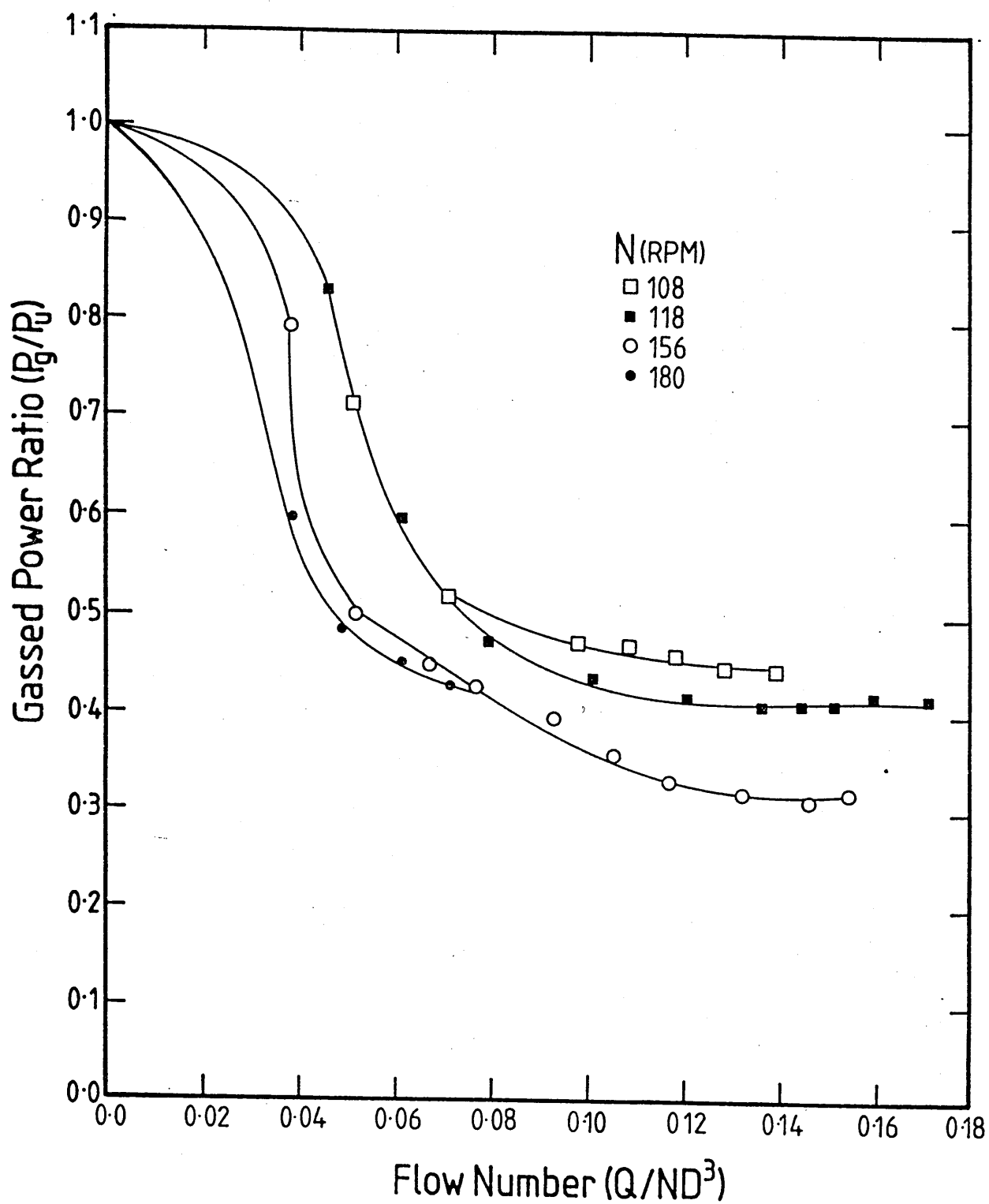
Ungassed power number as a function of Reynolds number 0.61 m tank

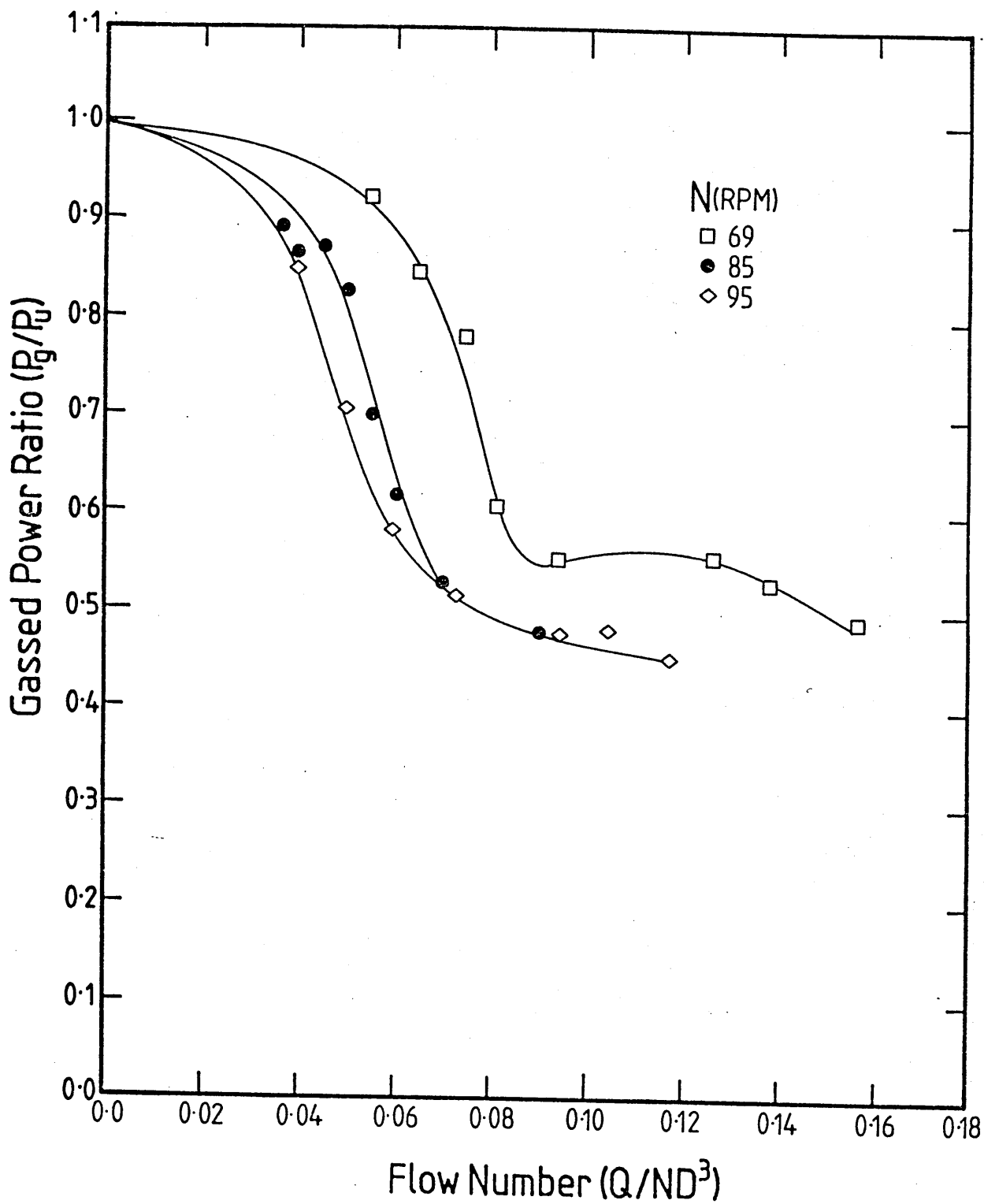


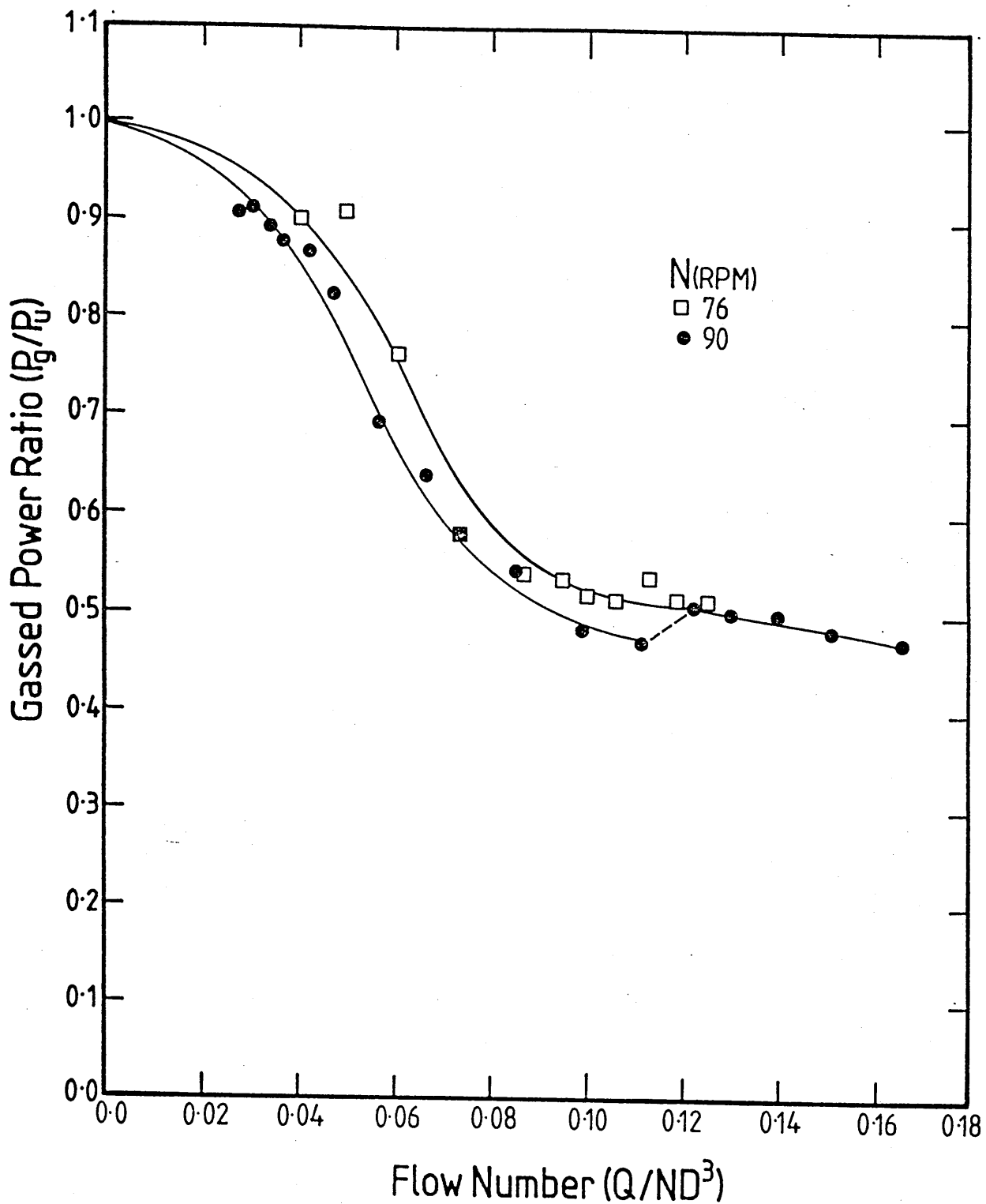
4.4

Single phase flow patterns: upwards-pumping PBT



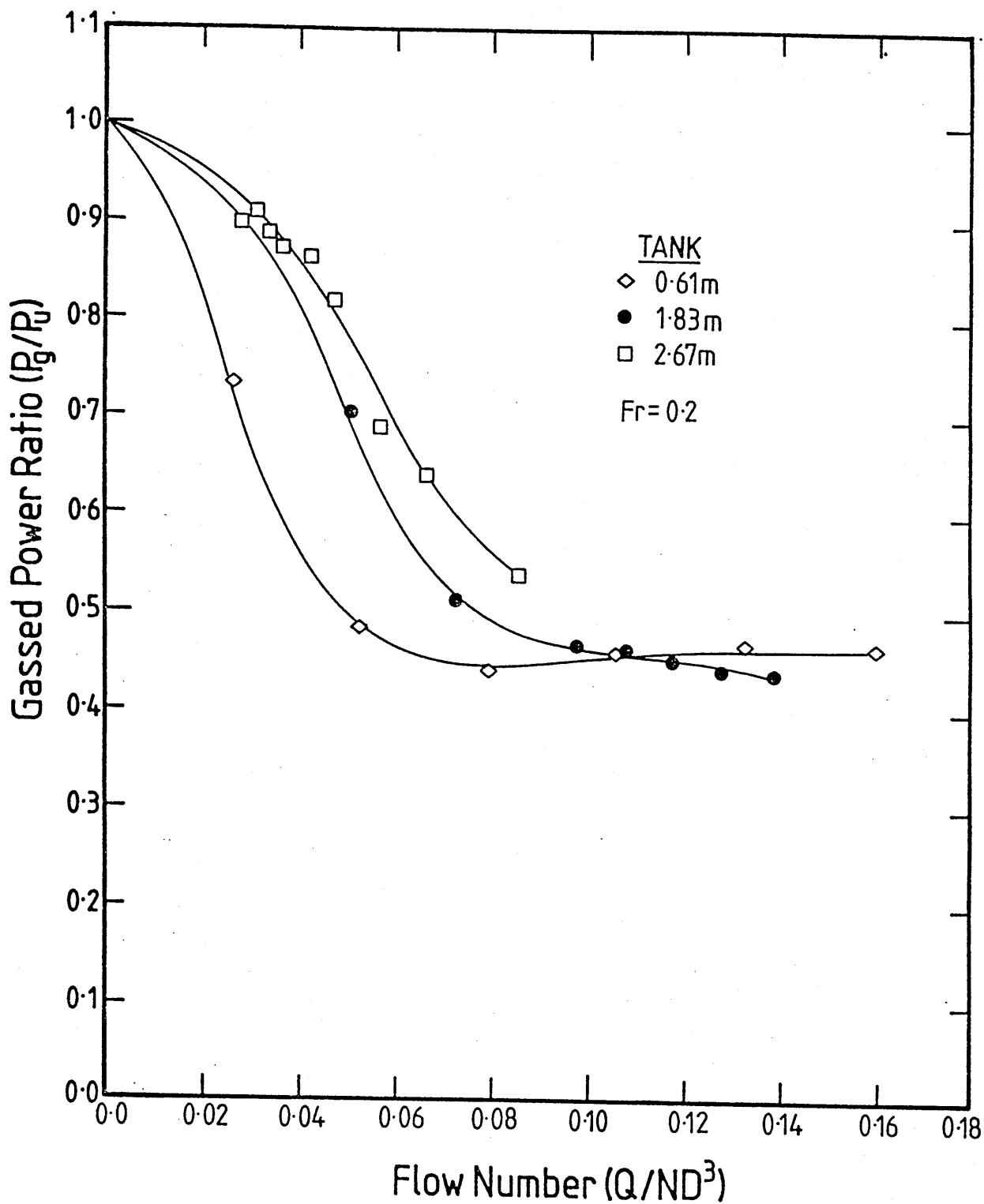


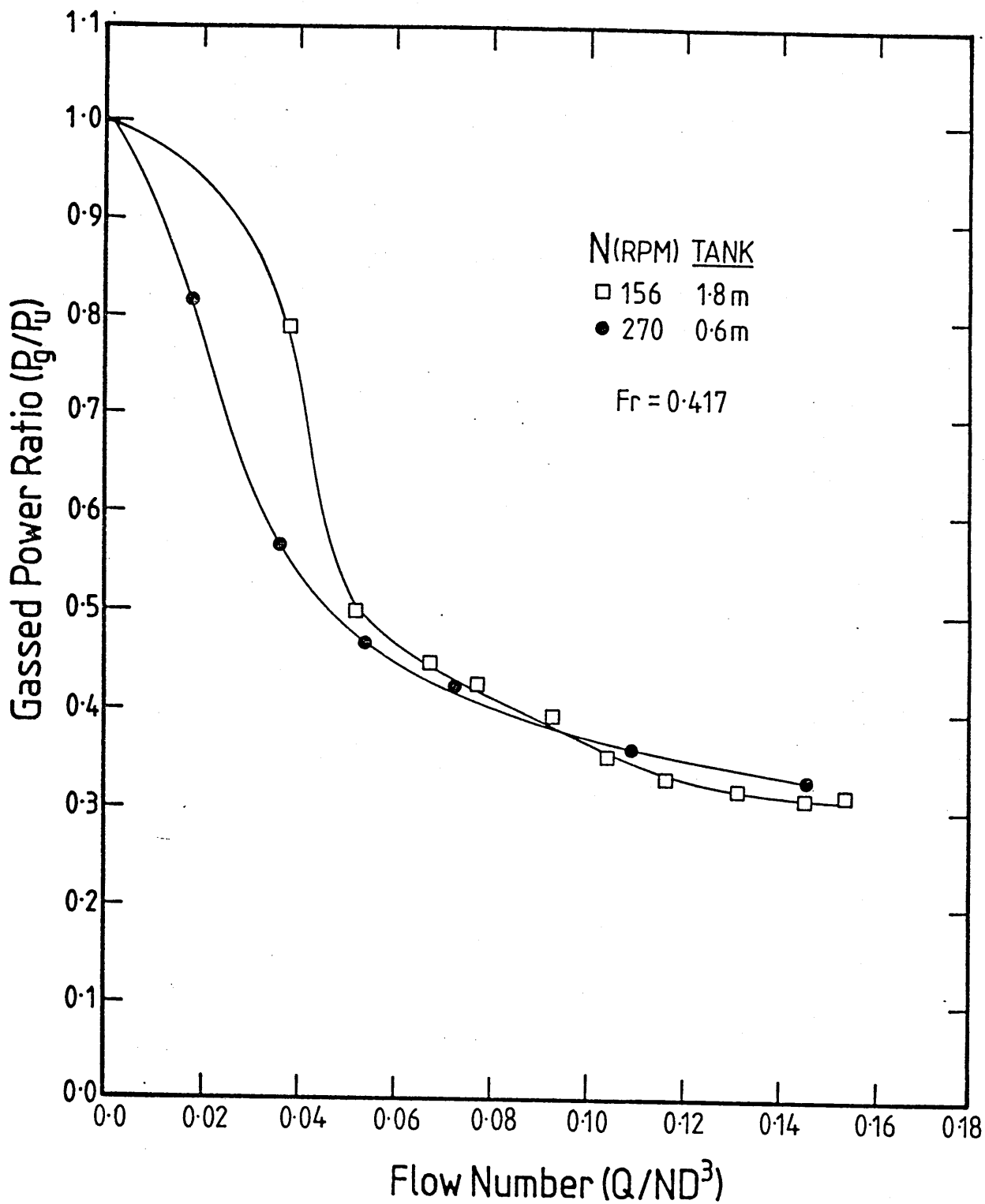


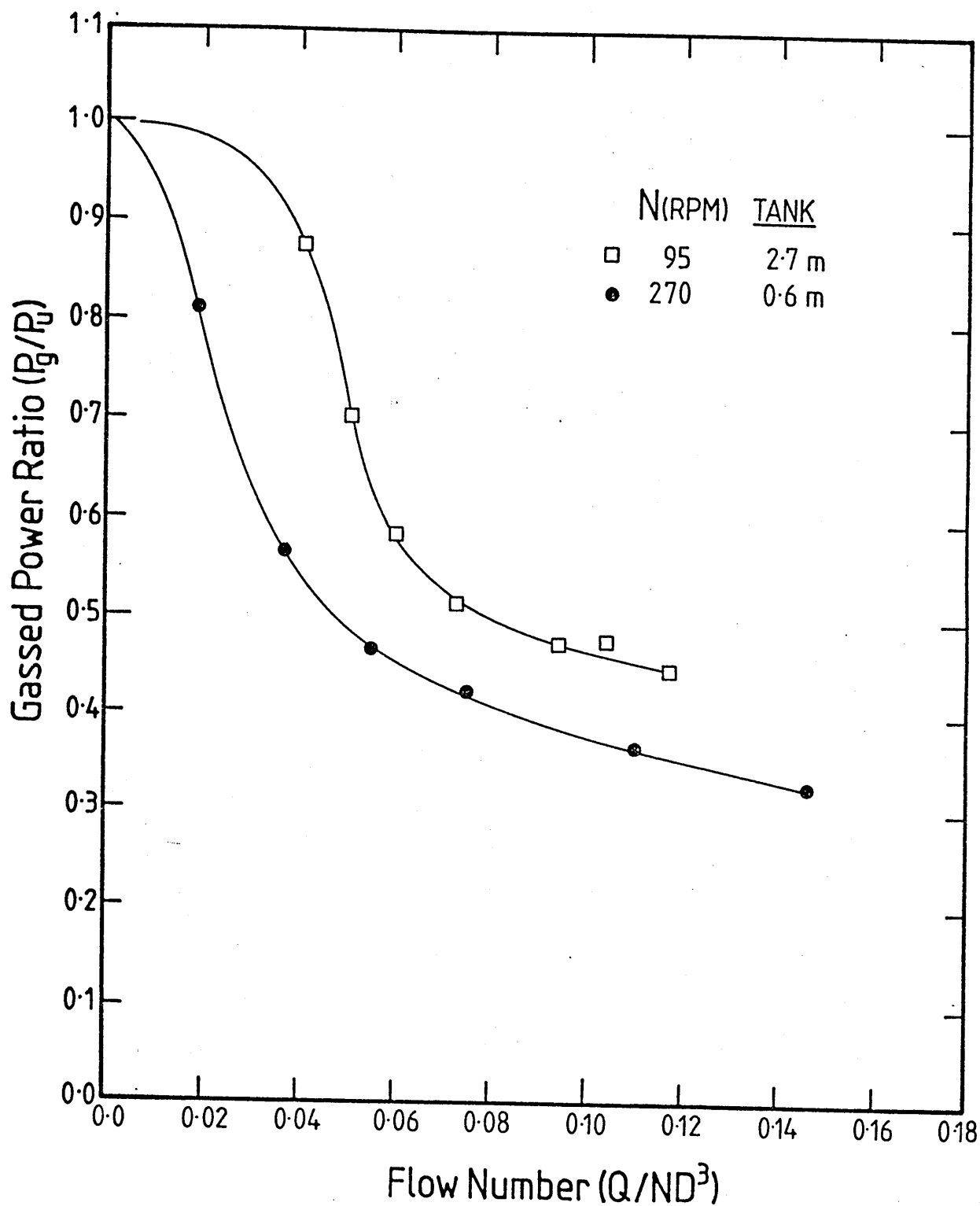


4.8

Gassed power curves: Rushton turbine: 2.67 m tank: octanol solution

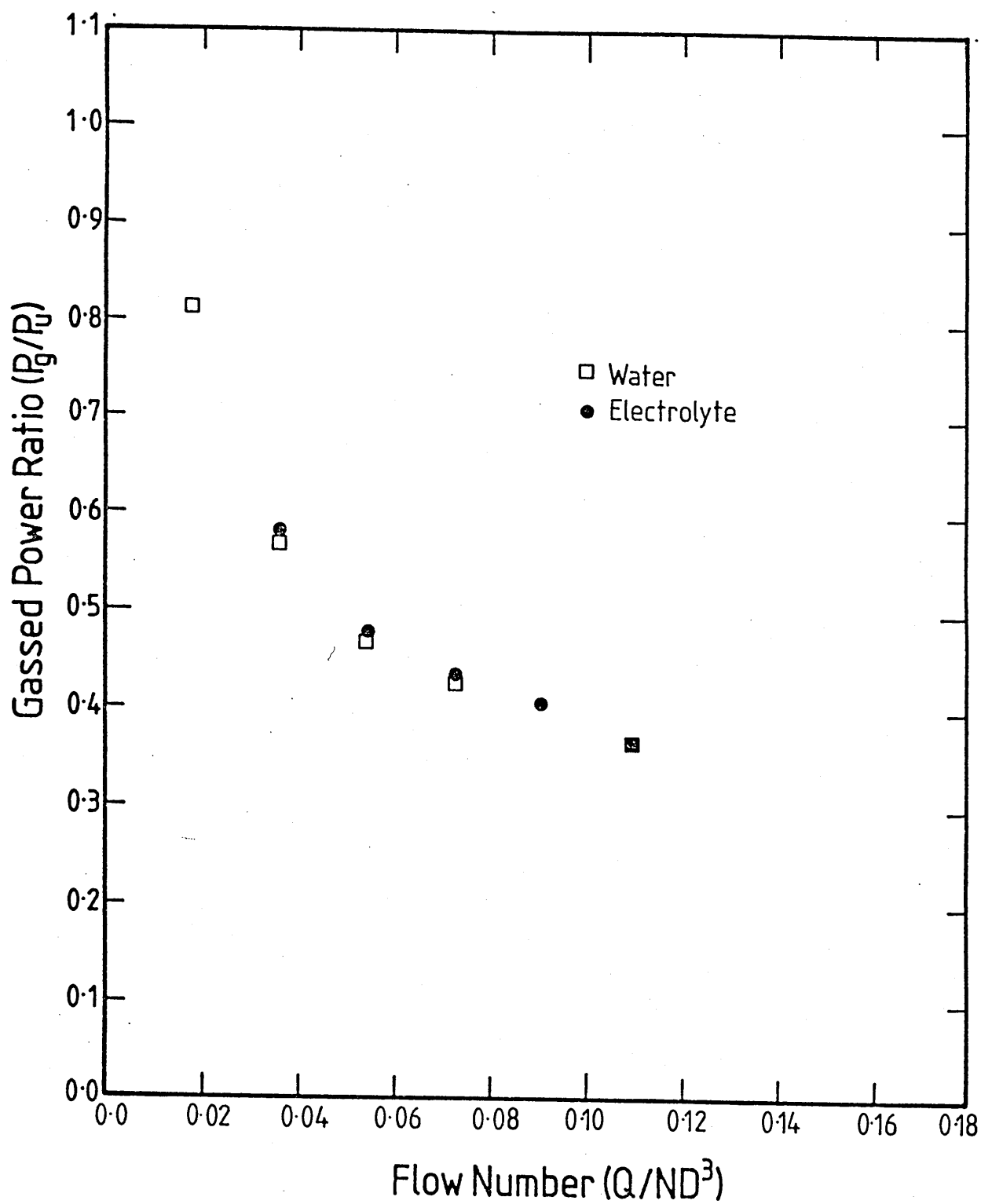






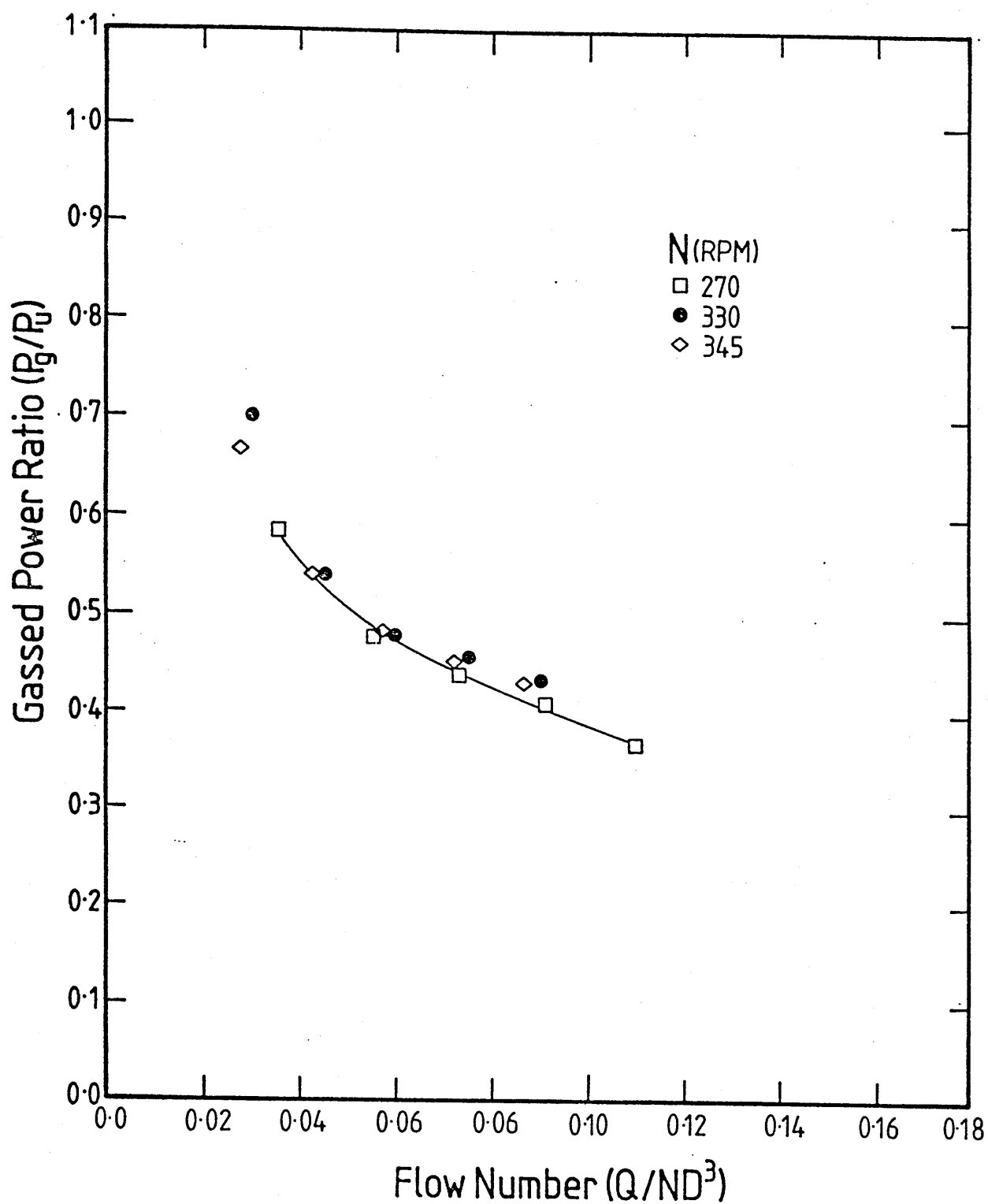
4.11

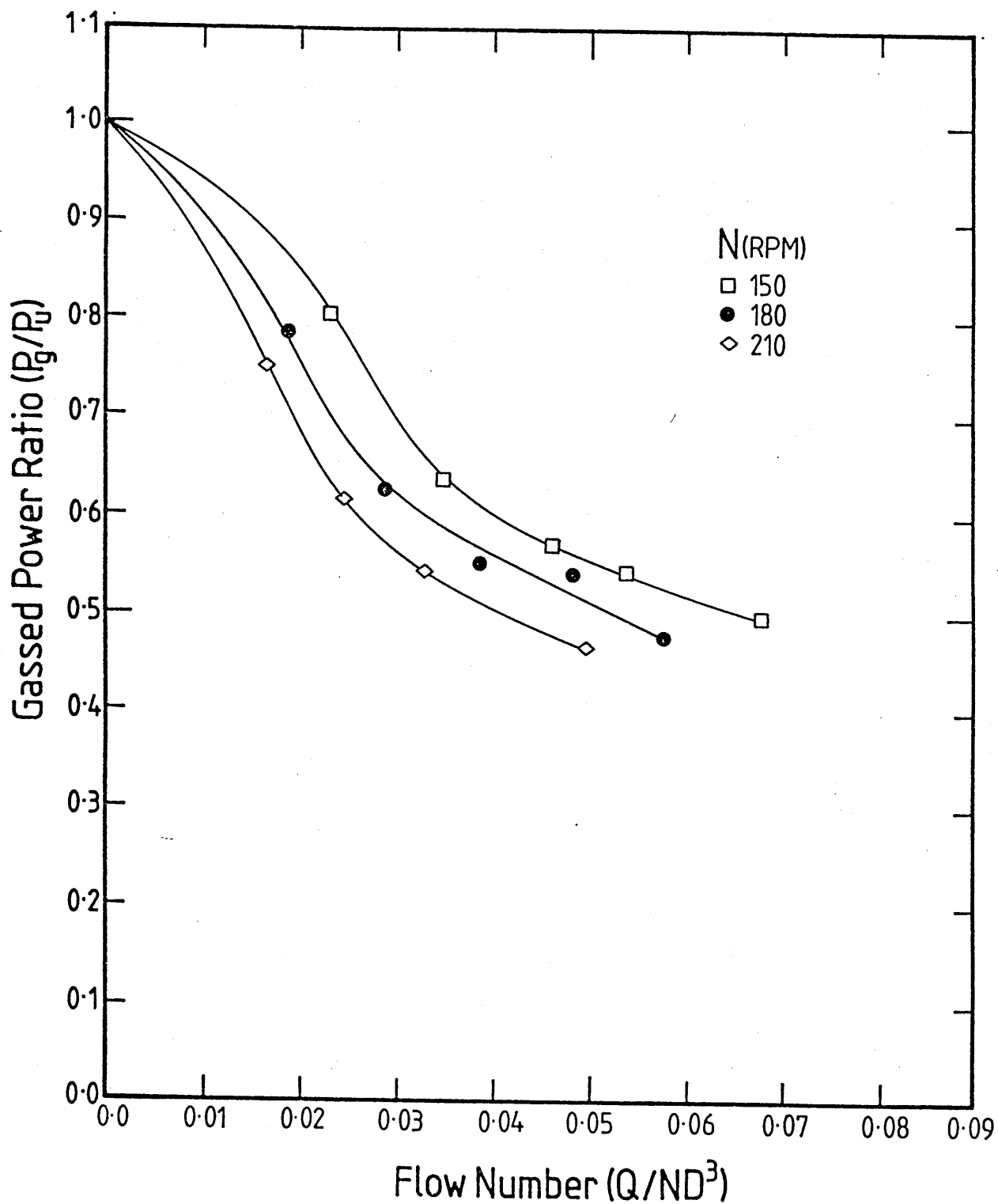
Rushton turbine: gassed power curves at constant ungassed power input per unit mass



4.12

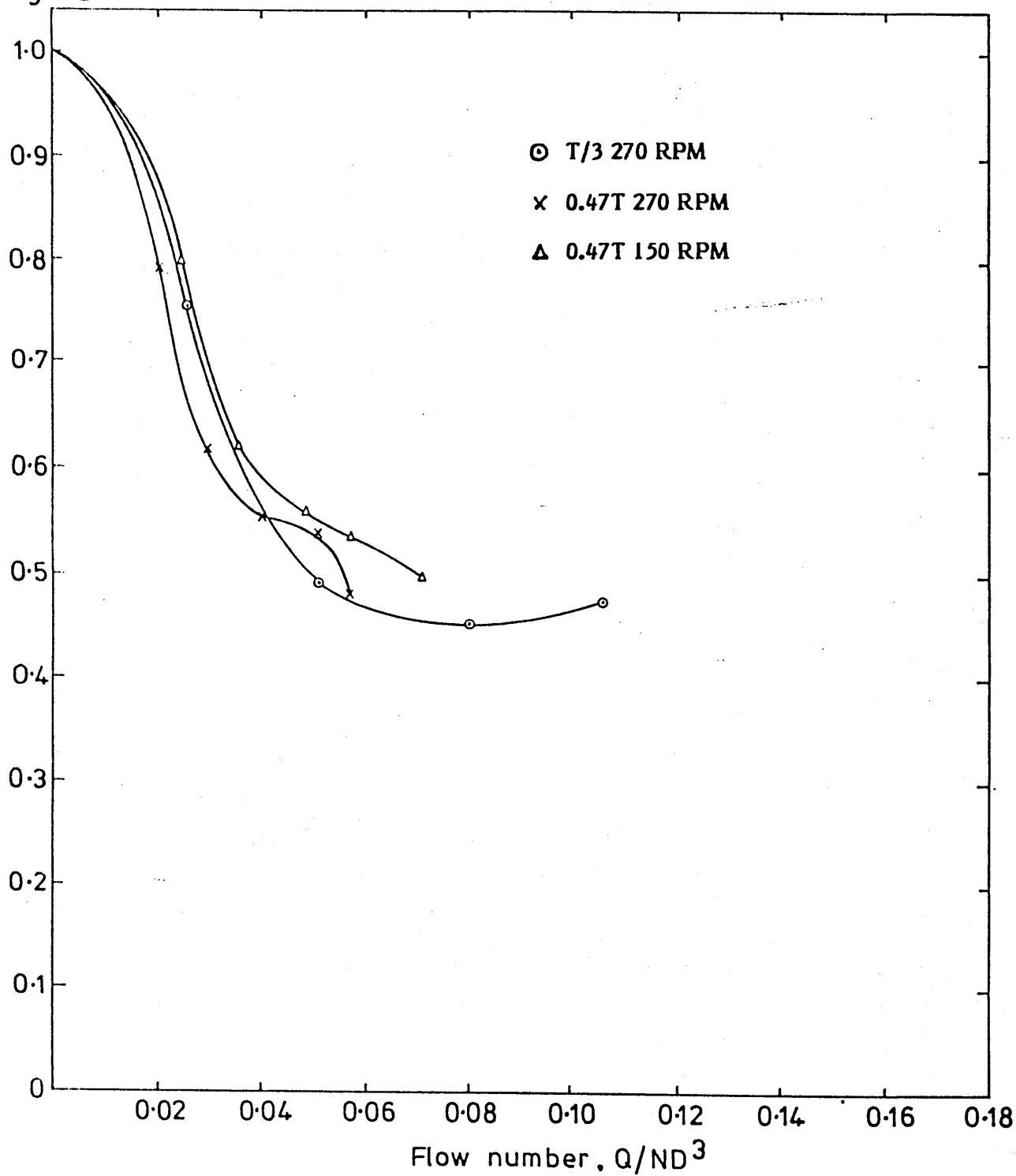
Rushton turbine: 0.61 m tank: effect of liquid type on gassed power curves

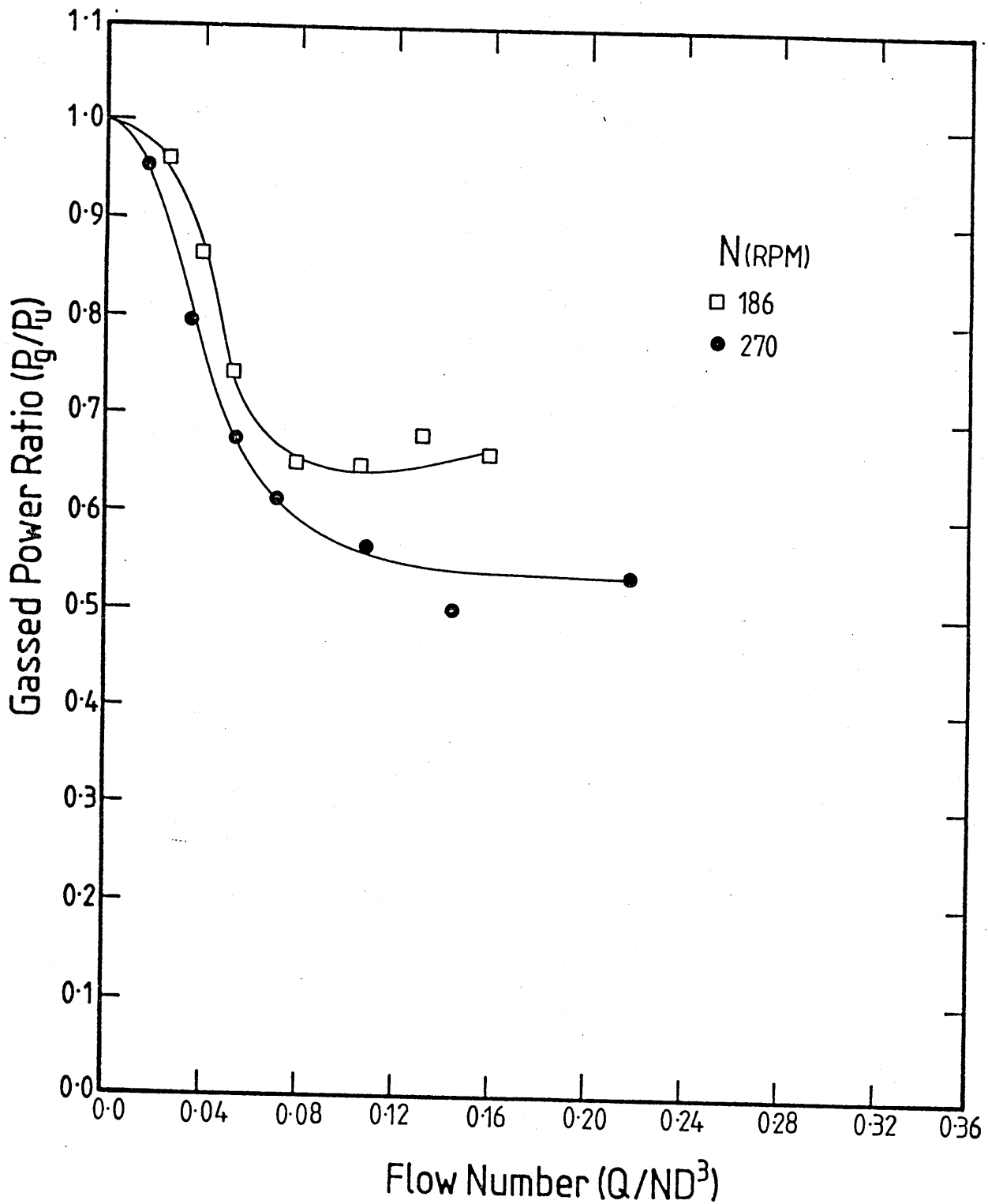




Gassed power ratio

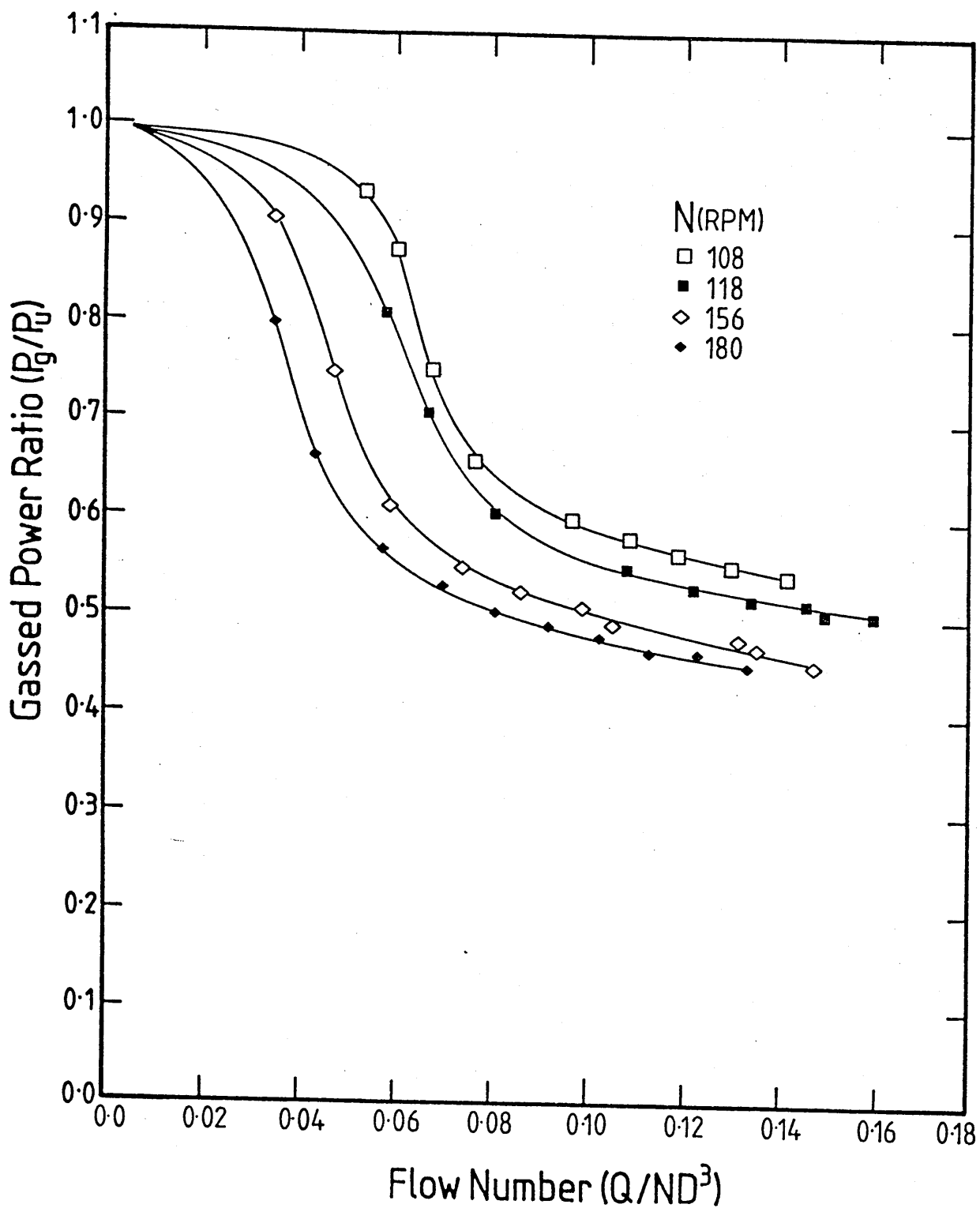
P_g/P_u





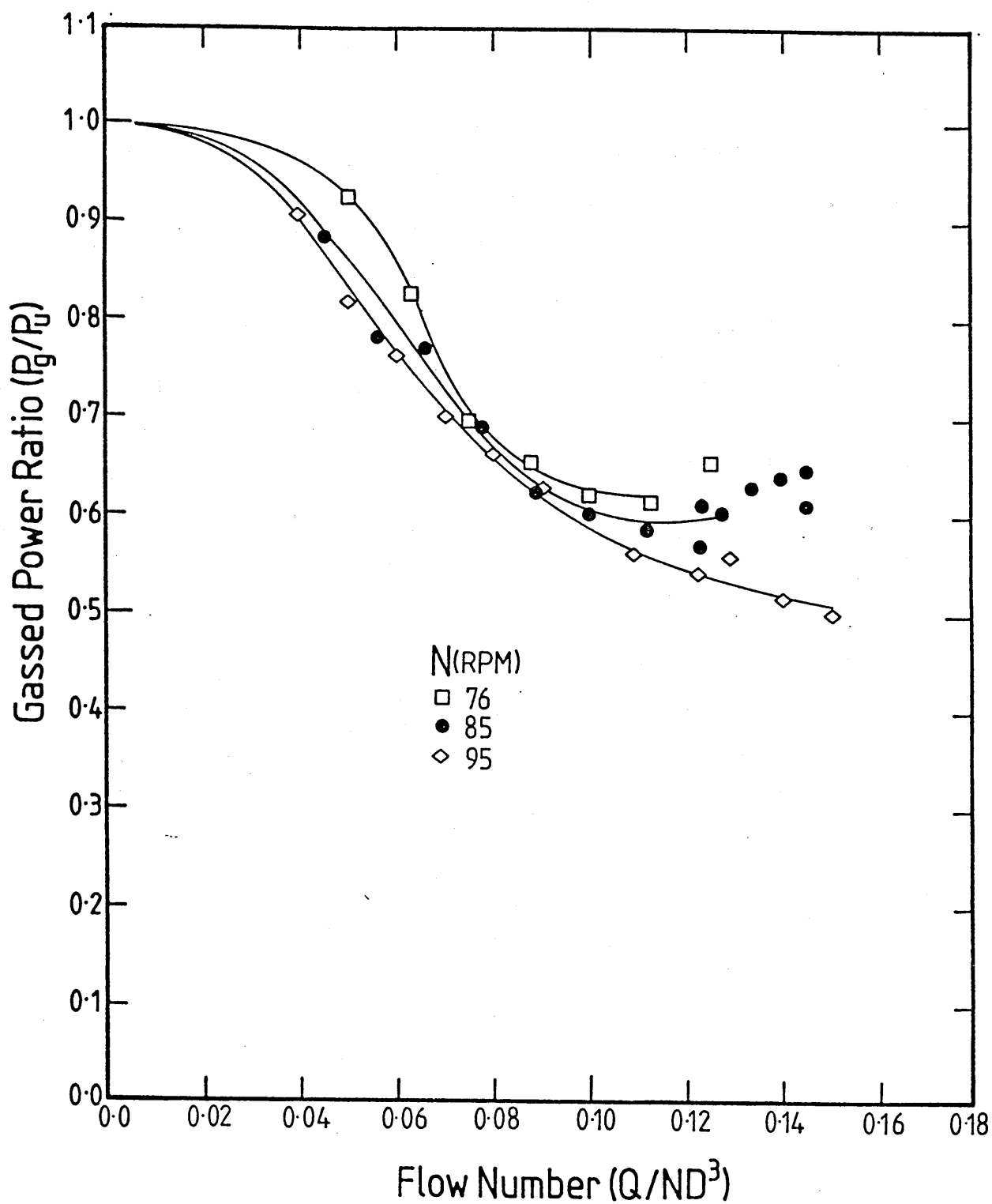
4.16

Concave turbine gassed power curves: 0.61 m tank: octanol solution



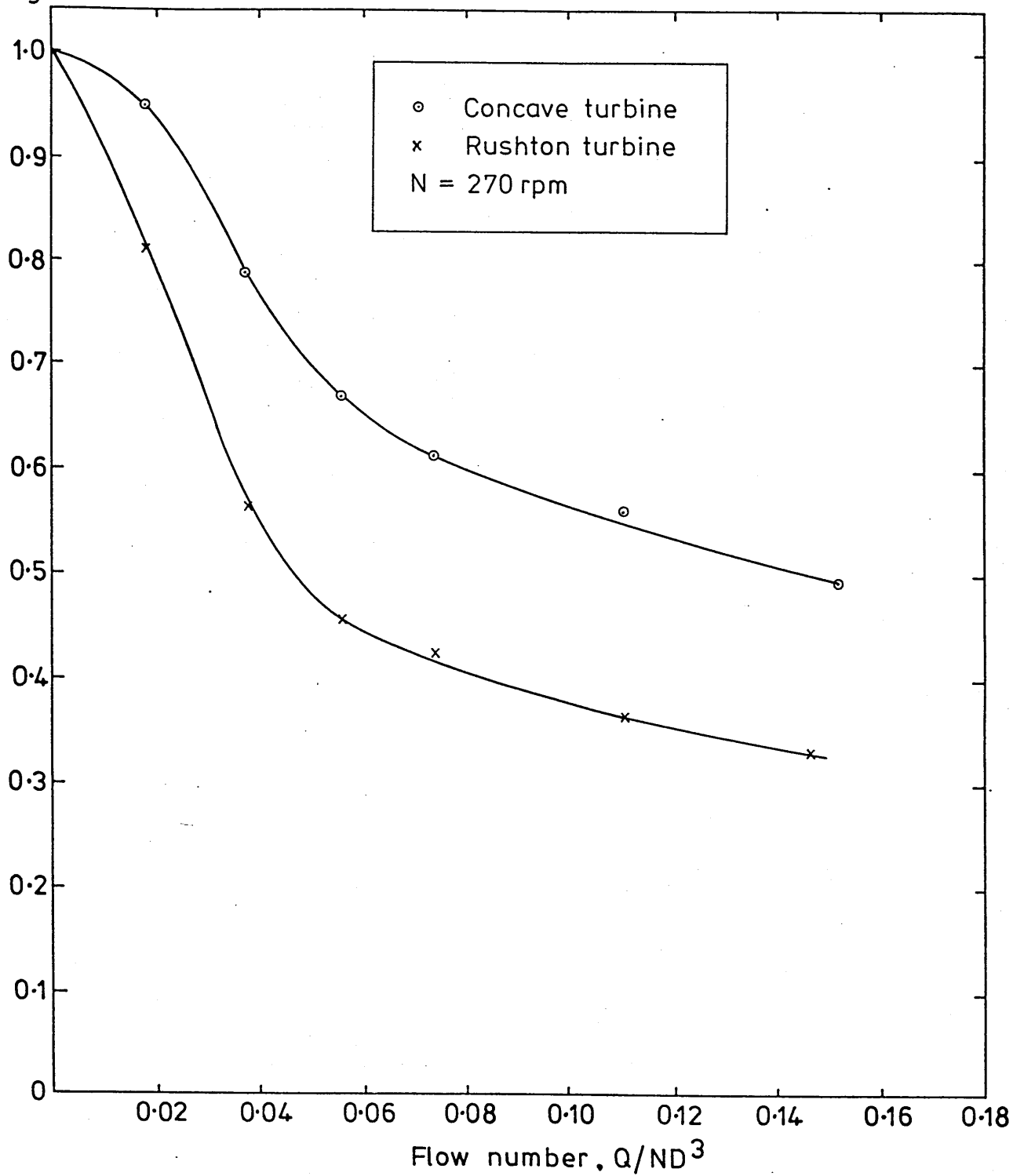
4.17

Concave turbine gassed power curves: 1.83 m tank: octanol solution



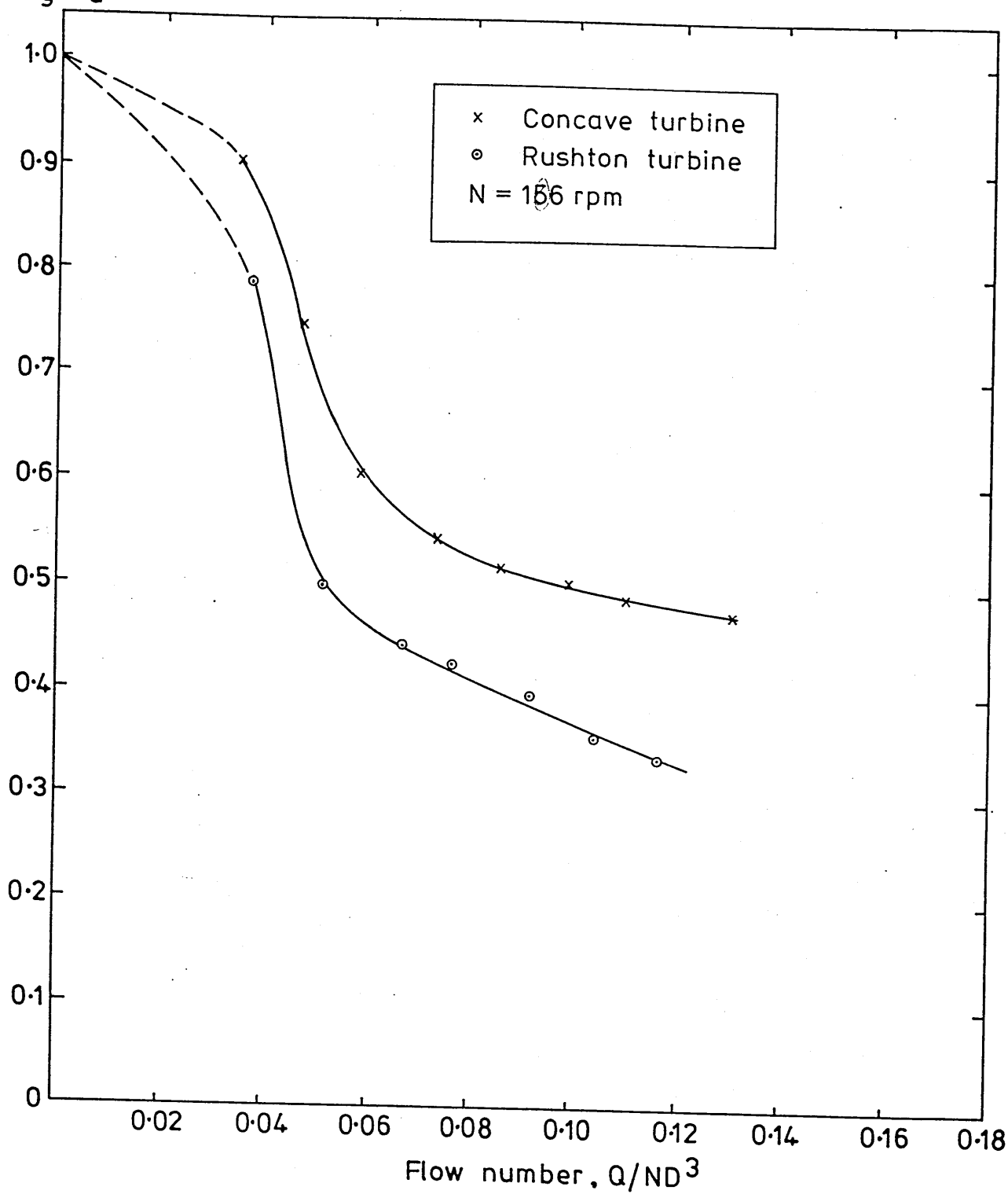
Gassed power ratio

P_g/P_u



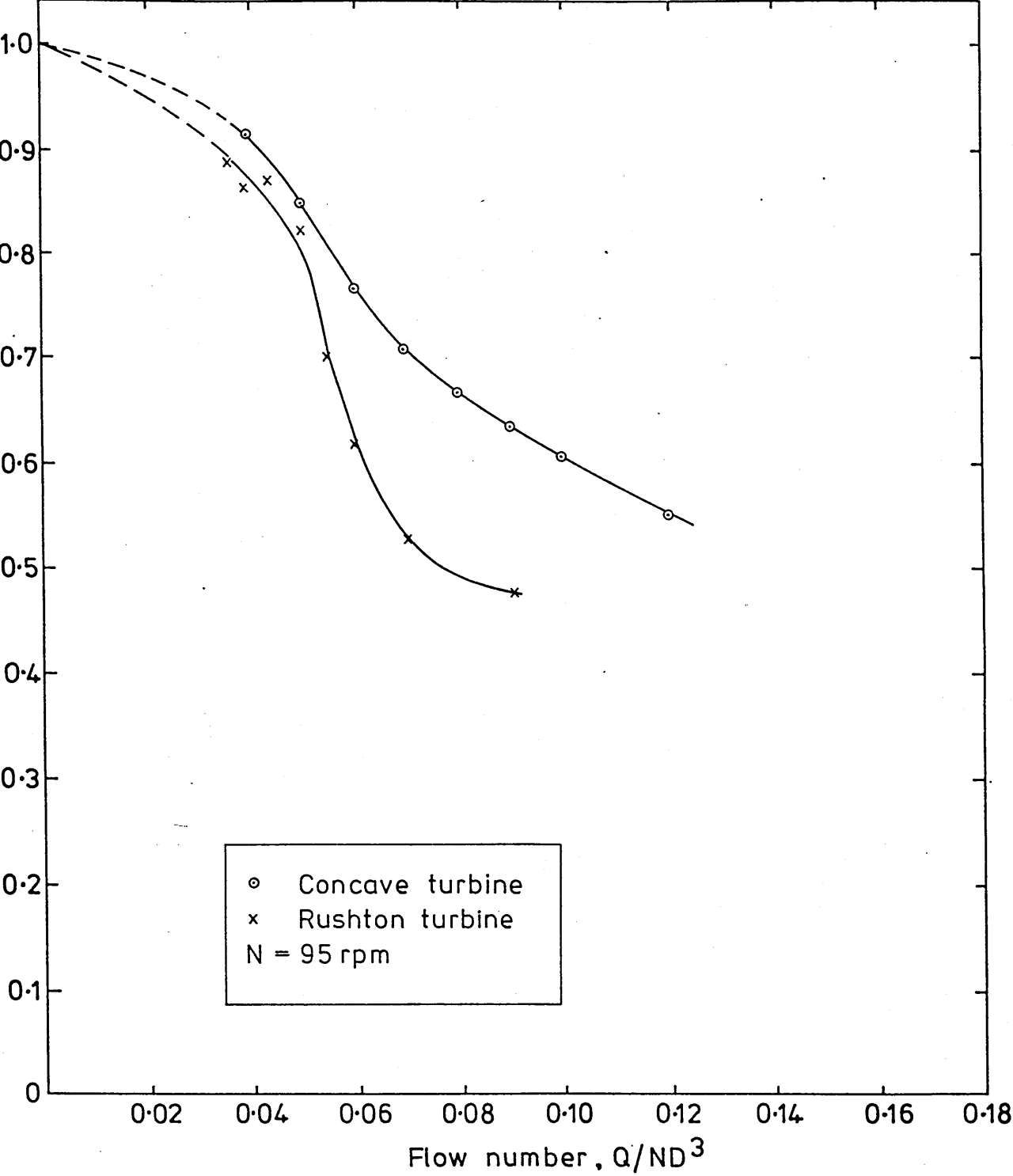
Gassed power ratio

P_g/P_u

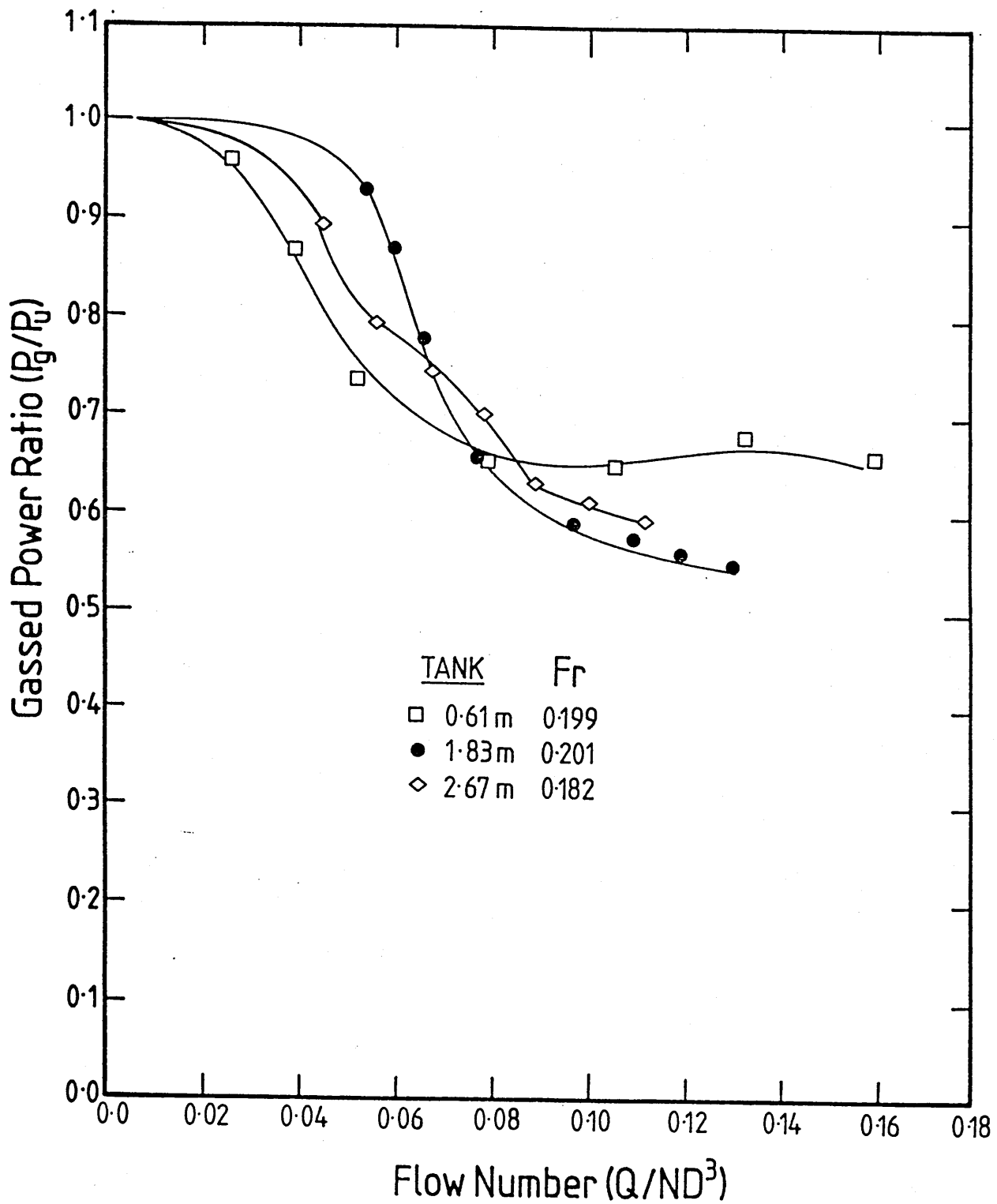


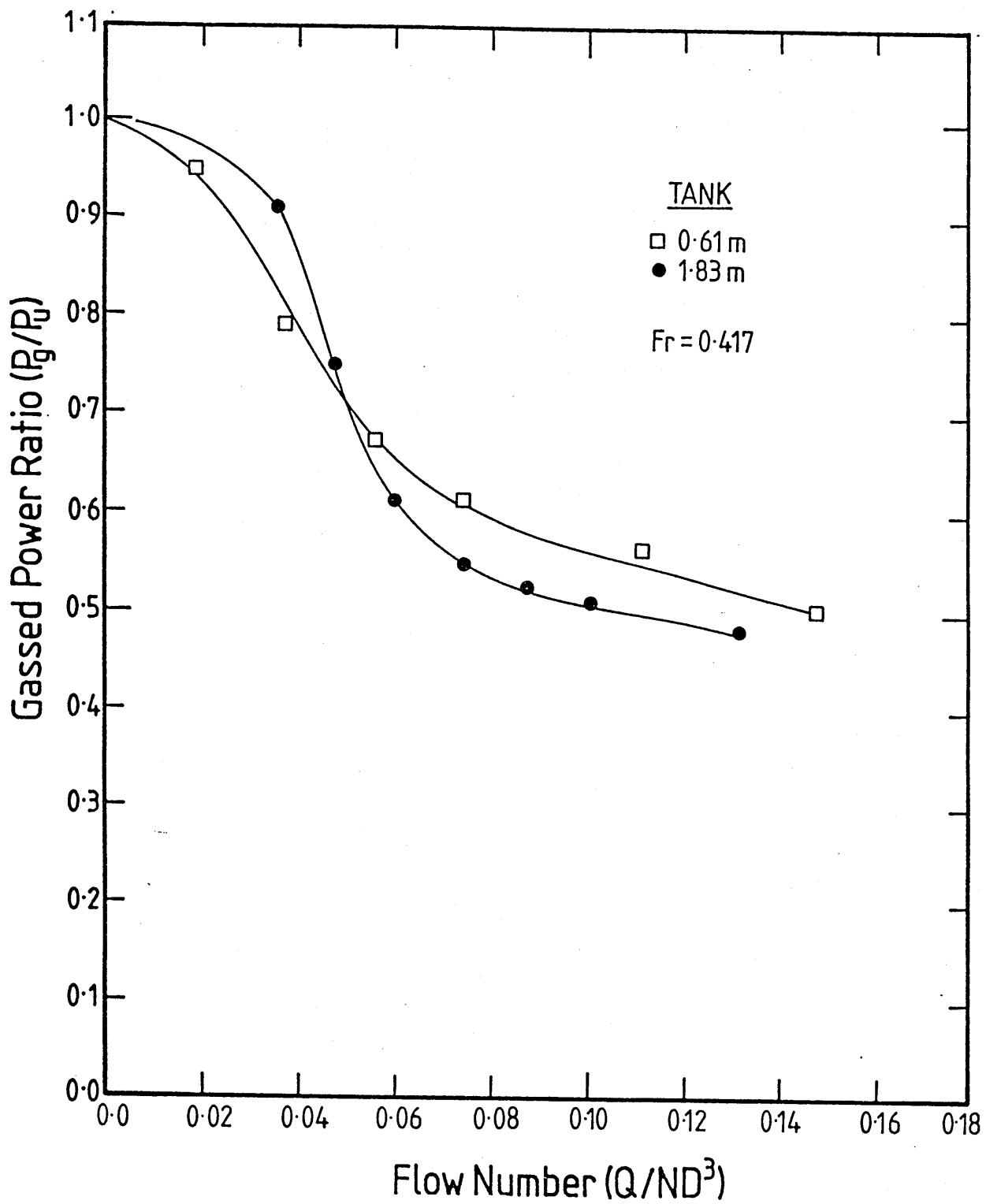
Gassed power ratio

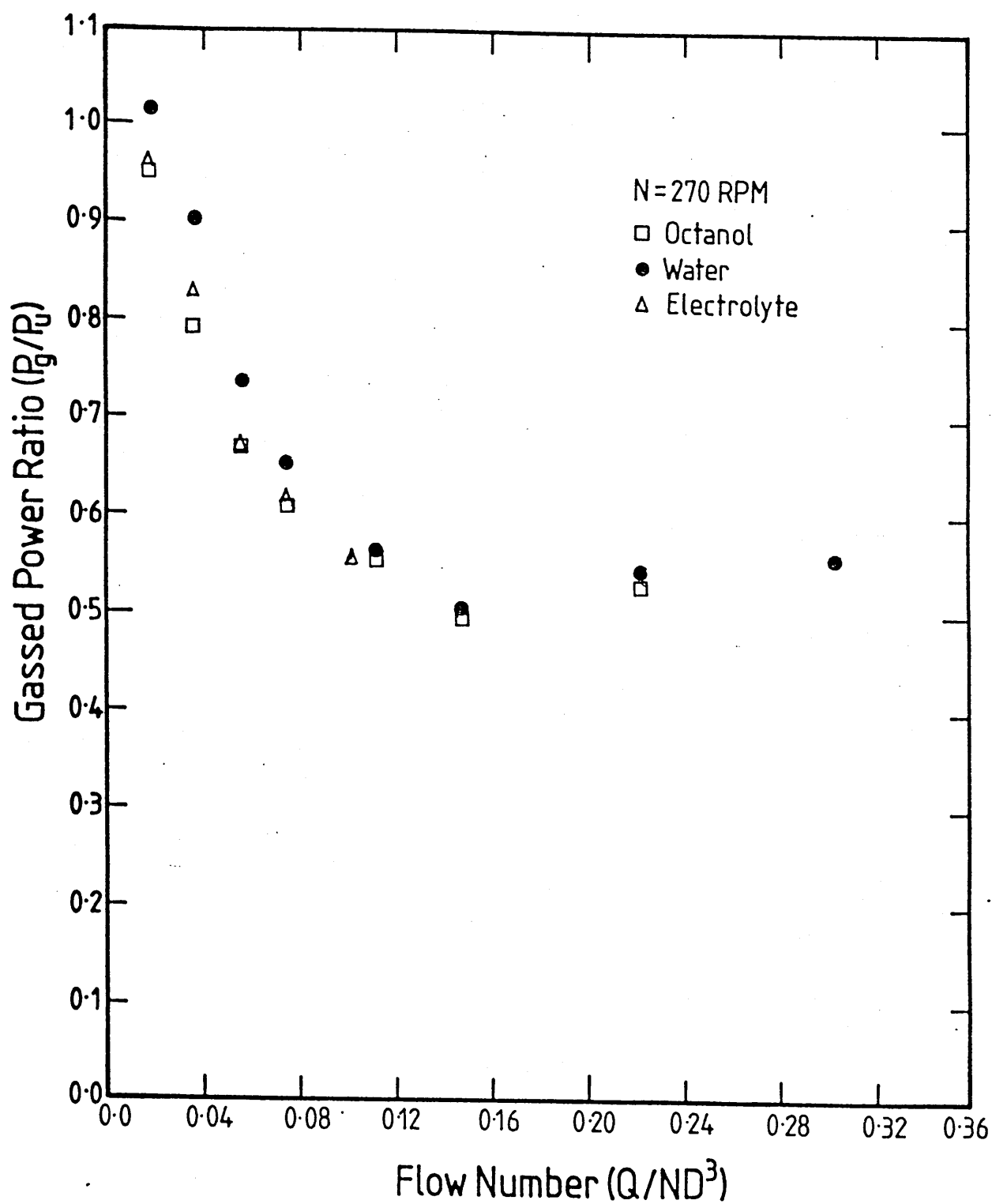
$$P_g/P_u$$

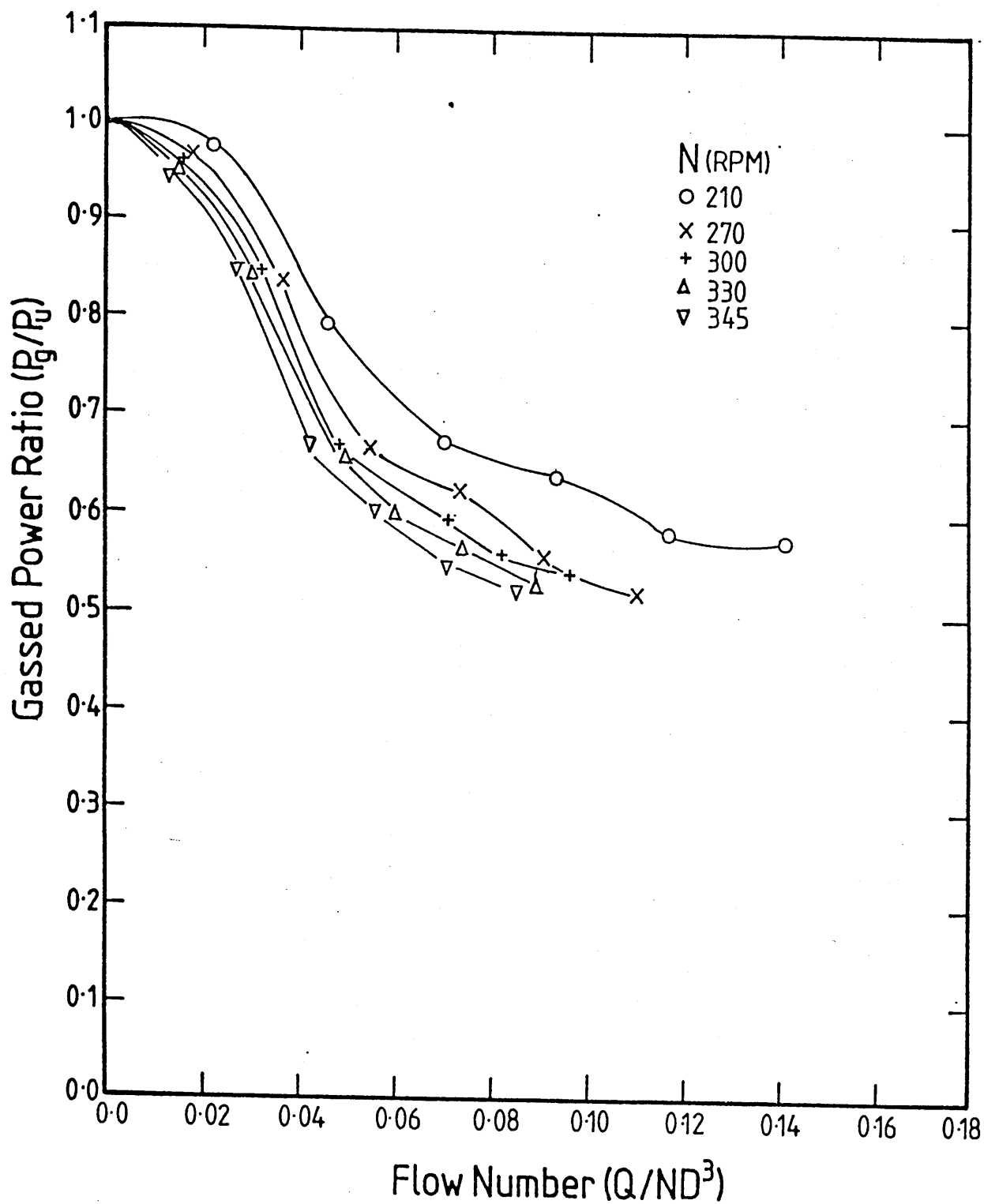


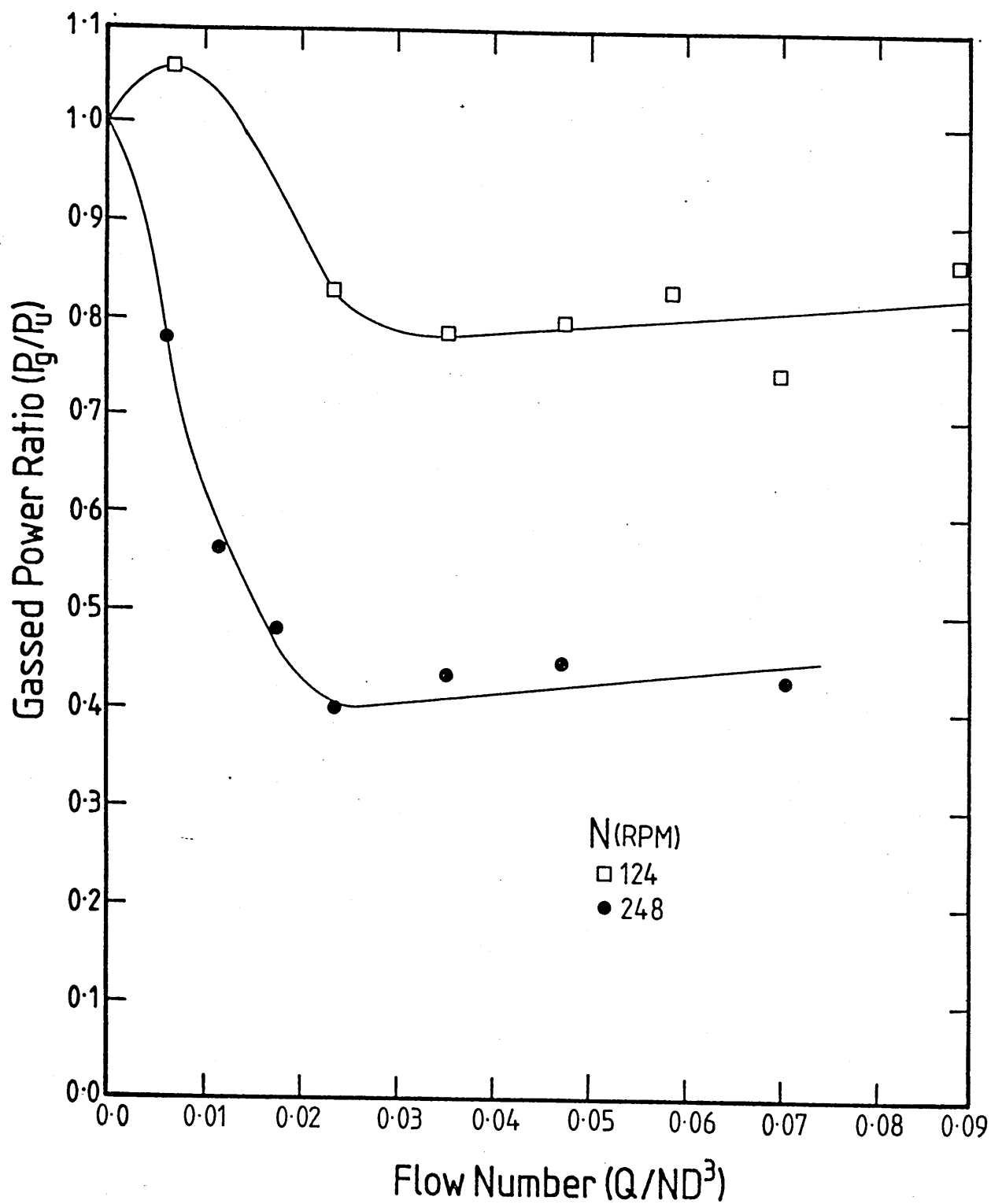
○ Concave turbine
x Rushton turbine
N = 95 rpm

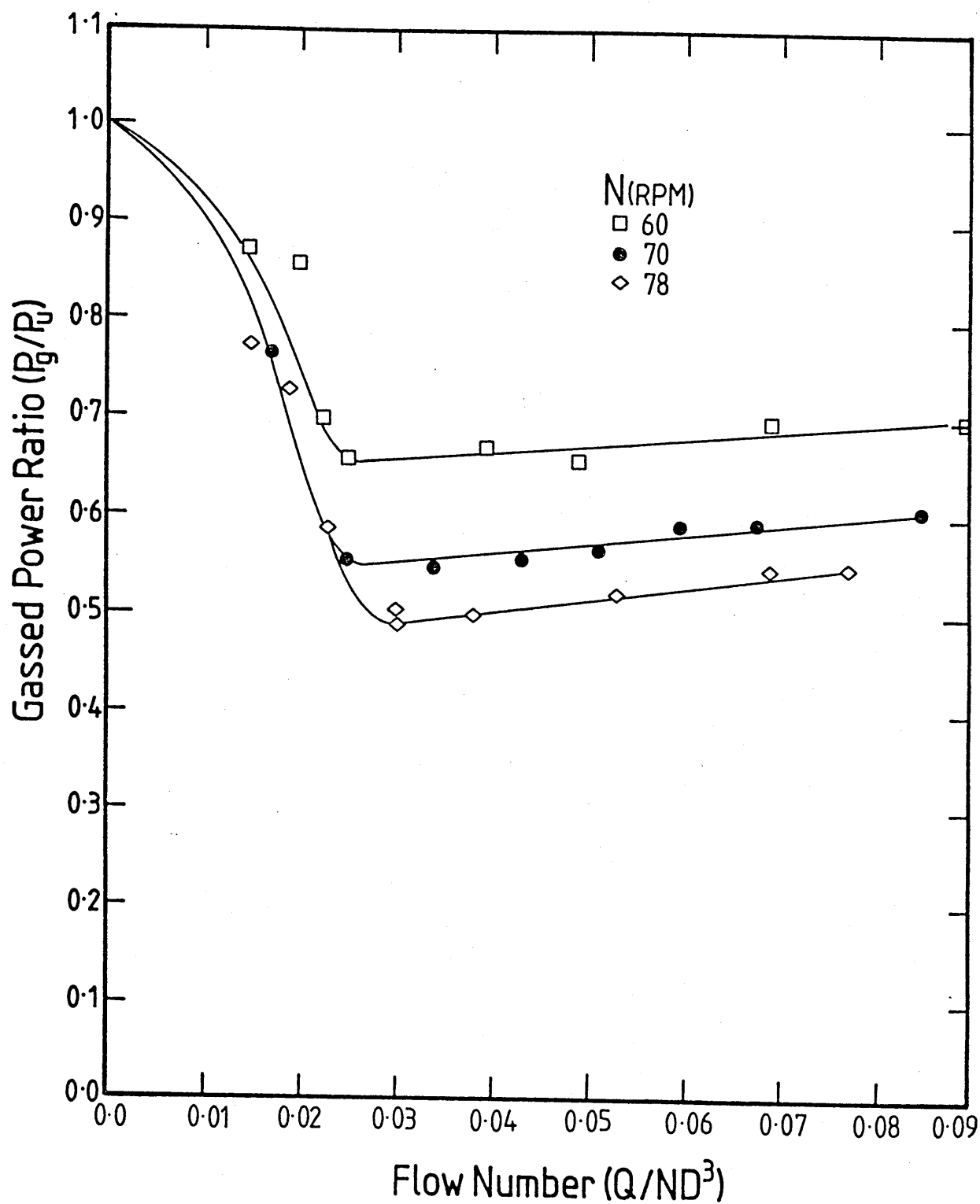


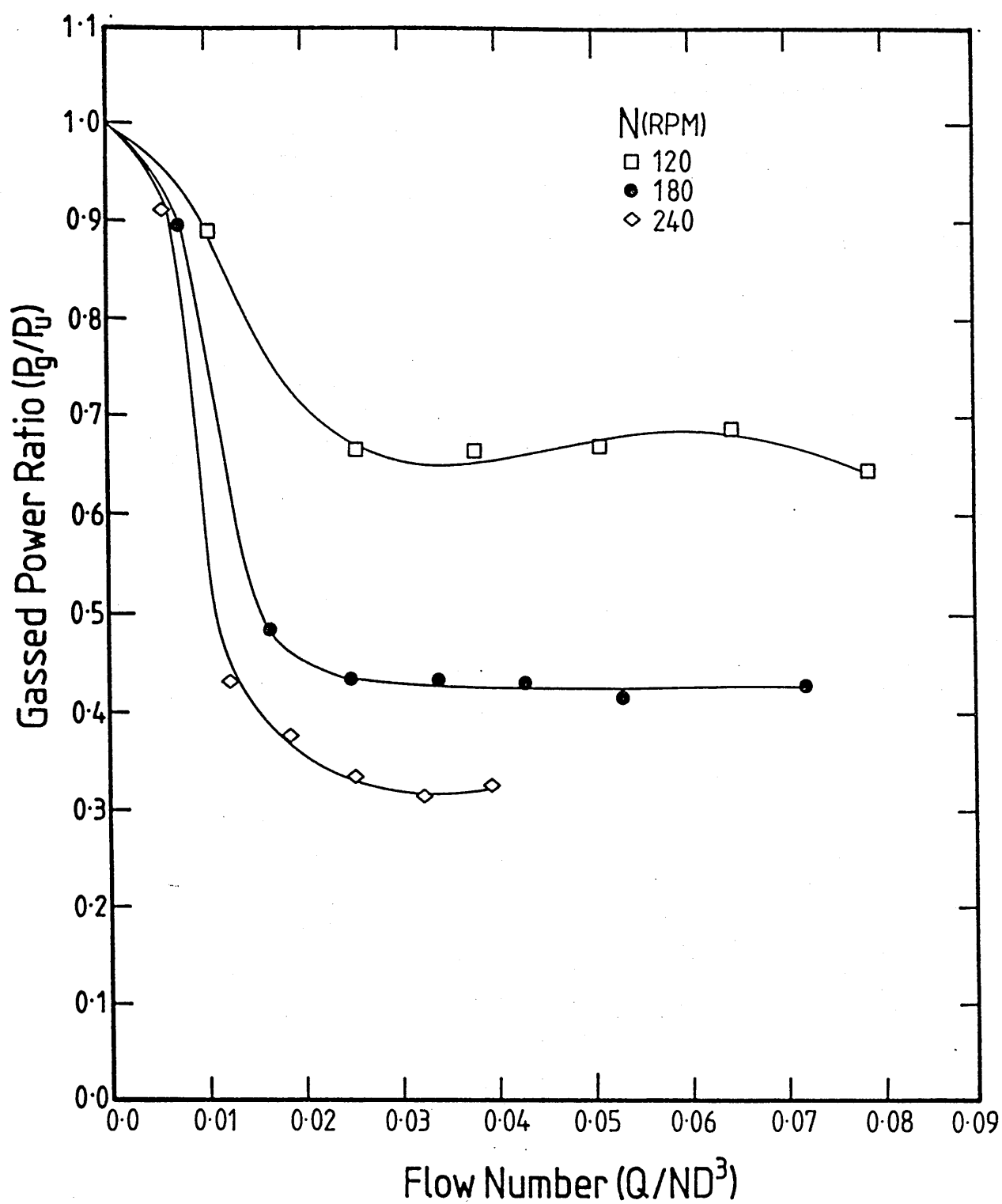


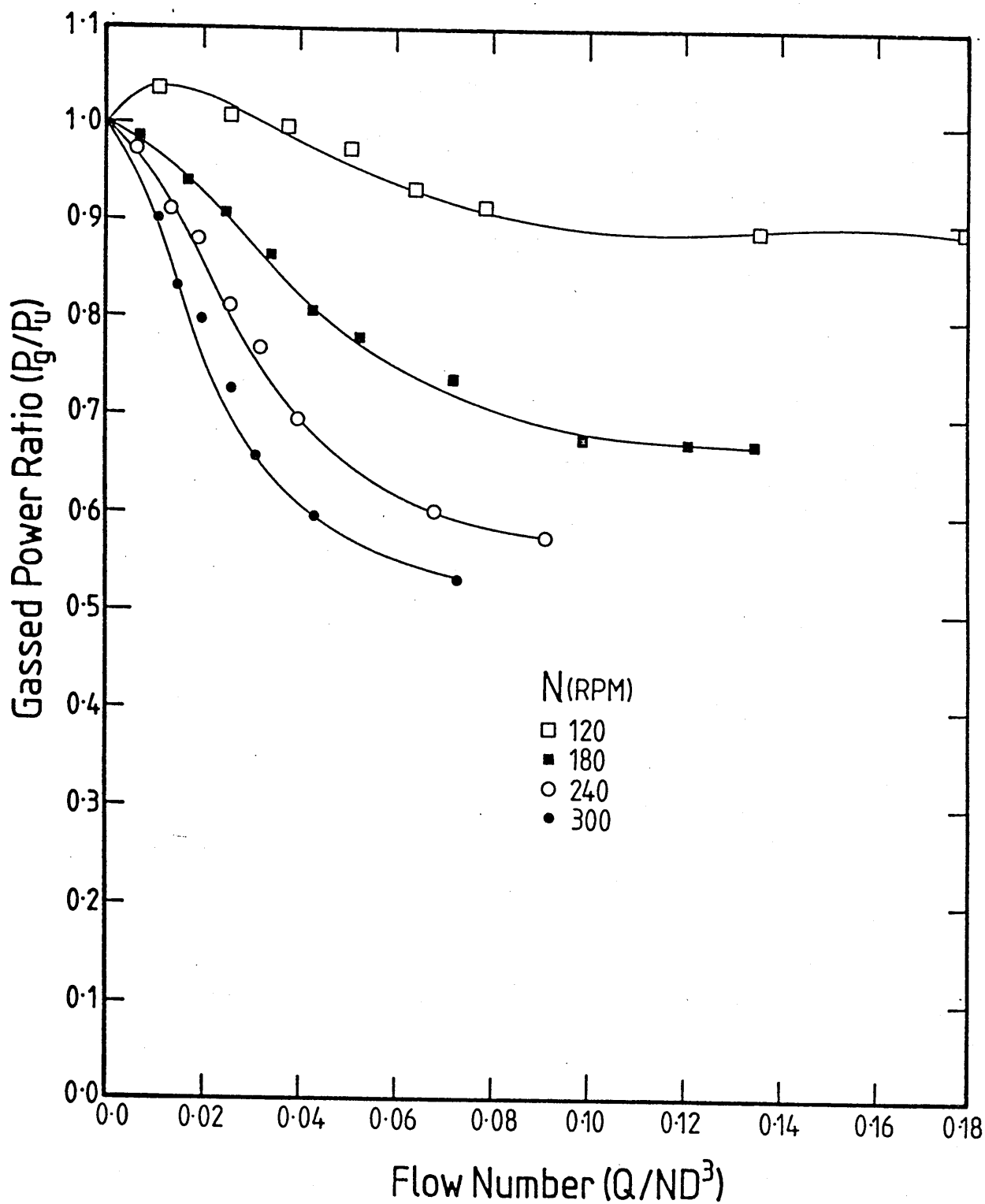


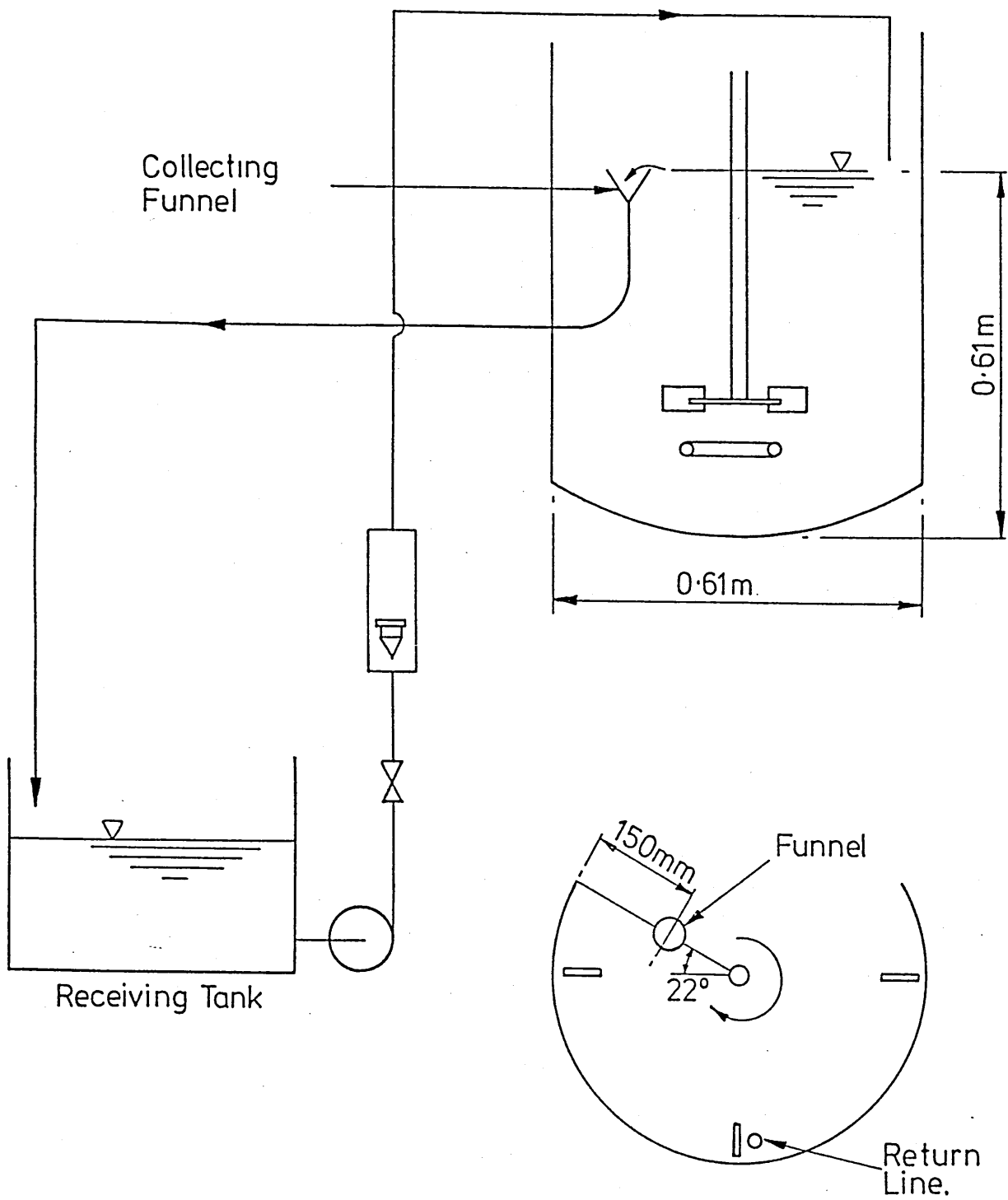


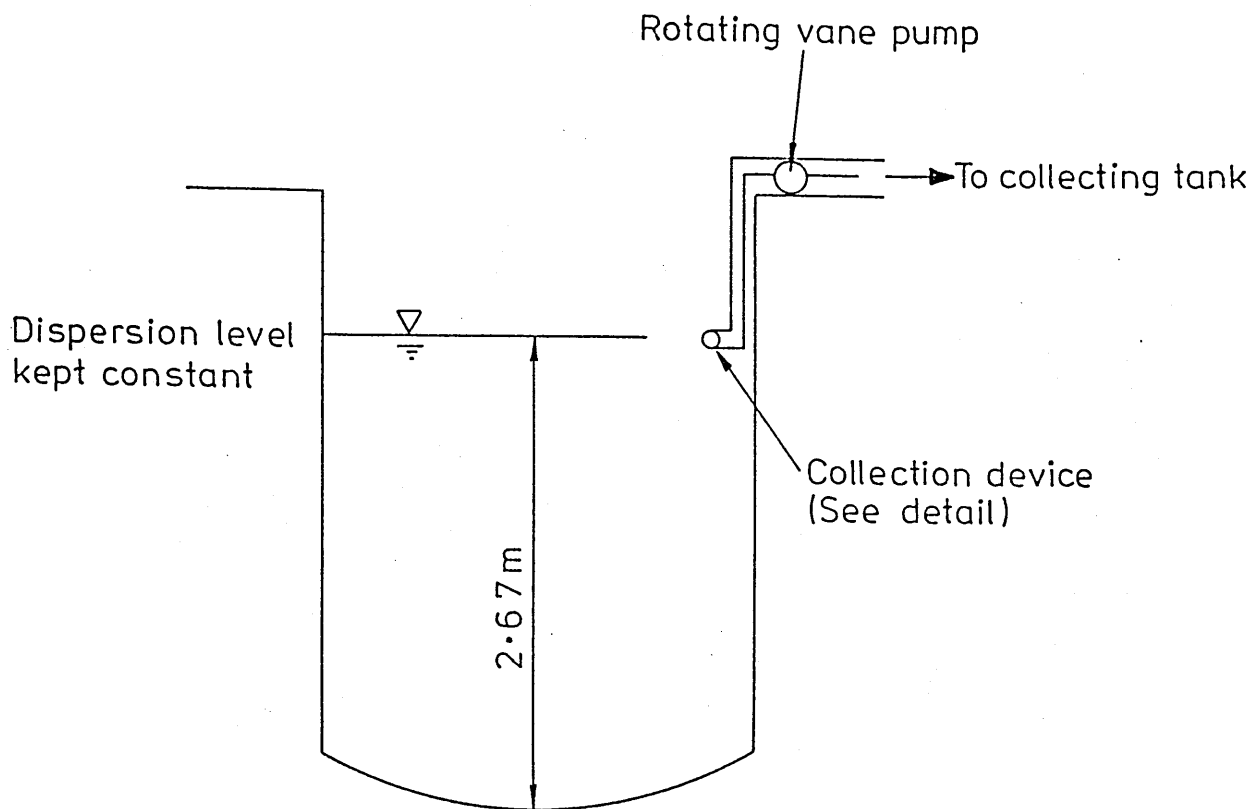
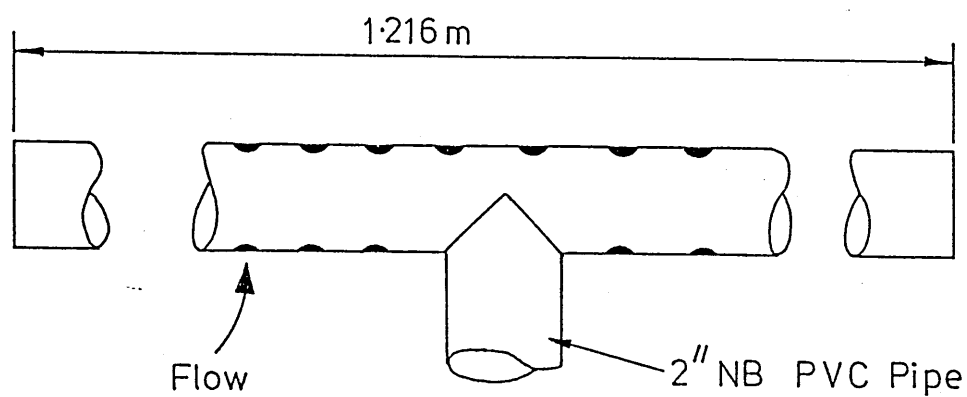








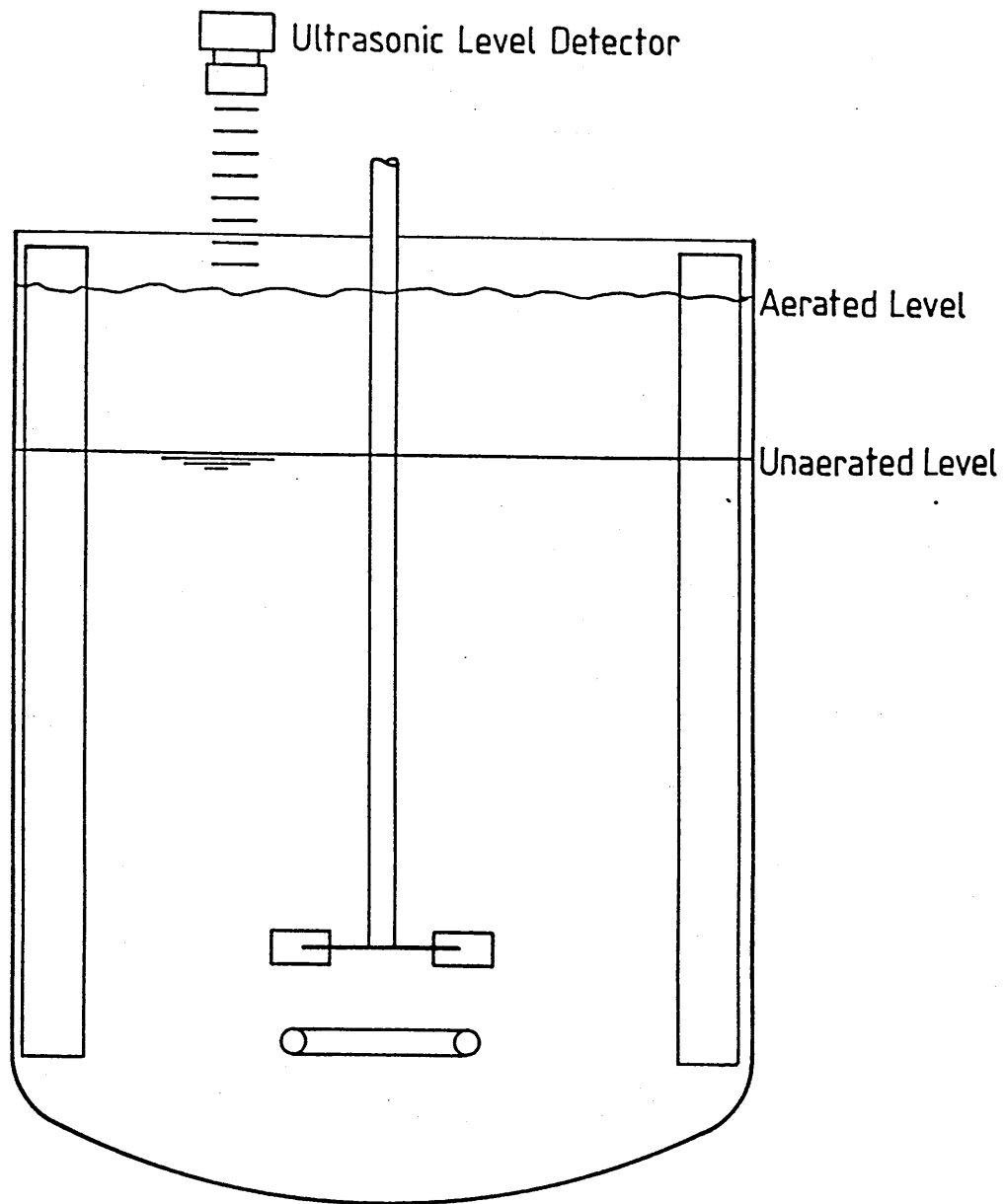




Collection device (plan)

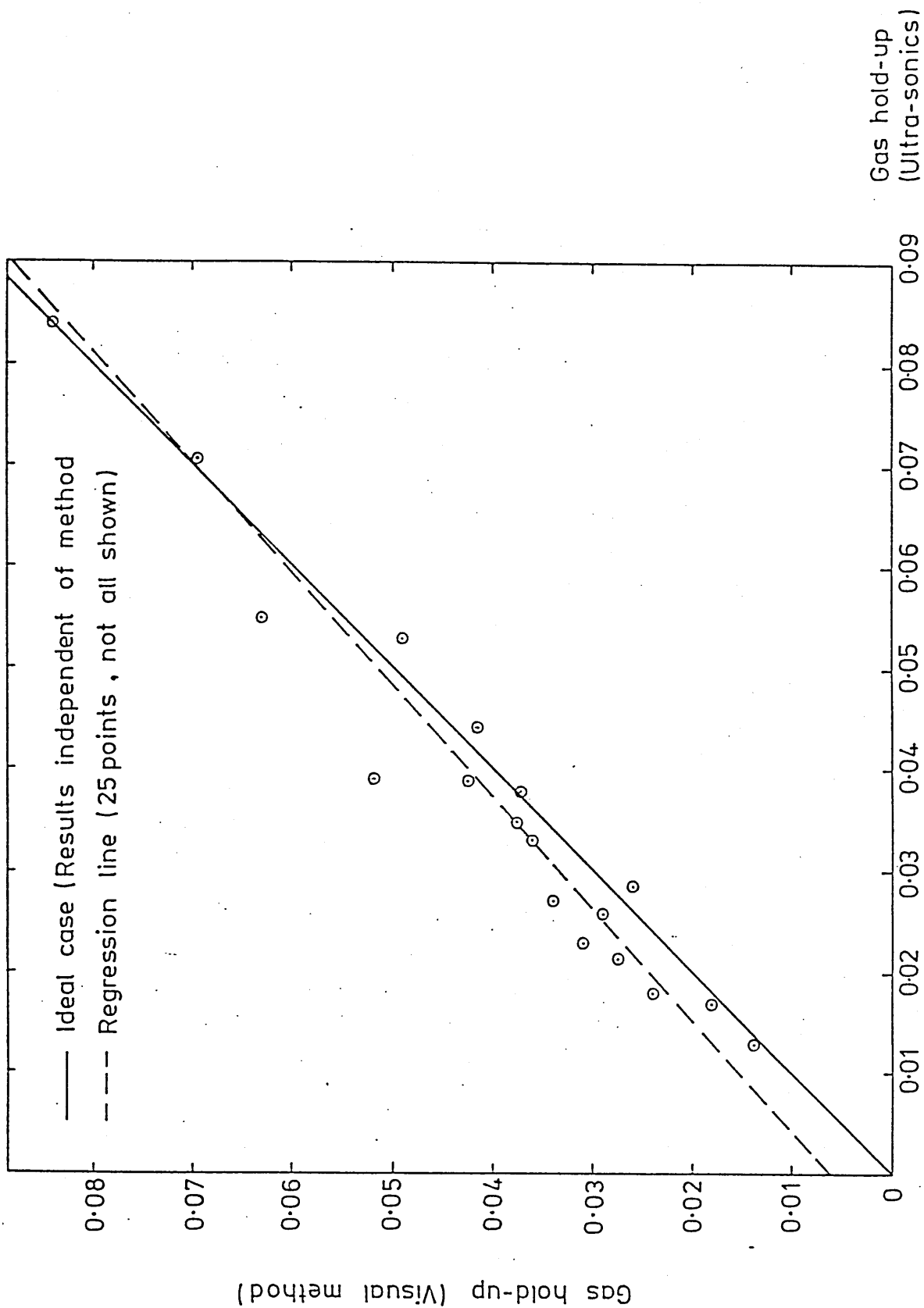
5.2

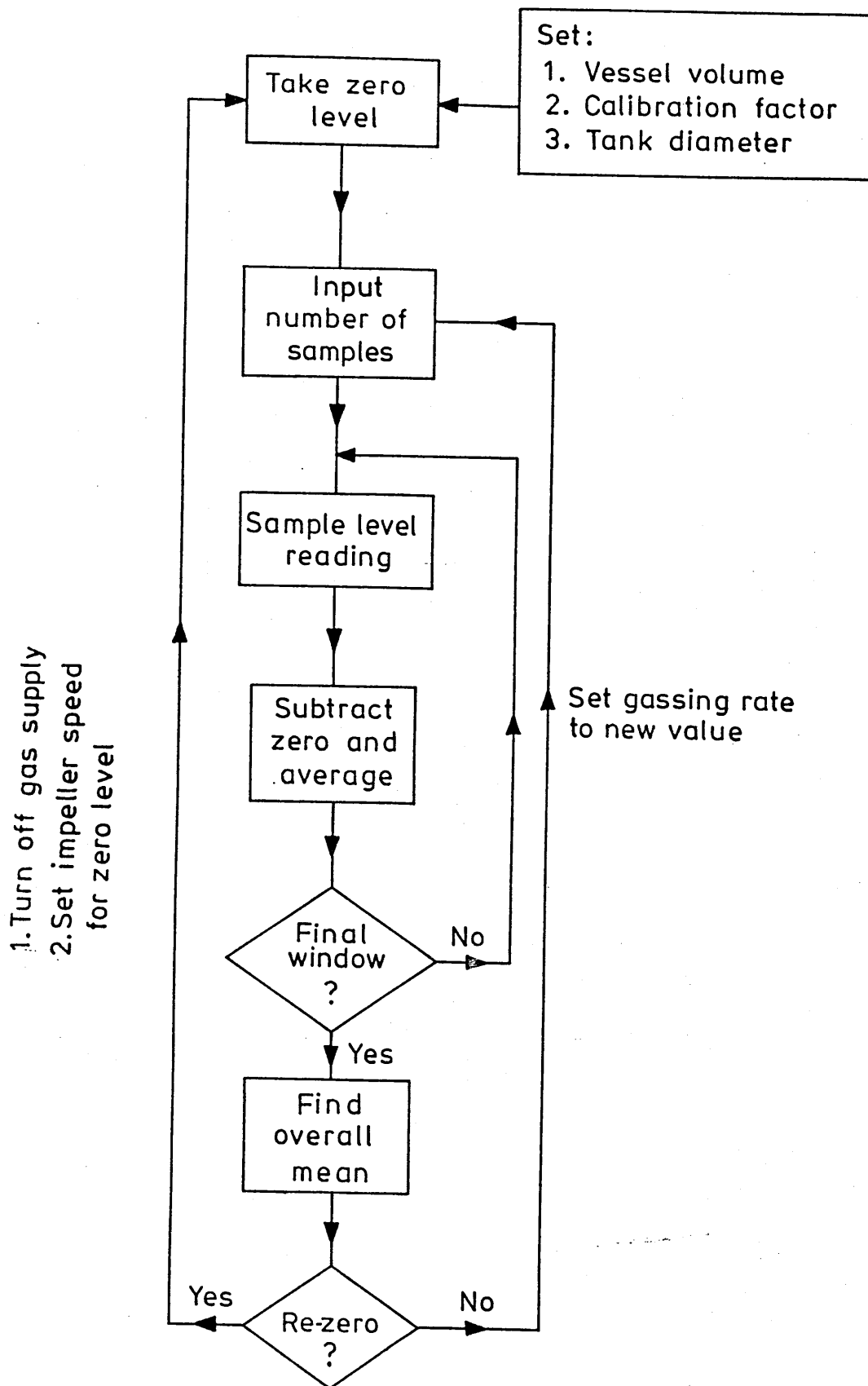
Spill-over method of measuring hold-up (2.67 m tank)

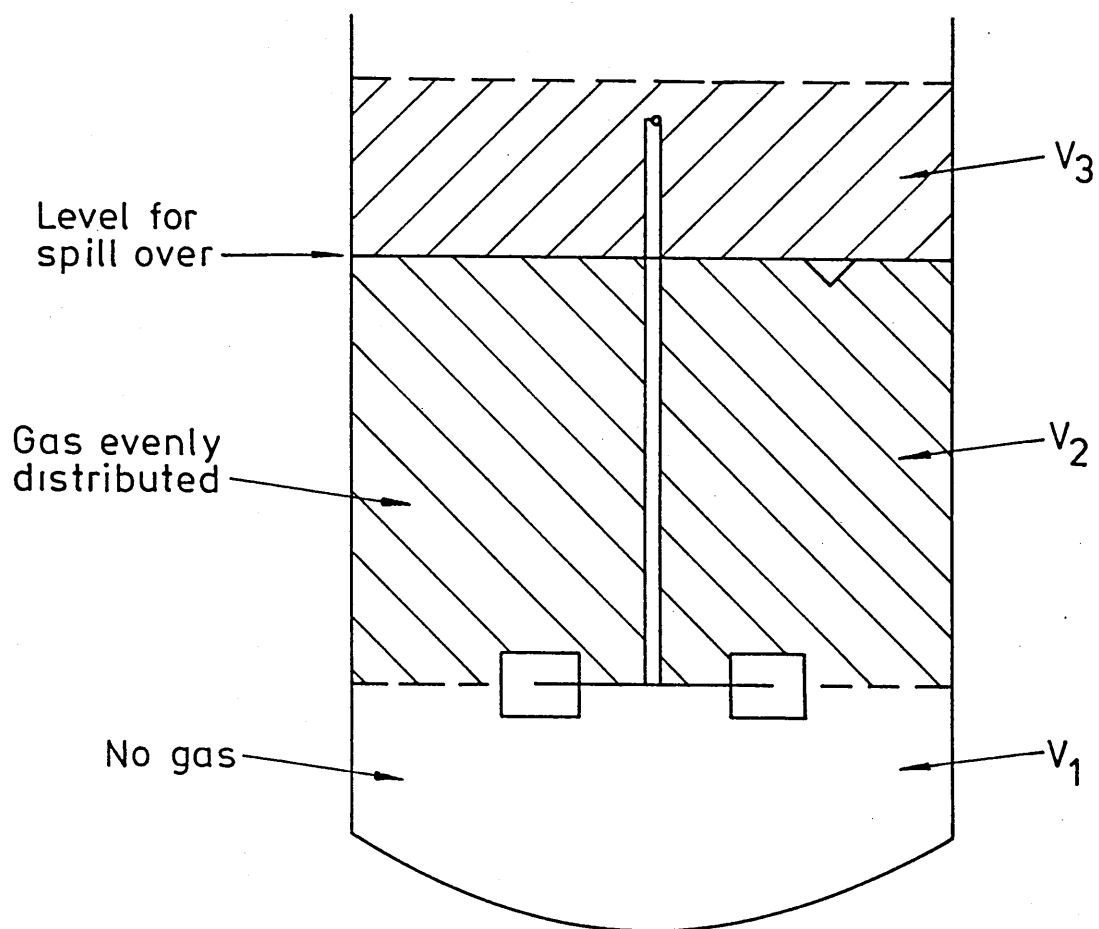


5.3

Comparison of ultra-sonic level detector with visual hold-up measurement

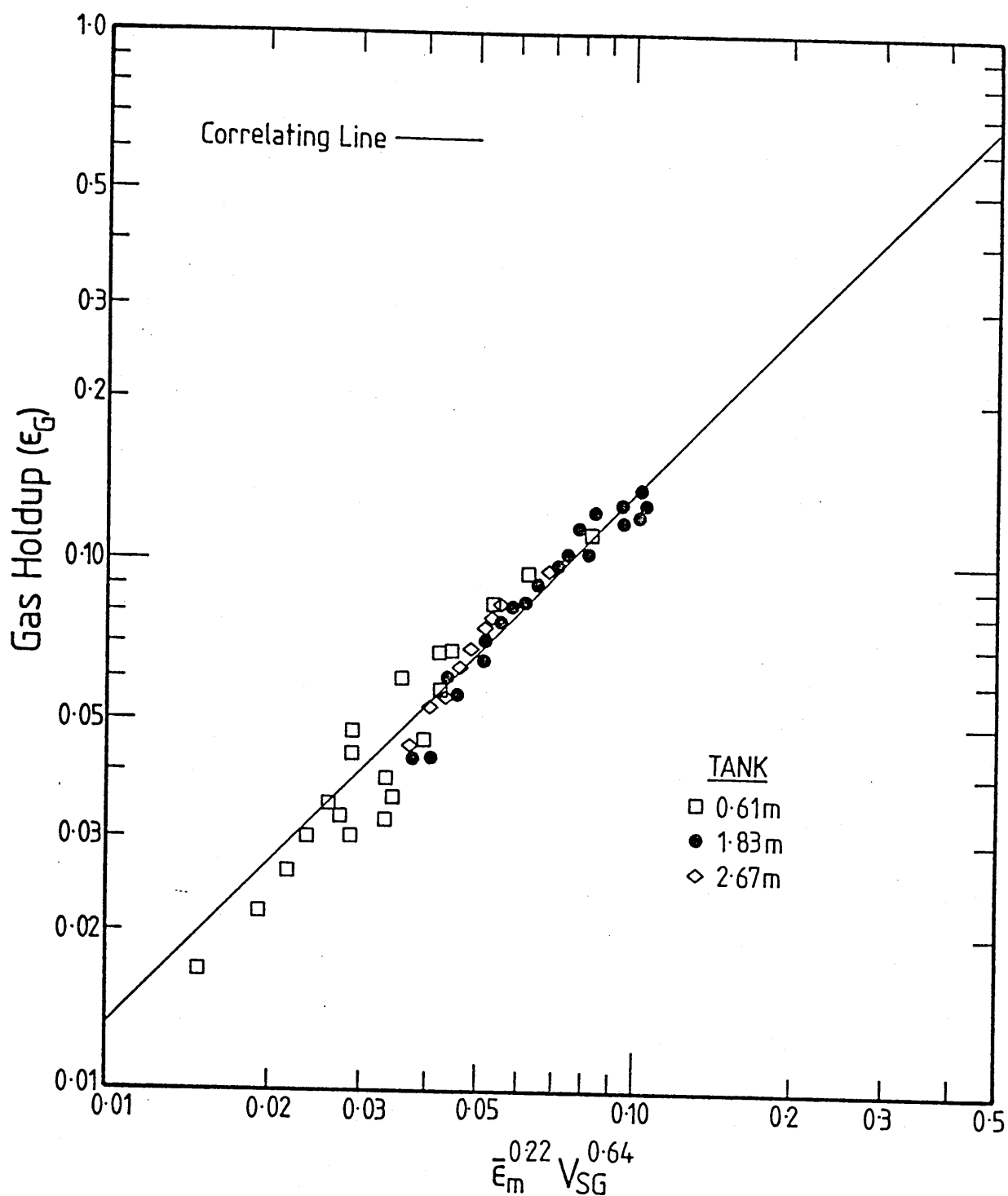






5.6

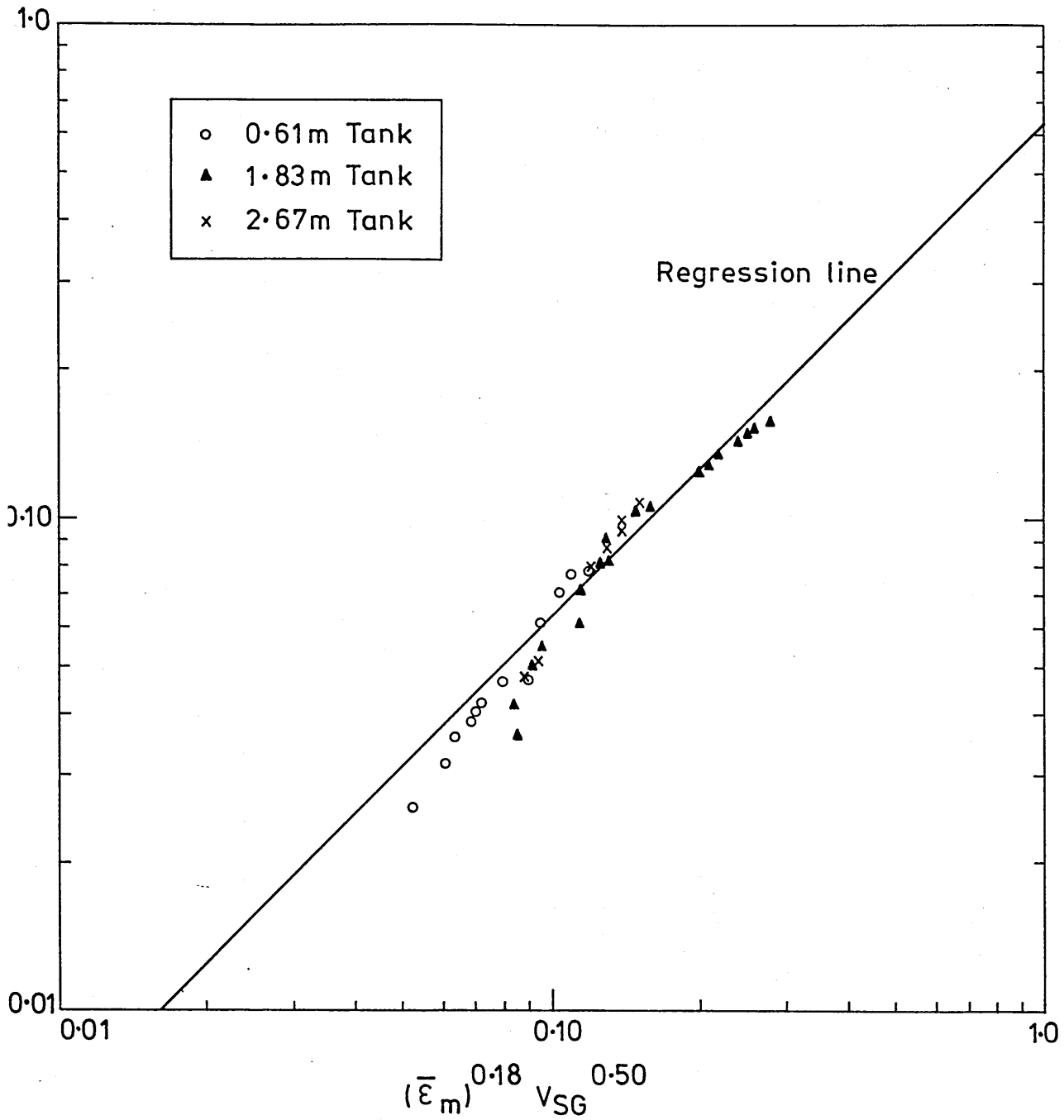
Distribution of gas in vessel

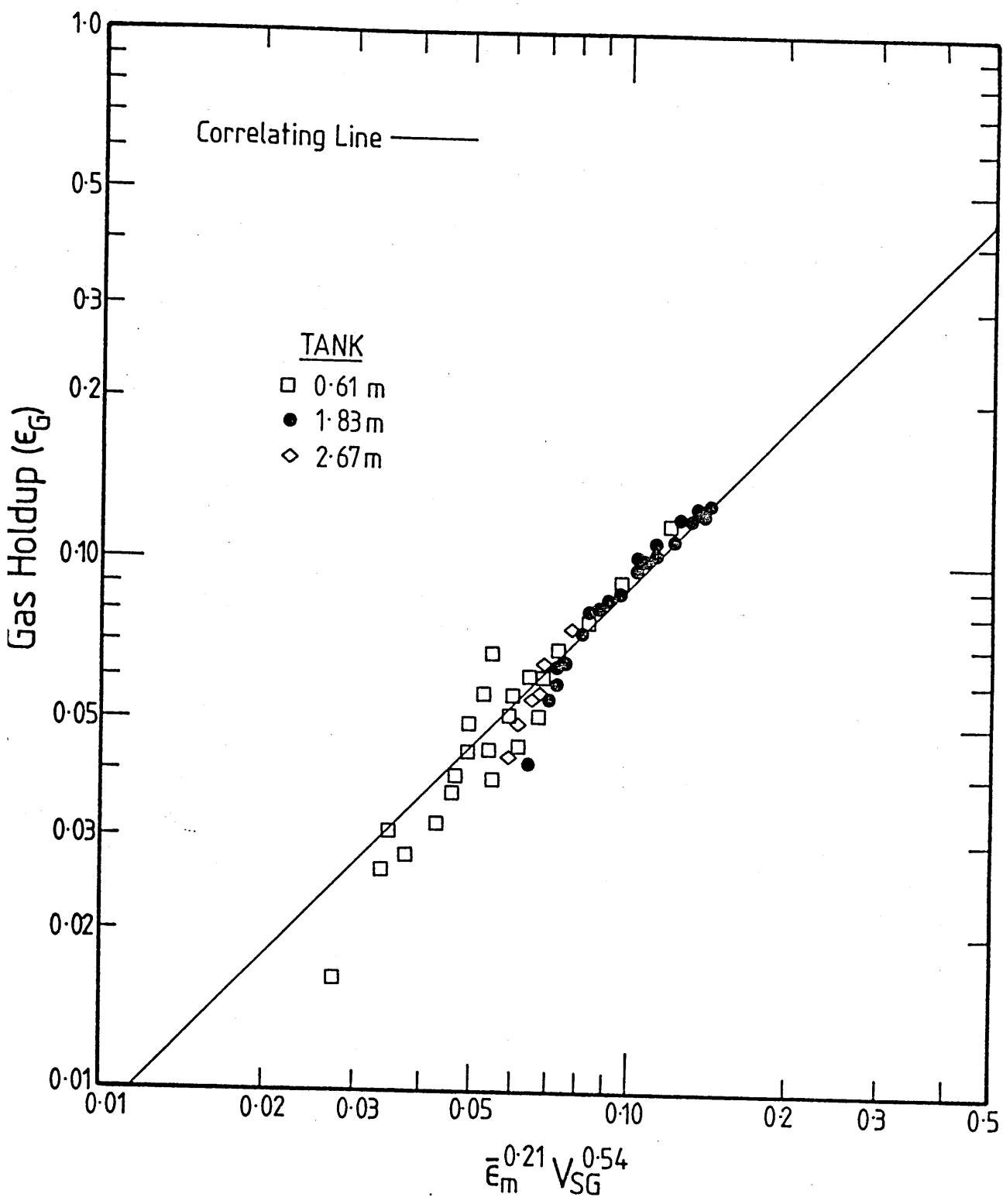


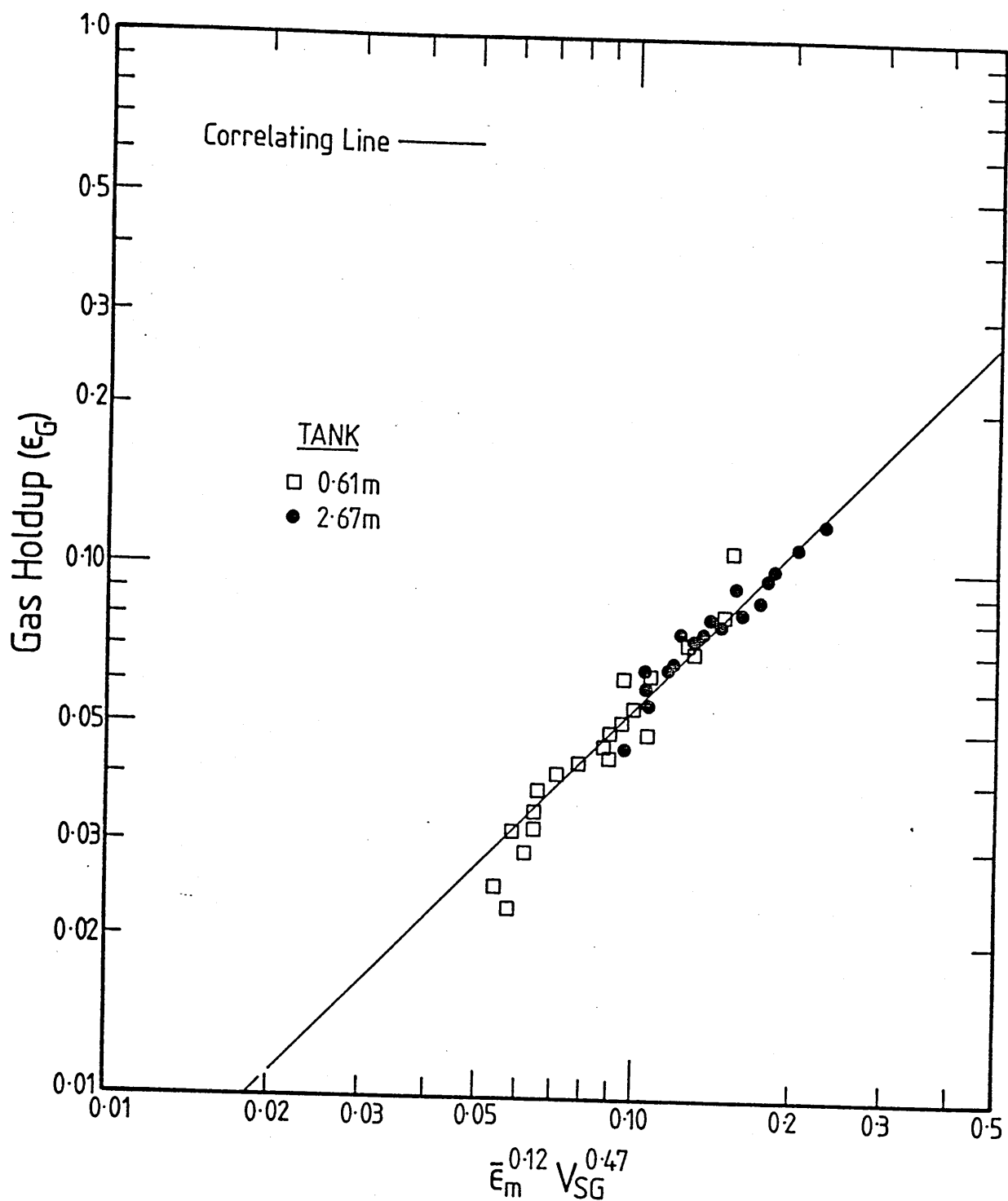
5.7

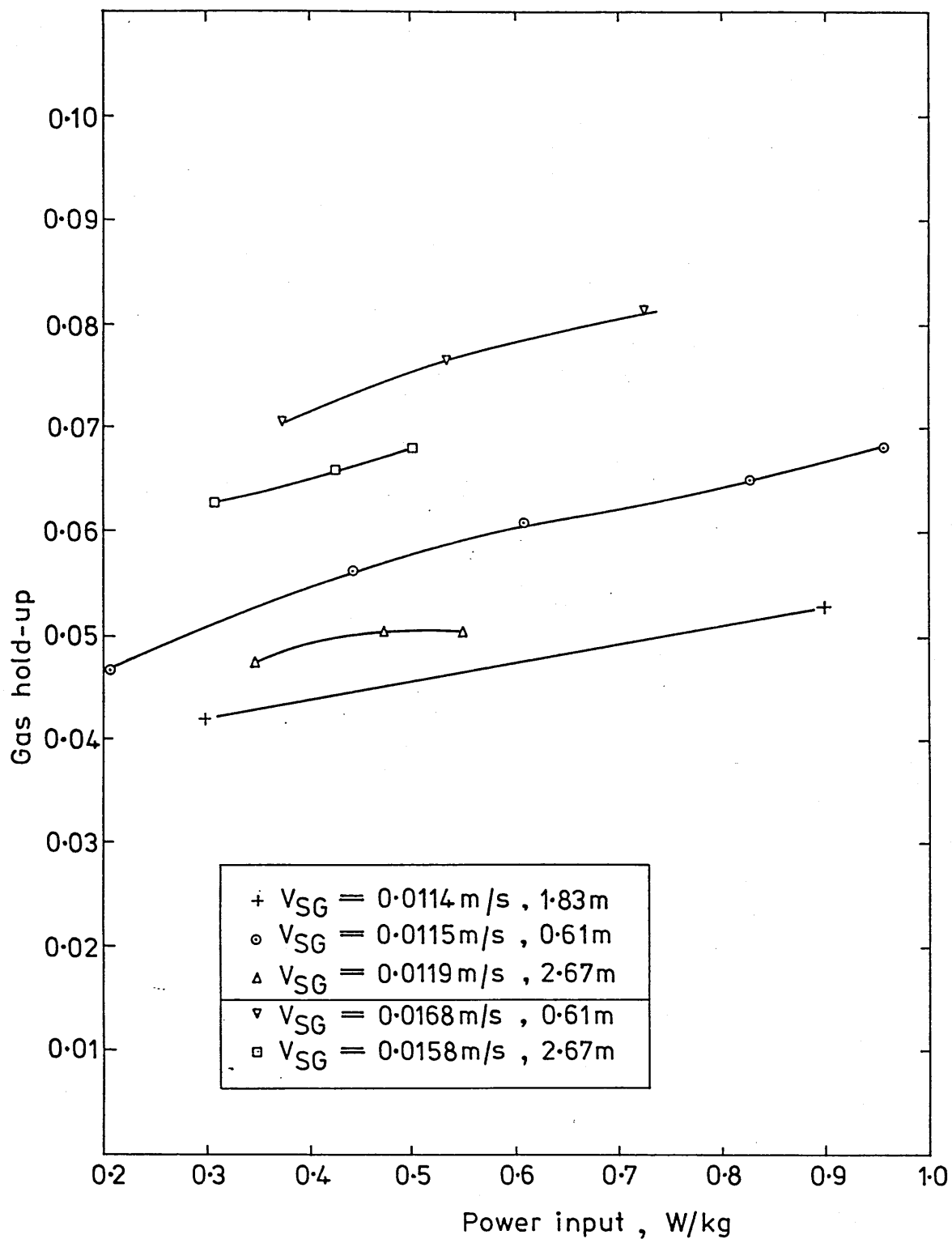
Hold-up data from all three scales: Rushton turbine: octanol solution

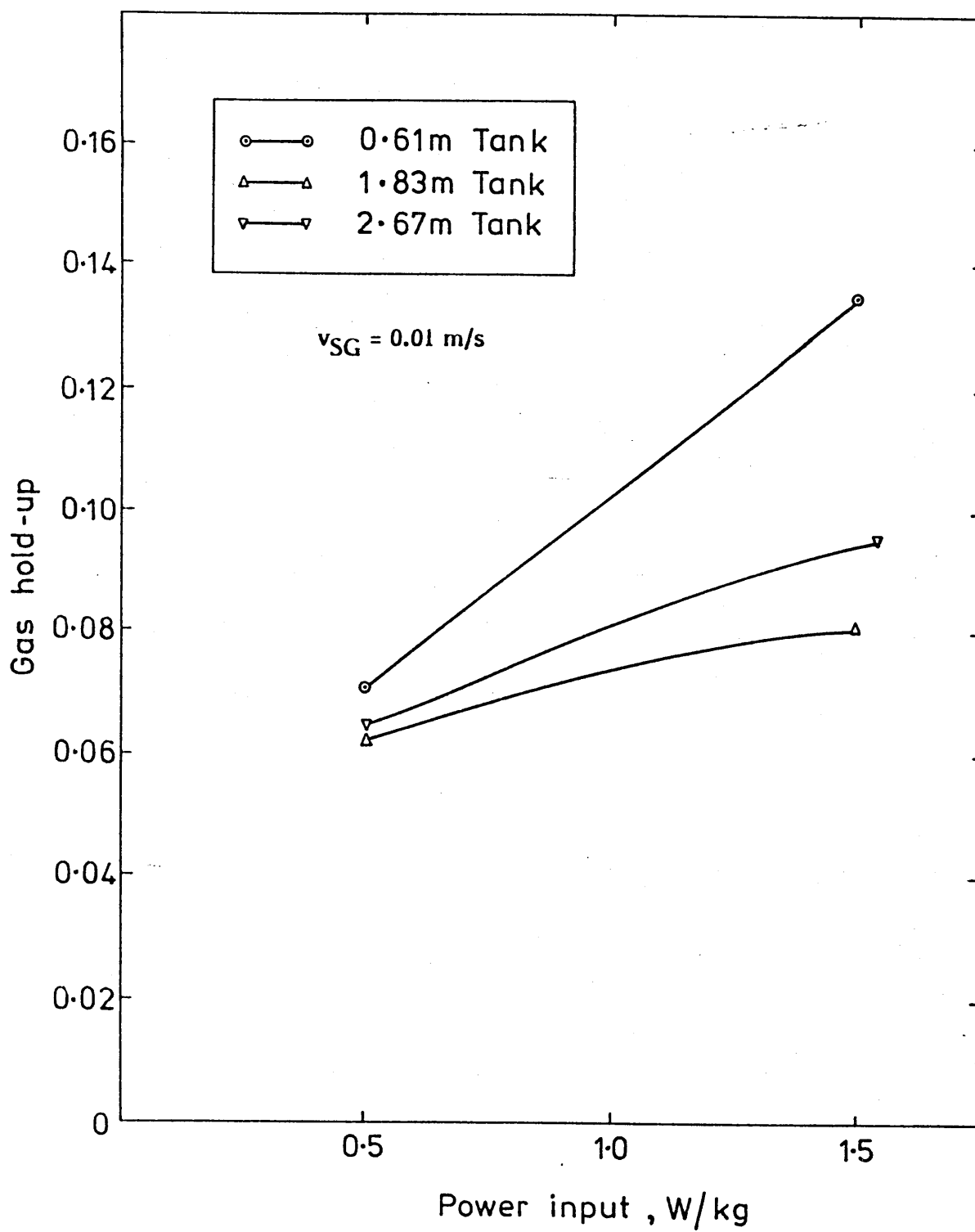
Gas hold-up

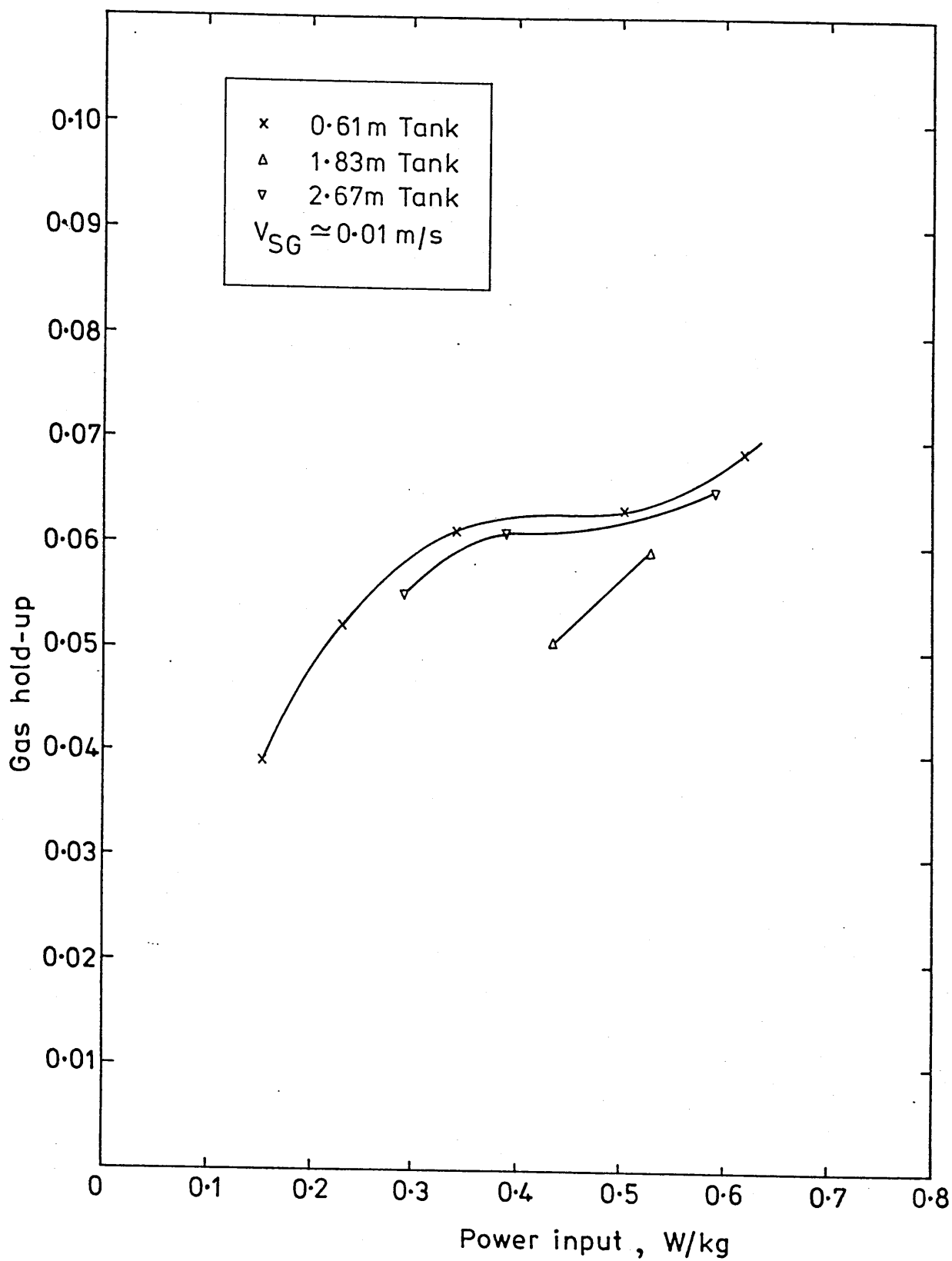


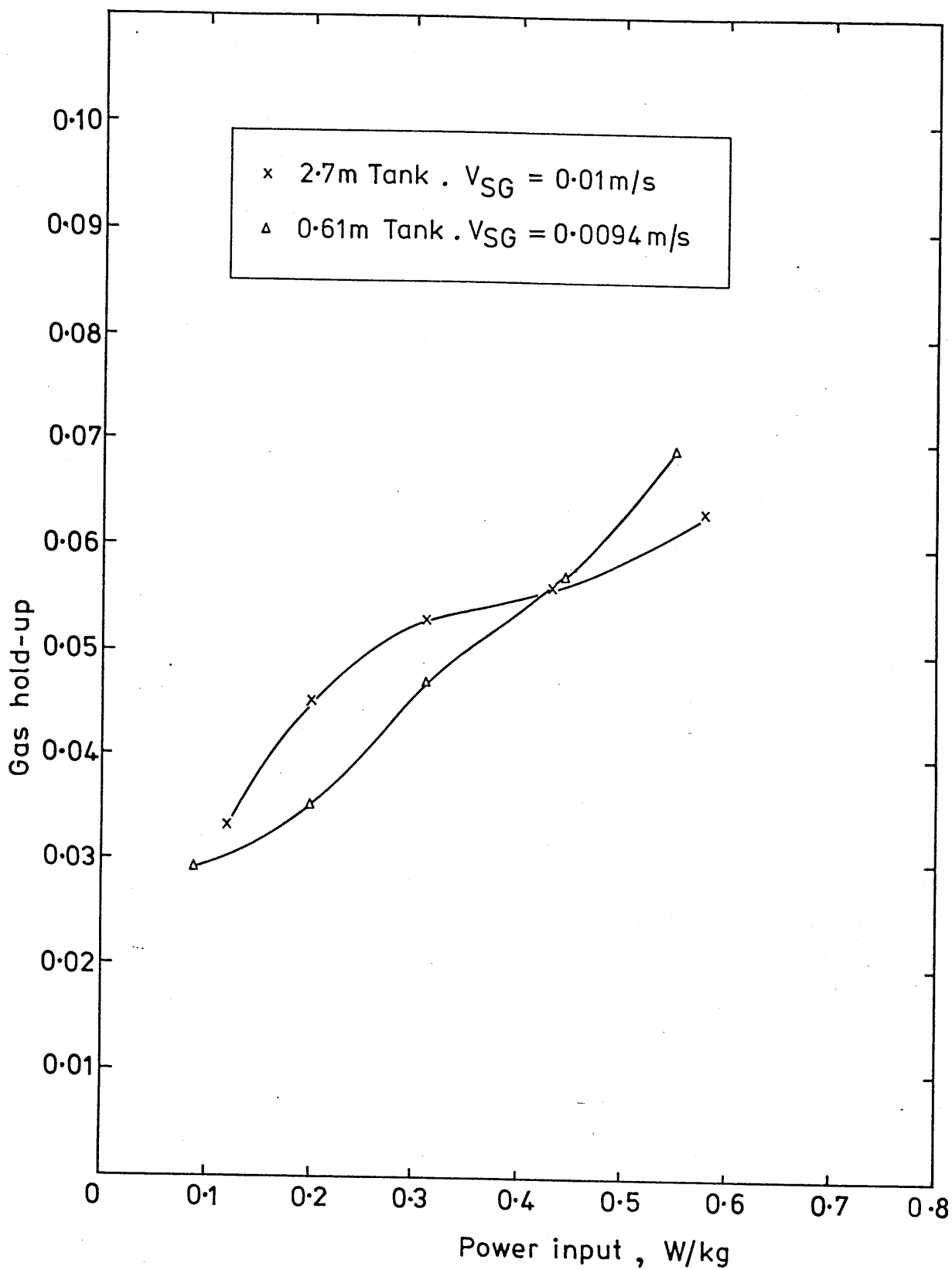


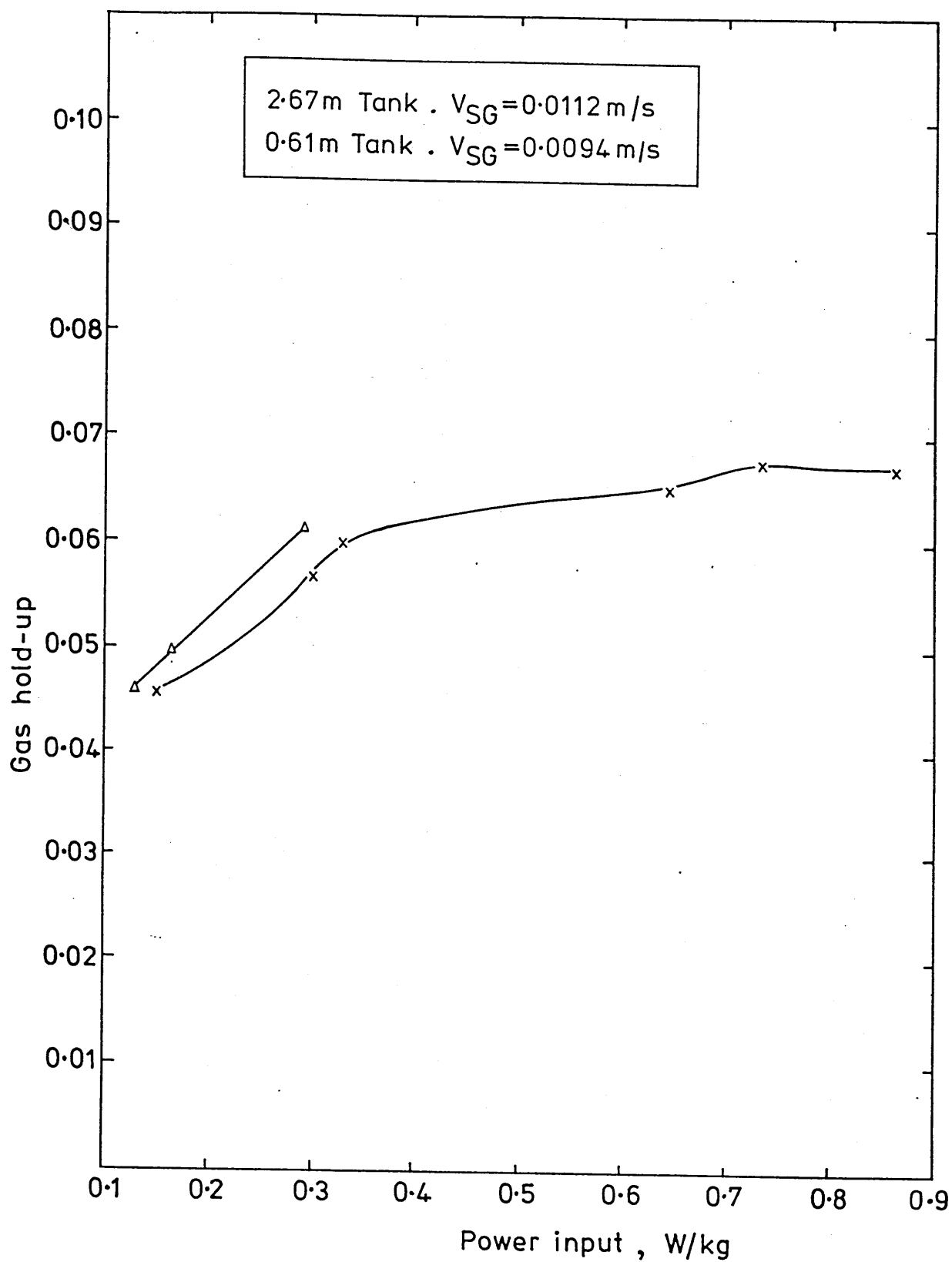


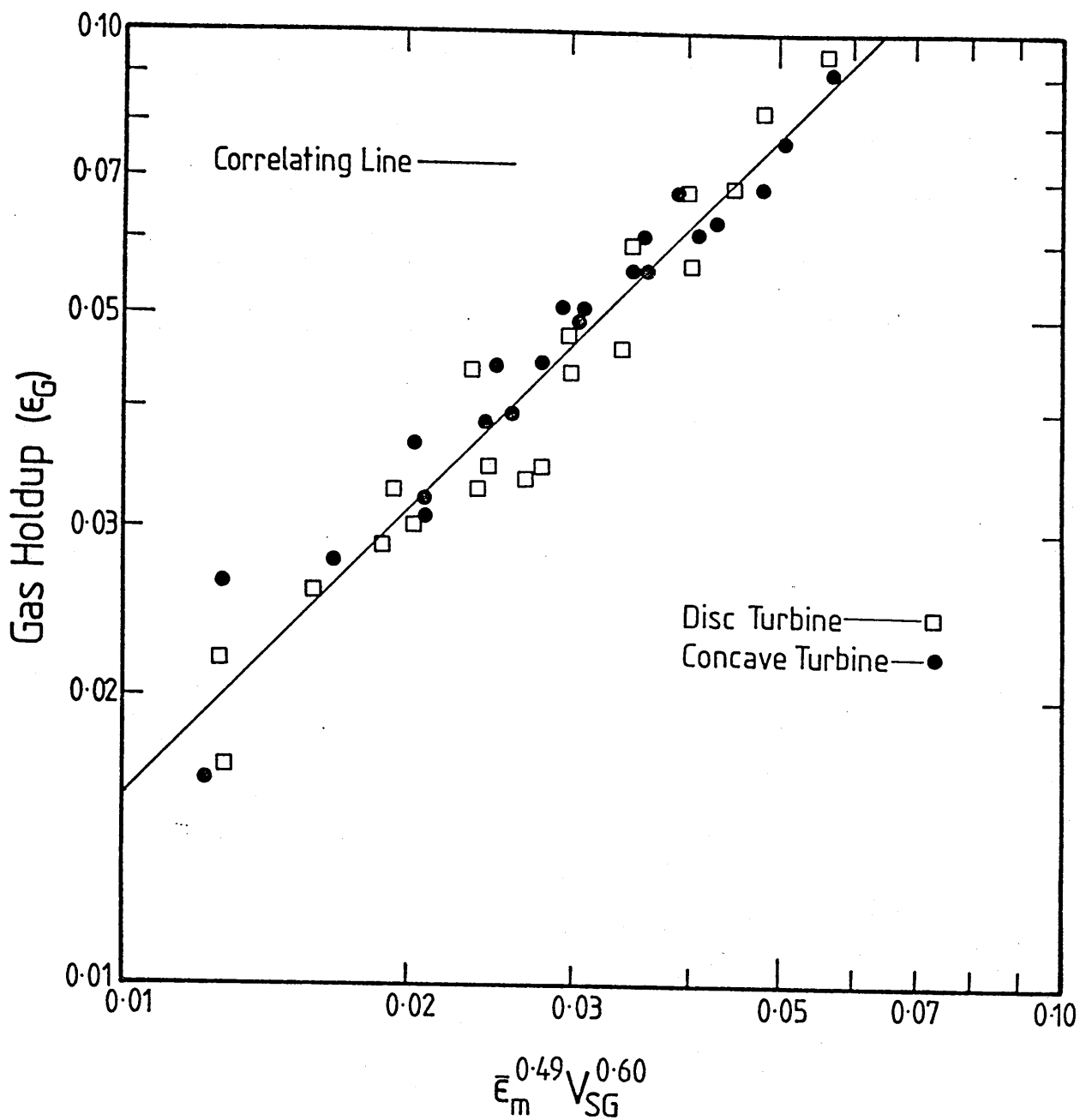






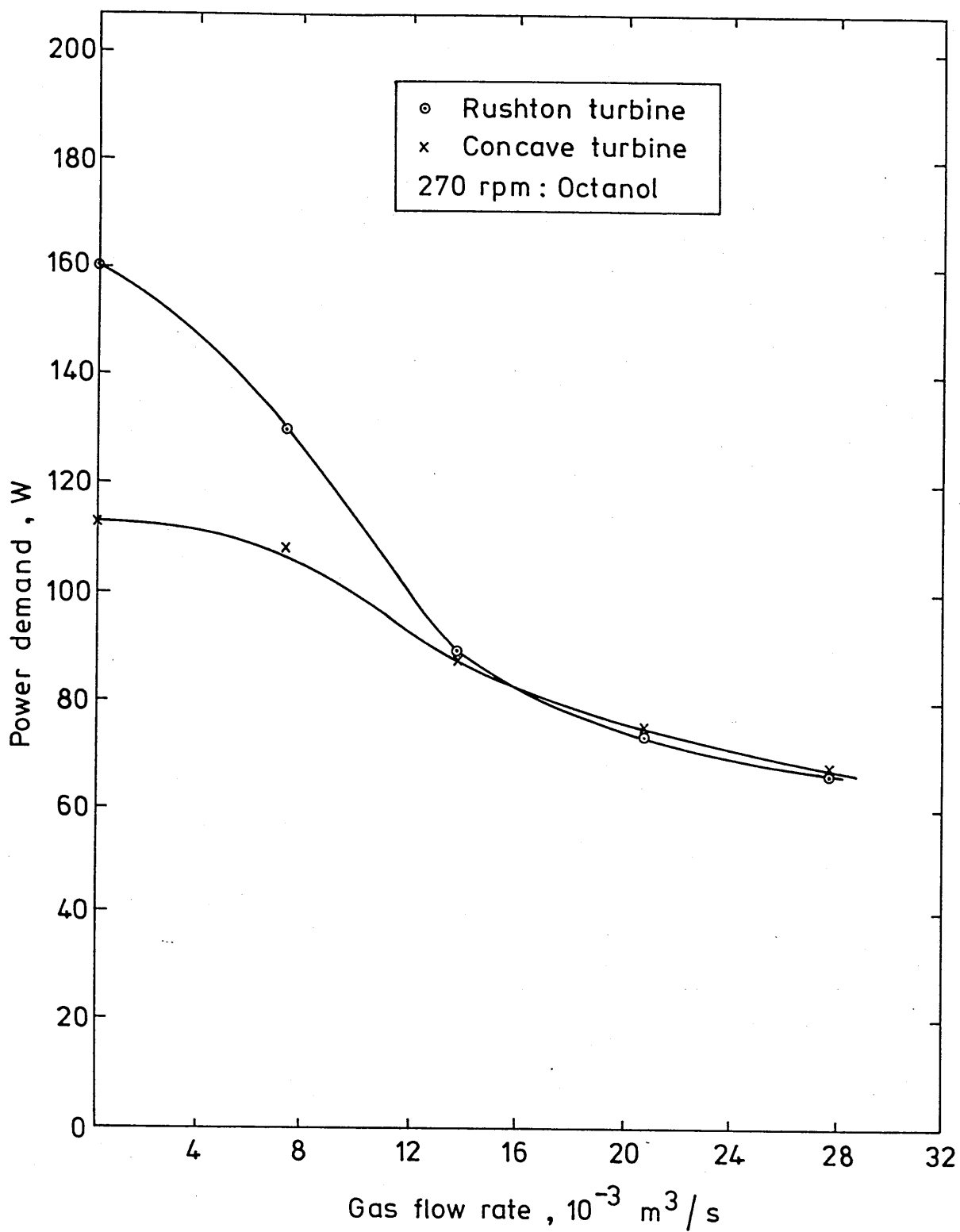


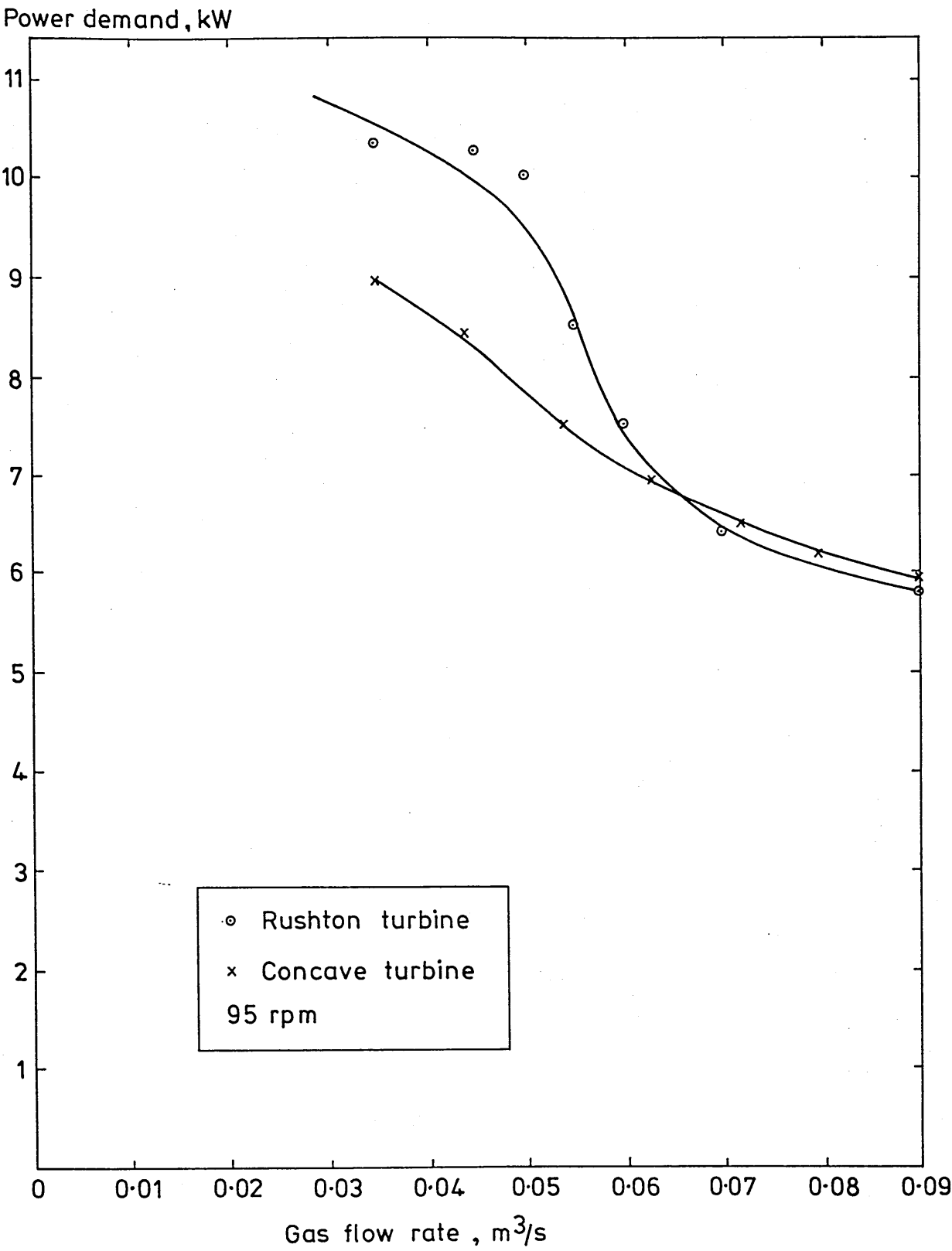


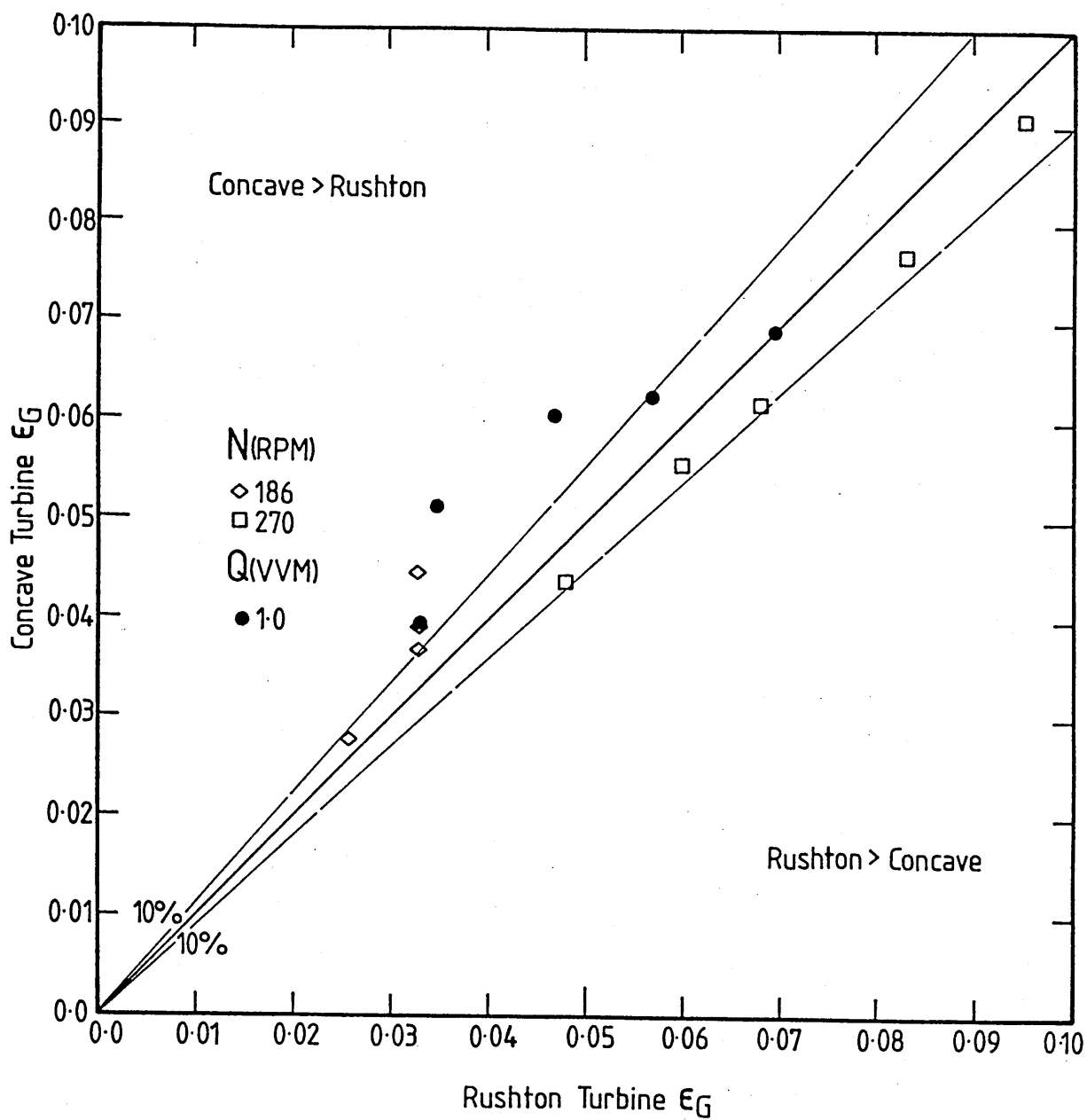


5.16

Combined Rushton turbine and concave turbine hold-up data:
0.61 m tank

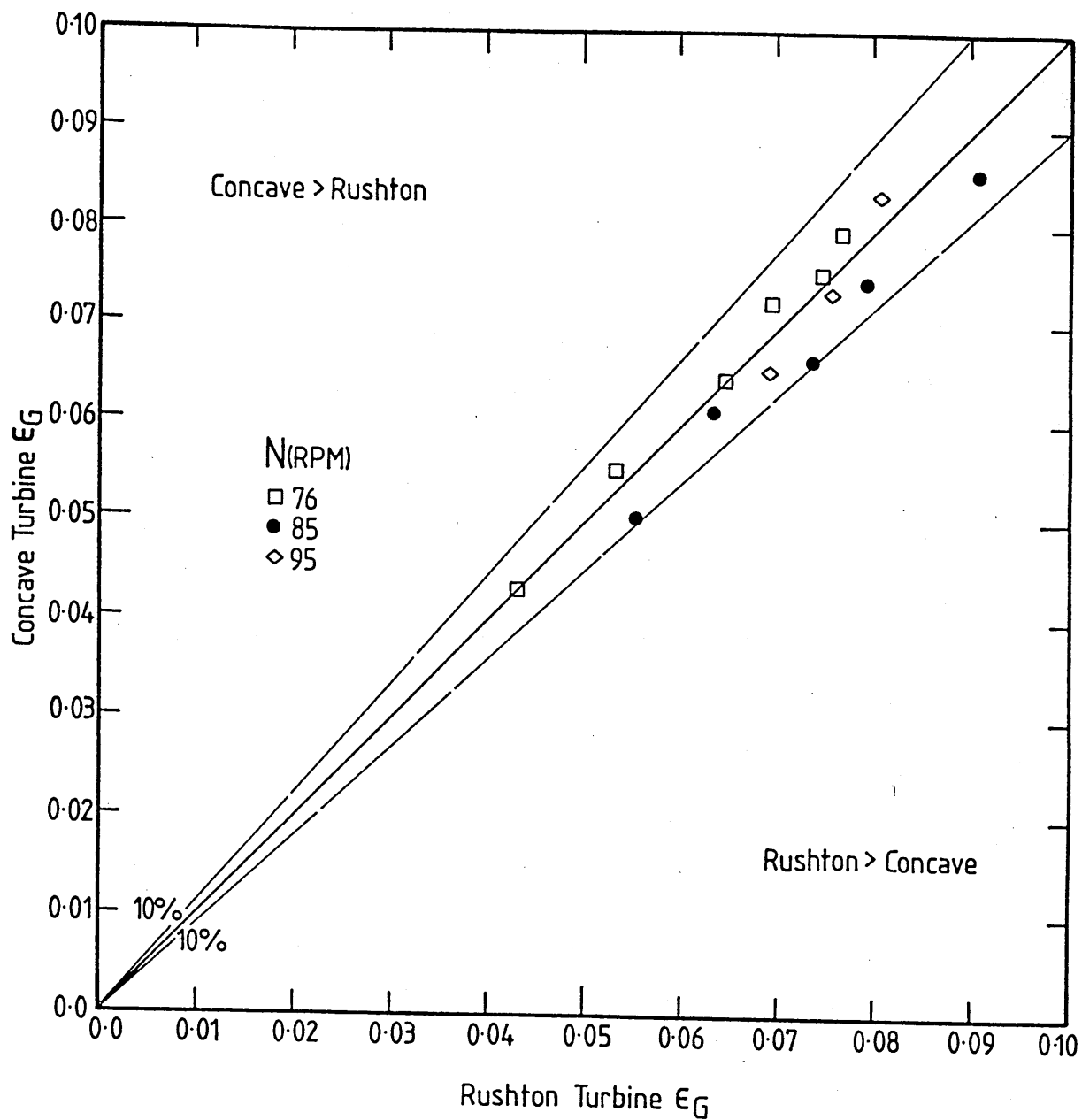






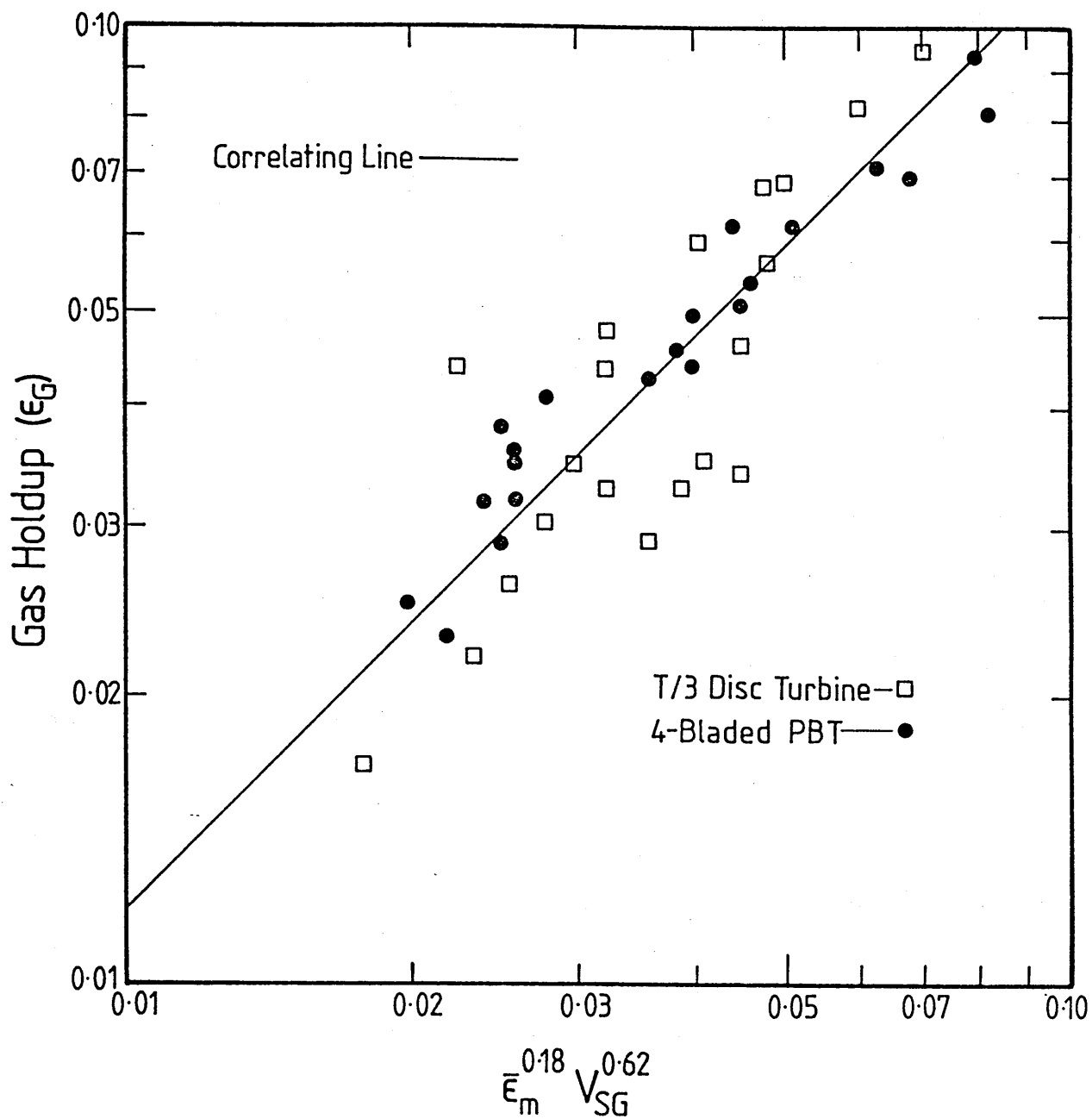
5.19

Comparison of hold-up data at identical operating conditions:
 0.61 m tank



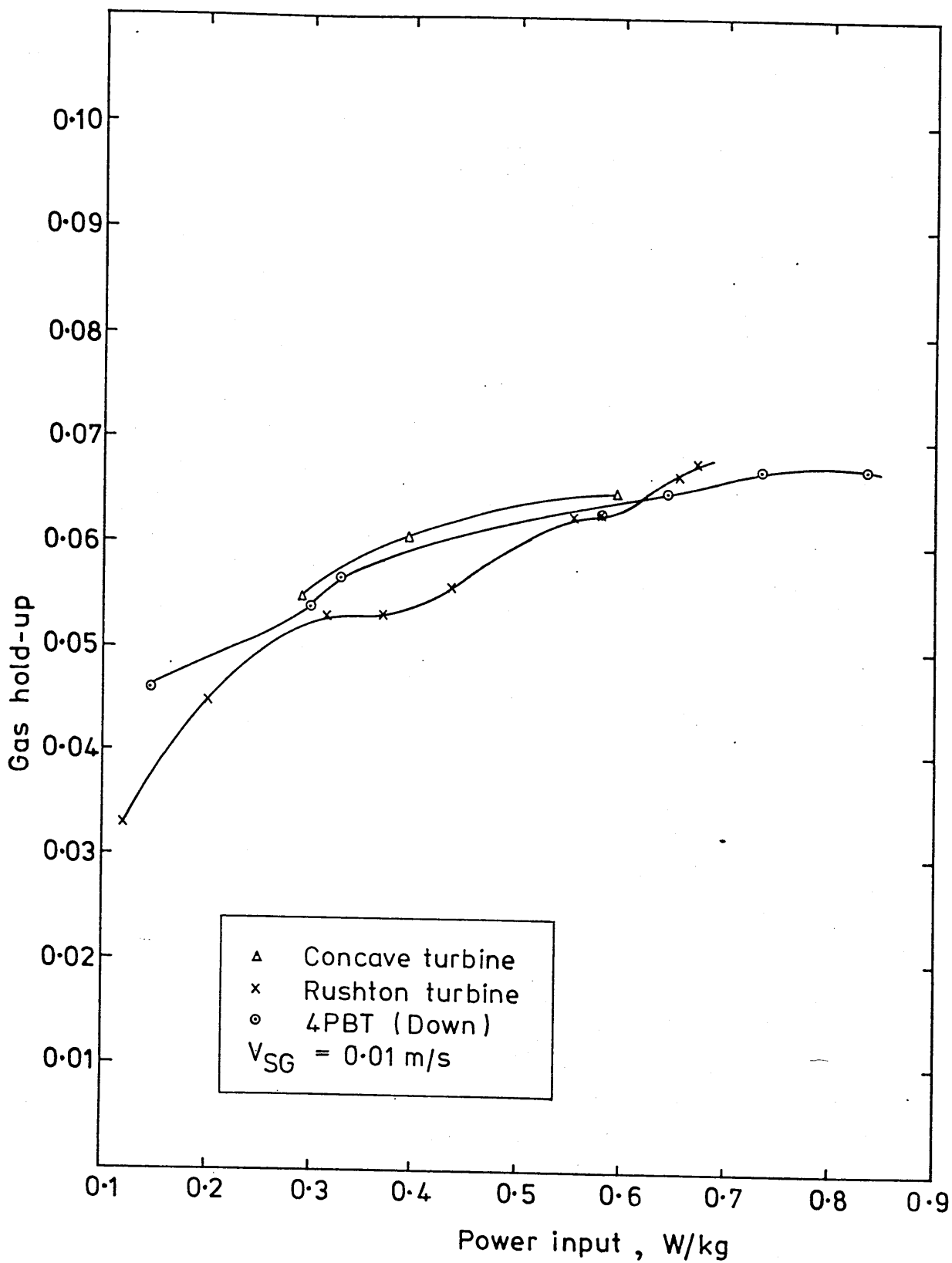
5.20

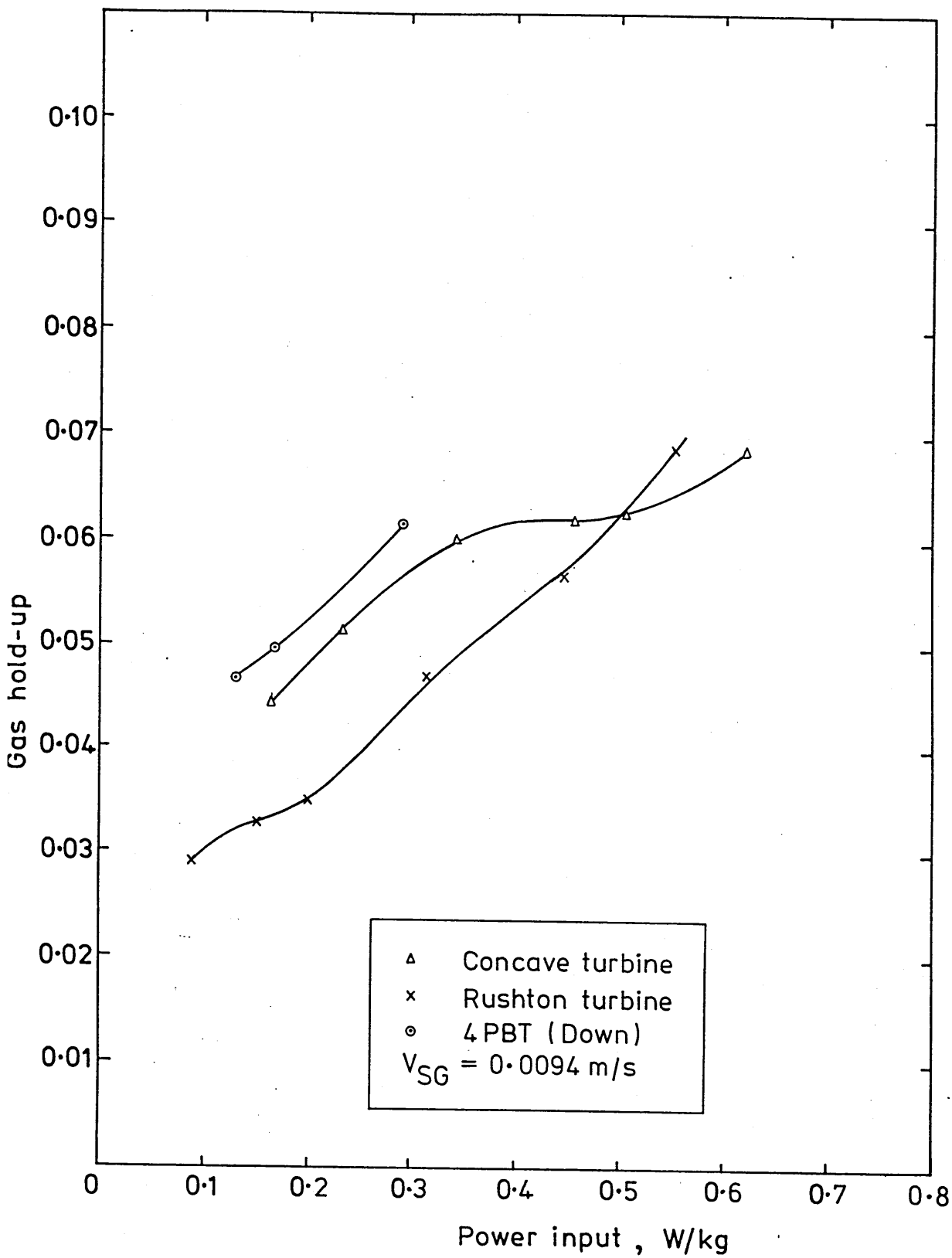
Comparison of hold-up data at identical operating conditions:
2.67 m tank

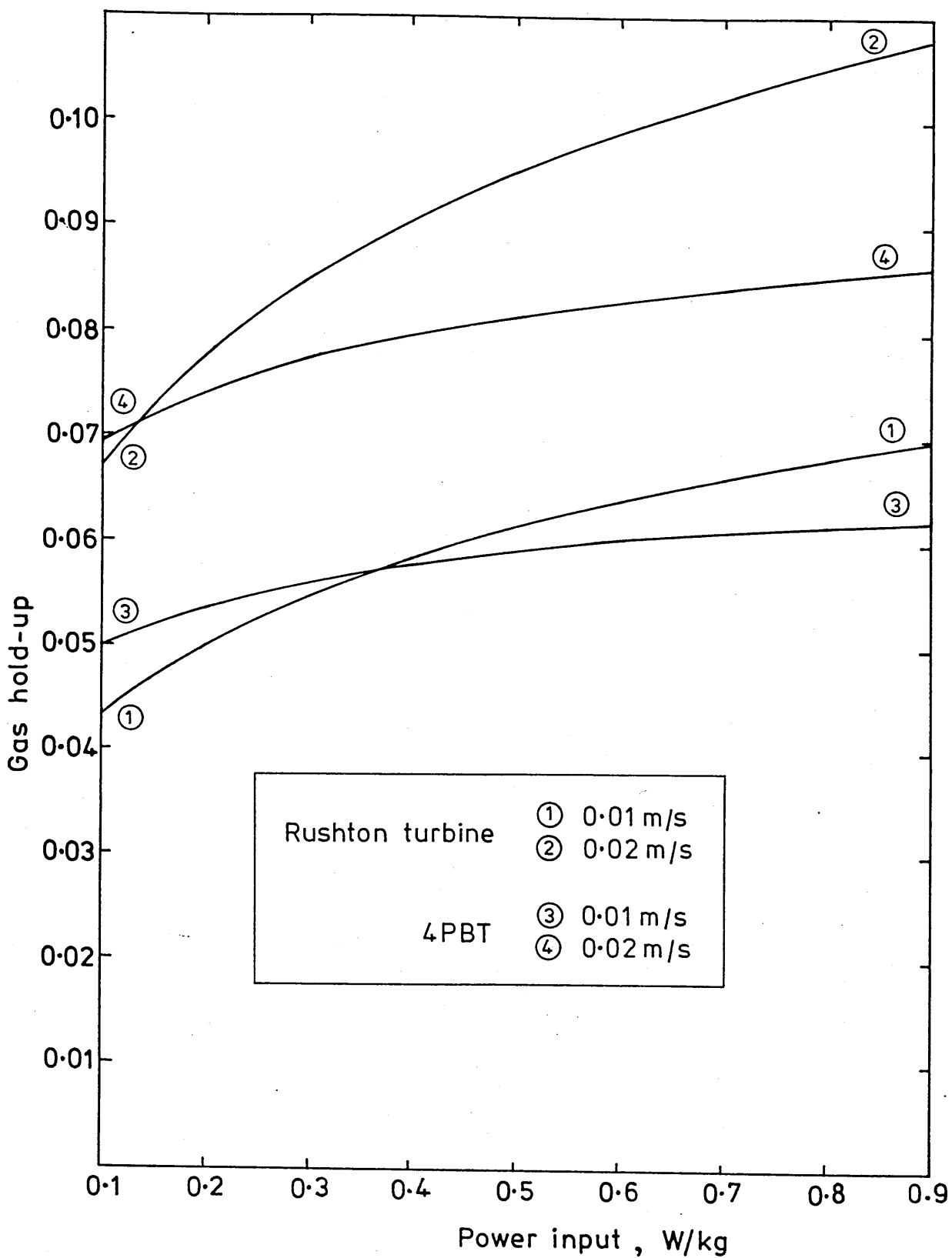


5.21

Combined 4BPT and Rushton turbine hold-up data: 0.61 m tank:
octanol solution

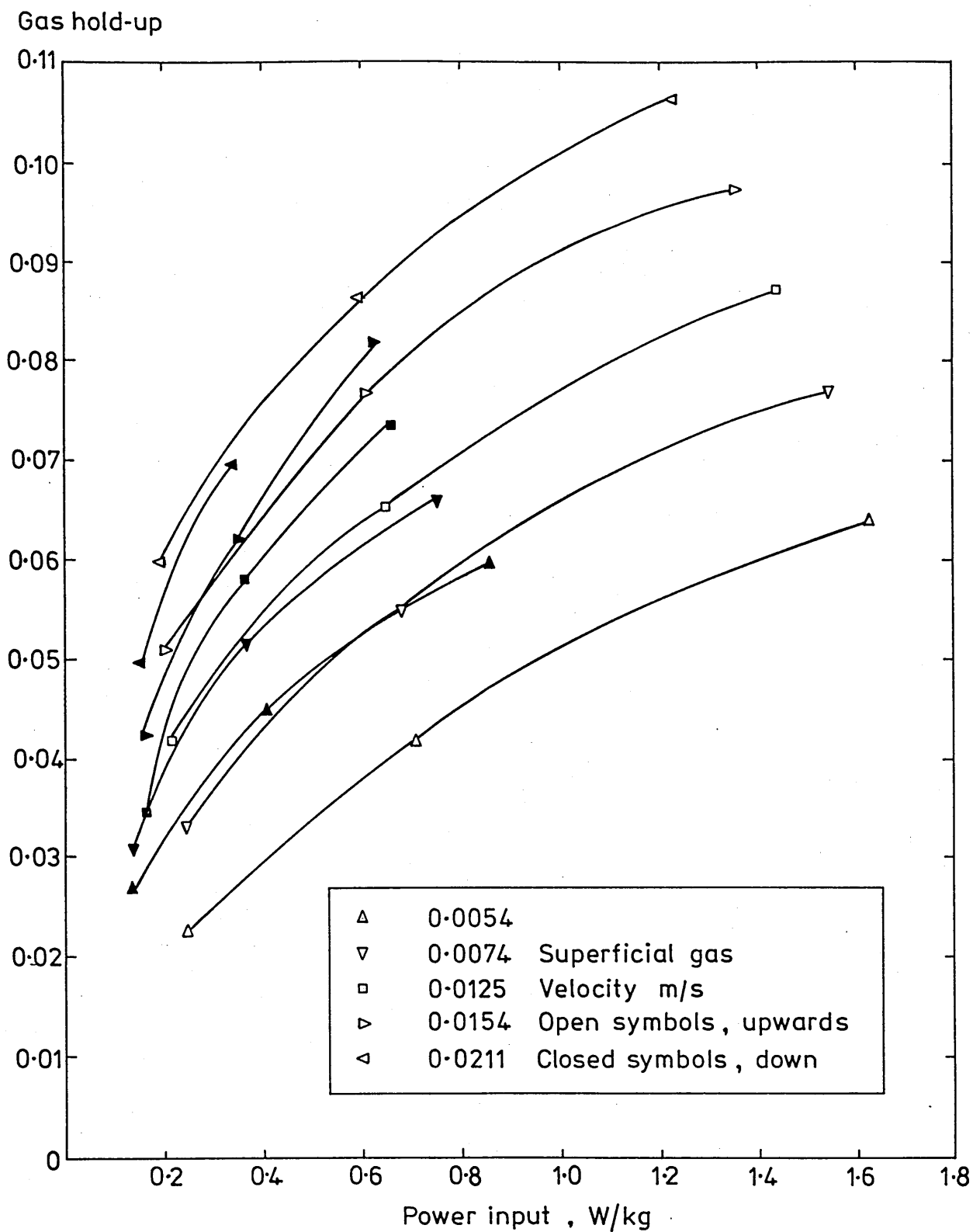


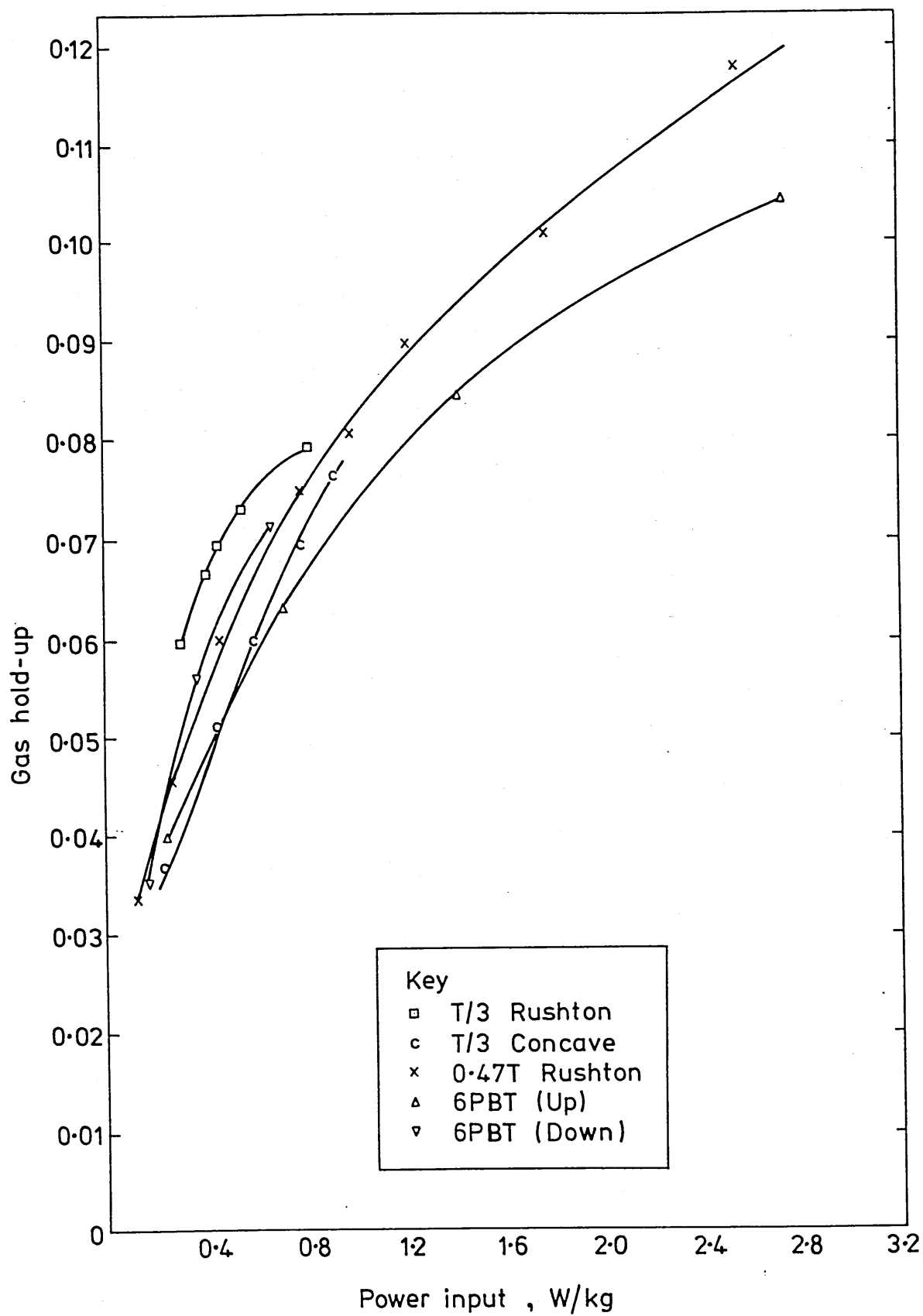


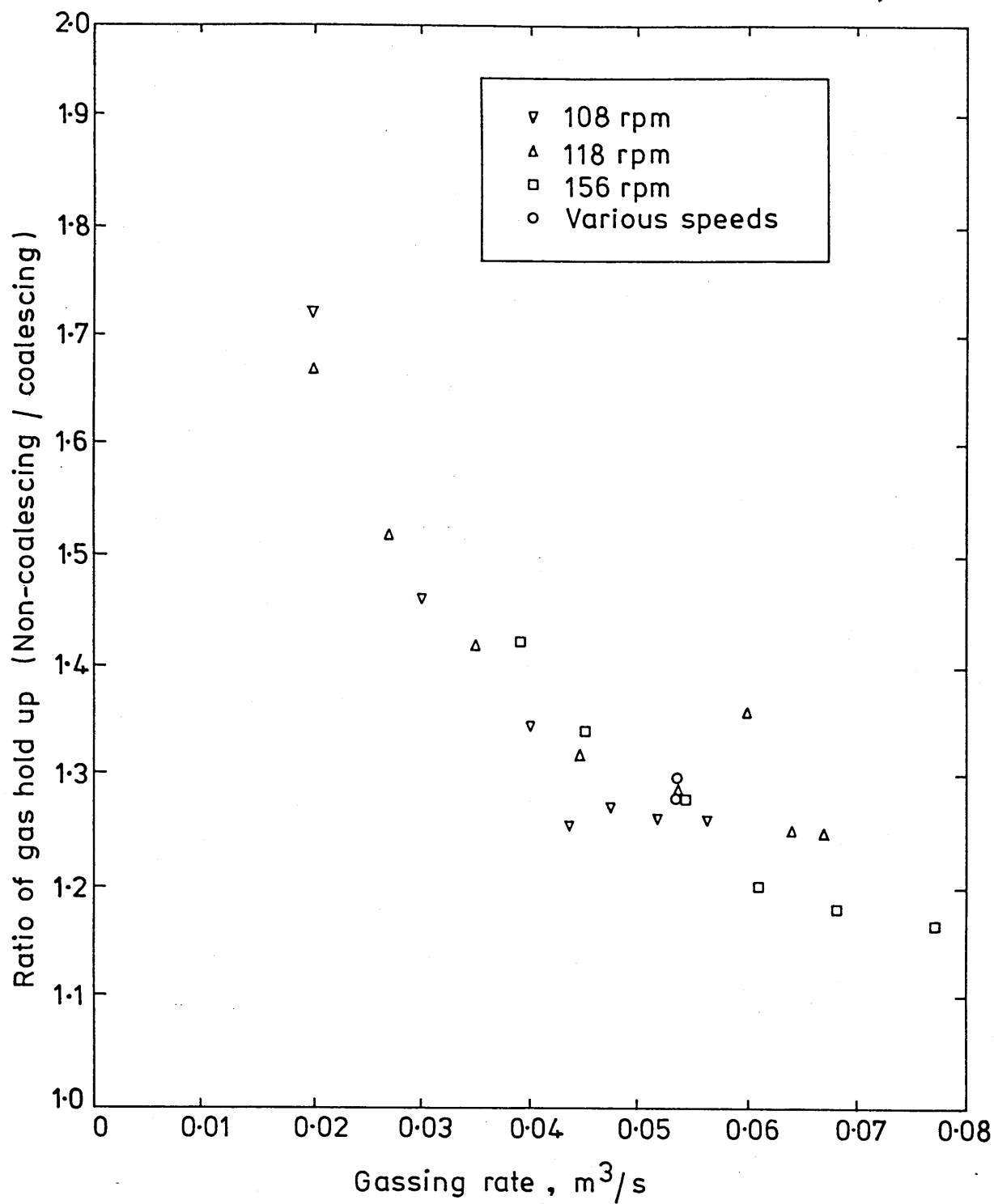


5.24

Comparison of correlation predictions: 4PBT and disc turbine:
combined correlations for 0.61 m and 2.67 m tanks

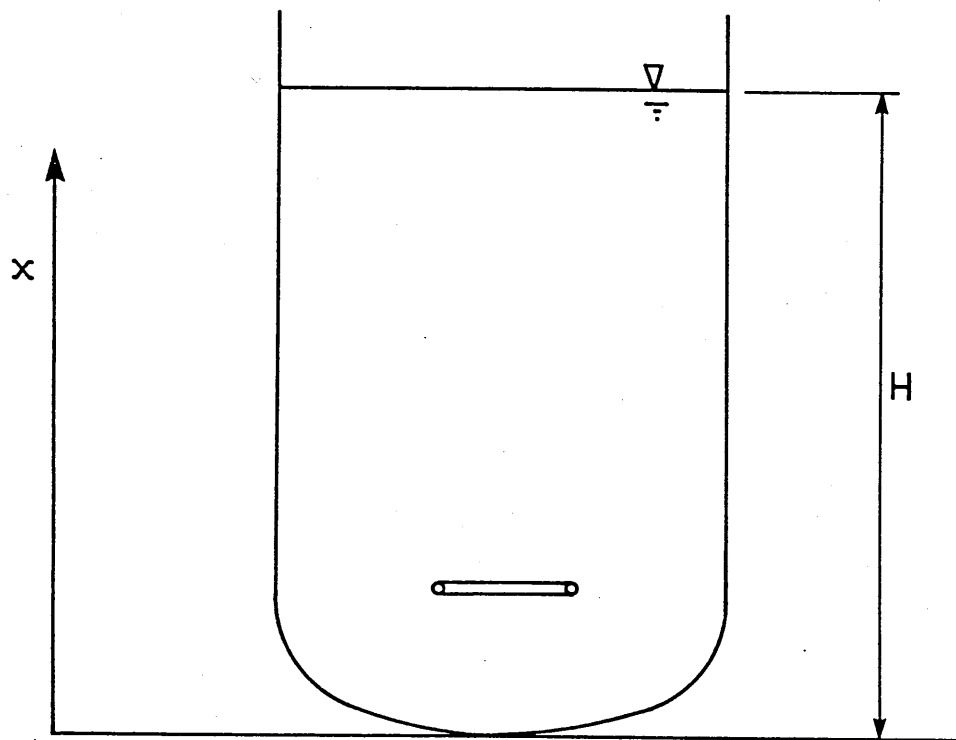






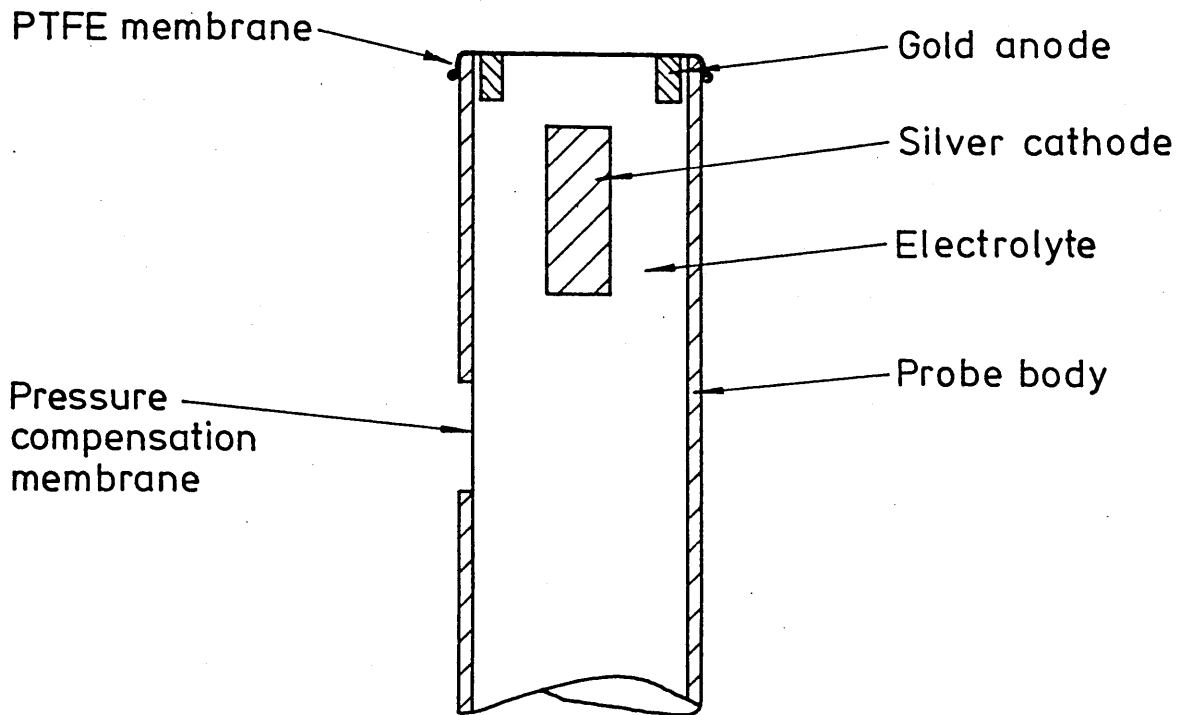
5.27

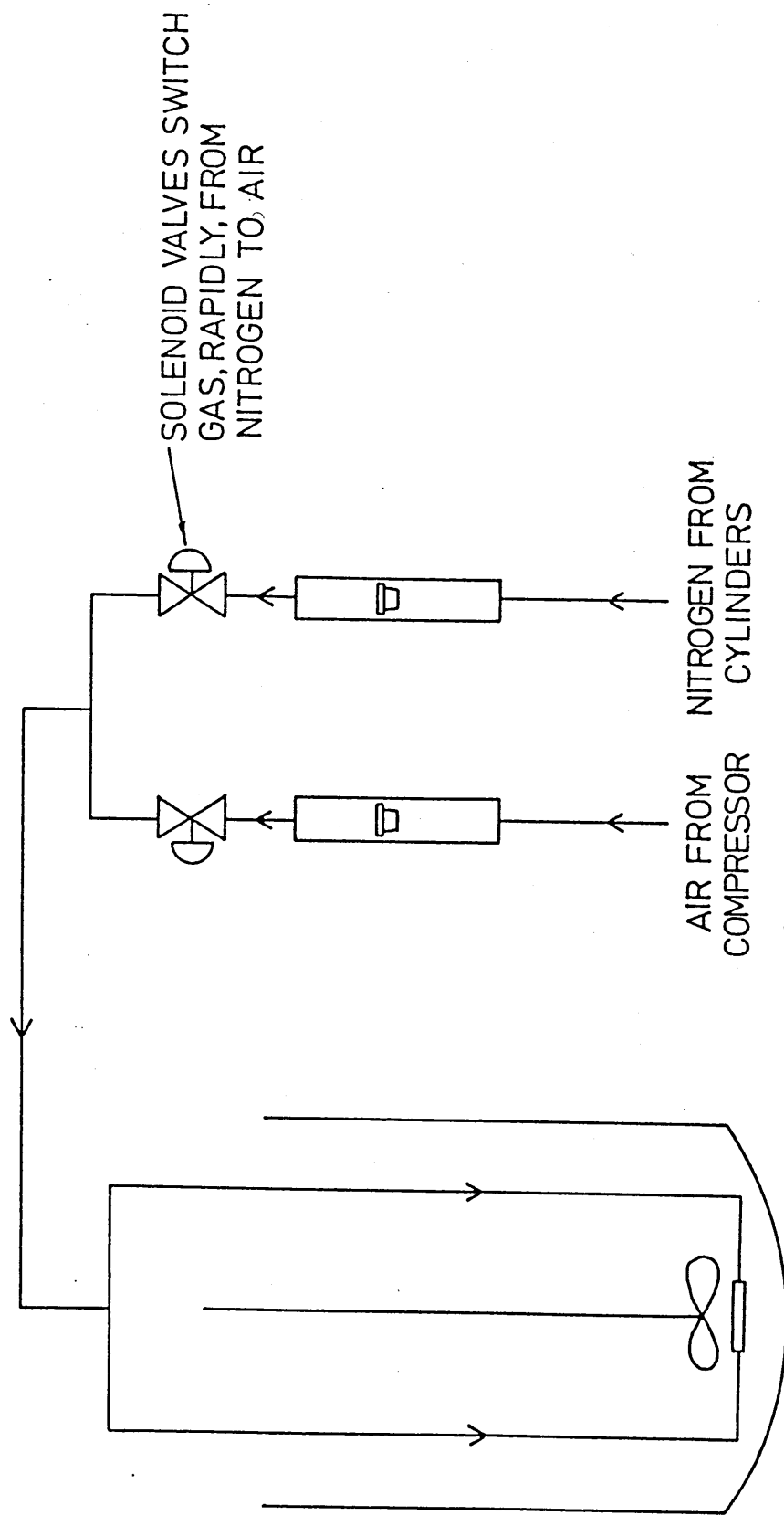
Ratio of hold-up in non-coalescing solution to hold-up in coalescing solution as function of gas flowrate: 1.83 m tank



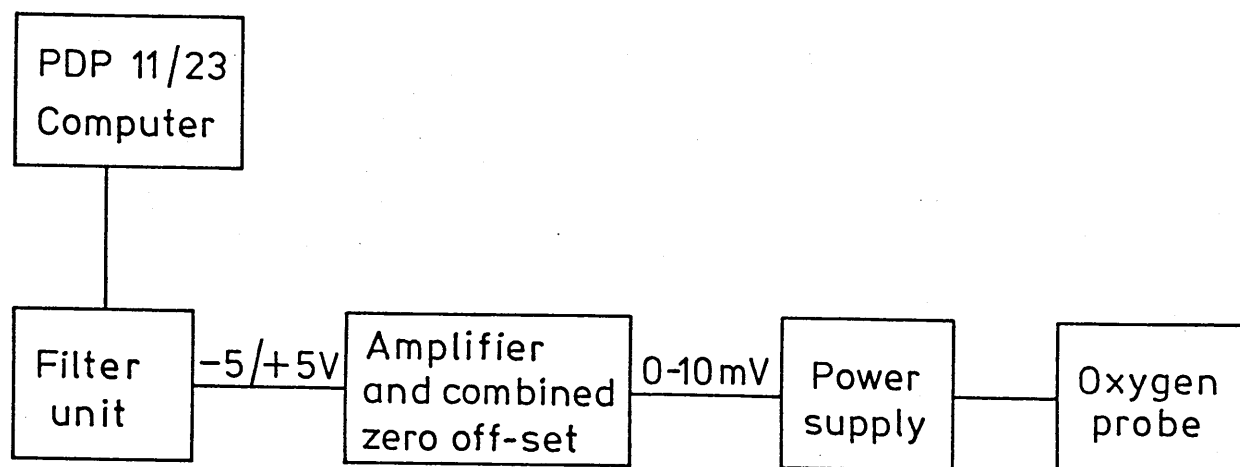
6.1

Co-ordinate system for bubble simulation



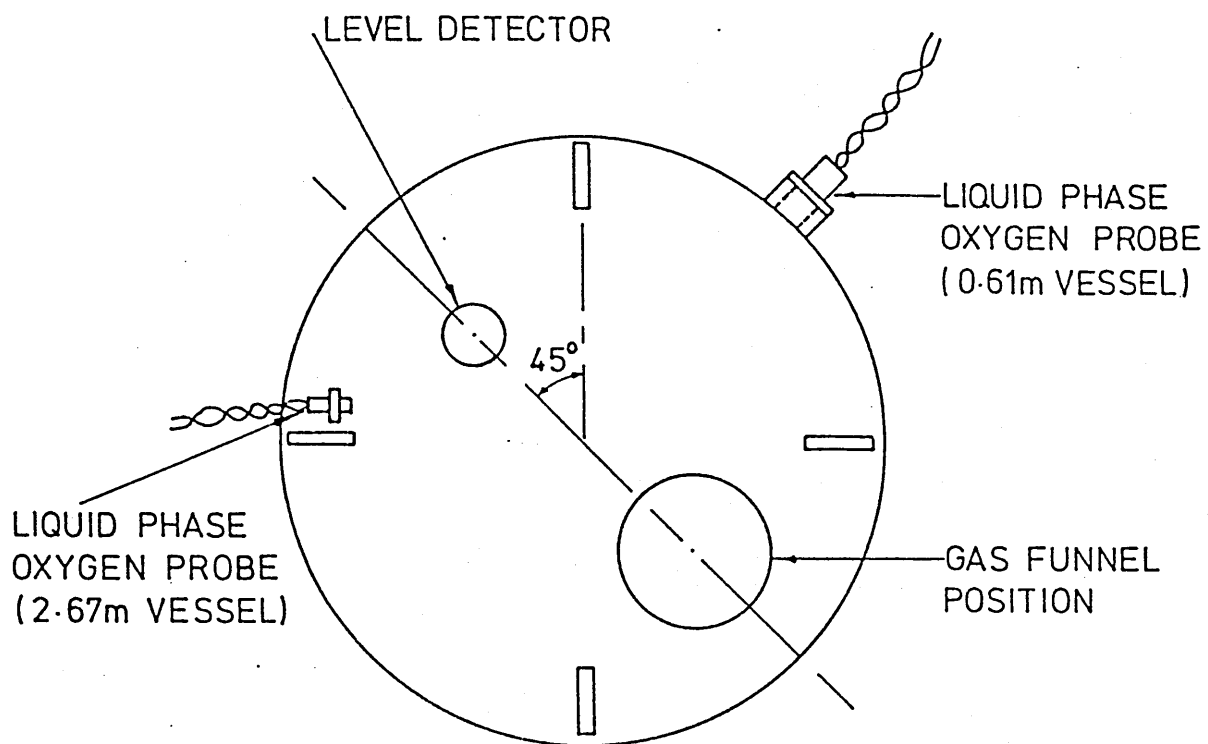


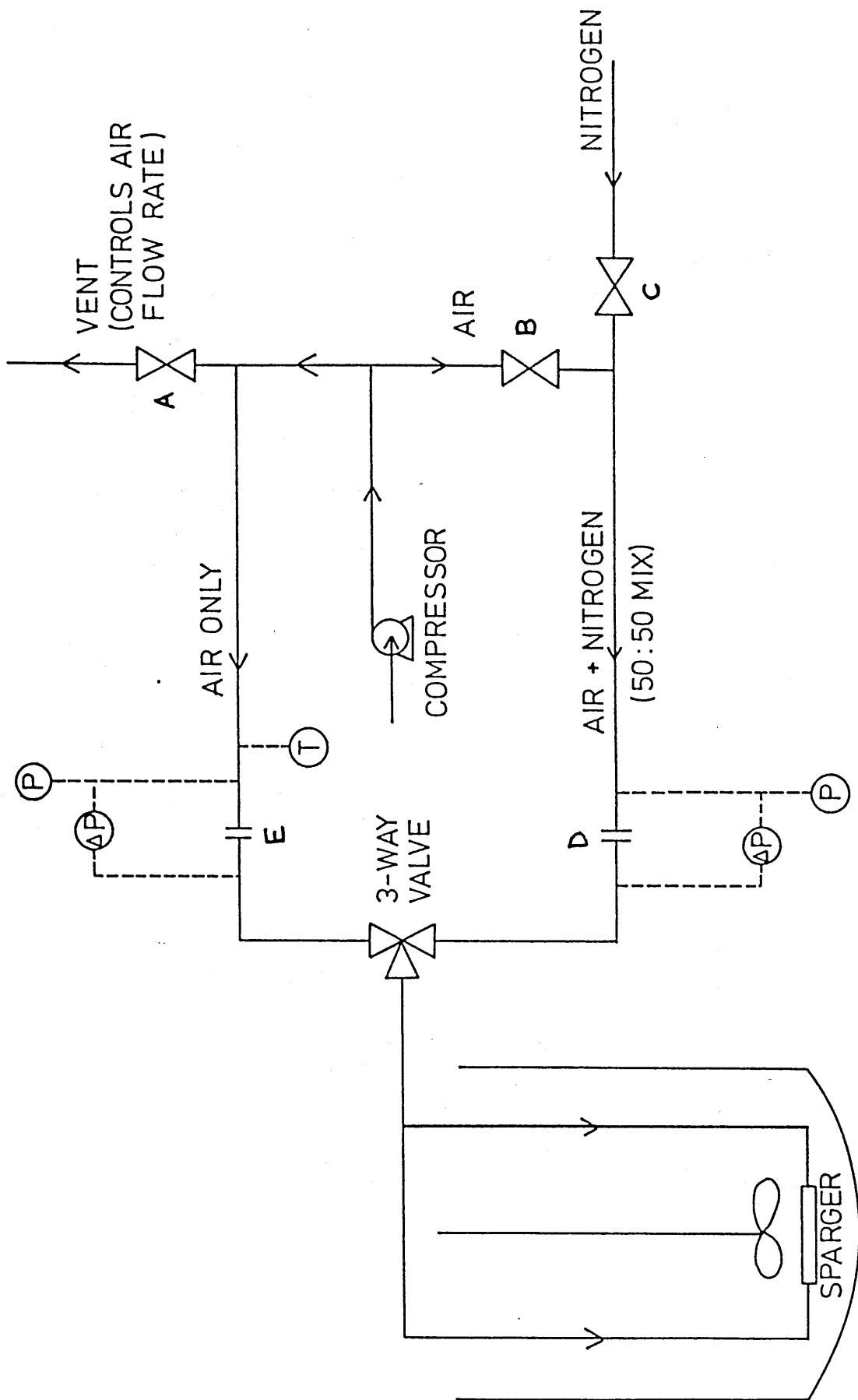
6.3 Gas supply lay-out: 0.61 m tank



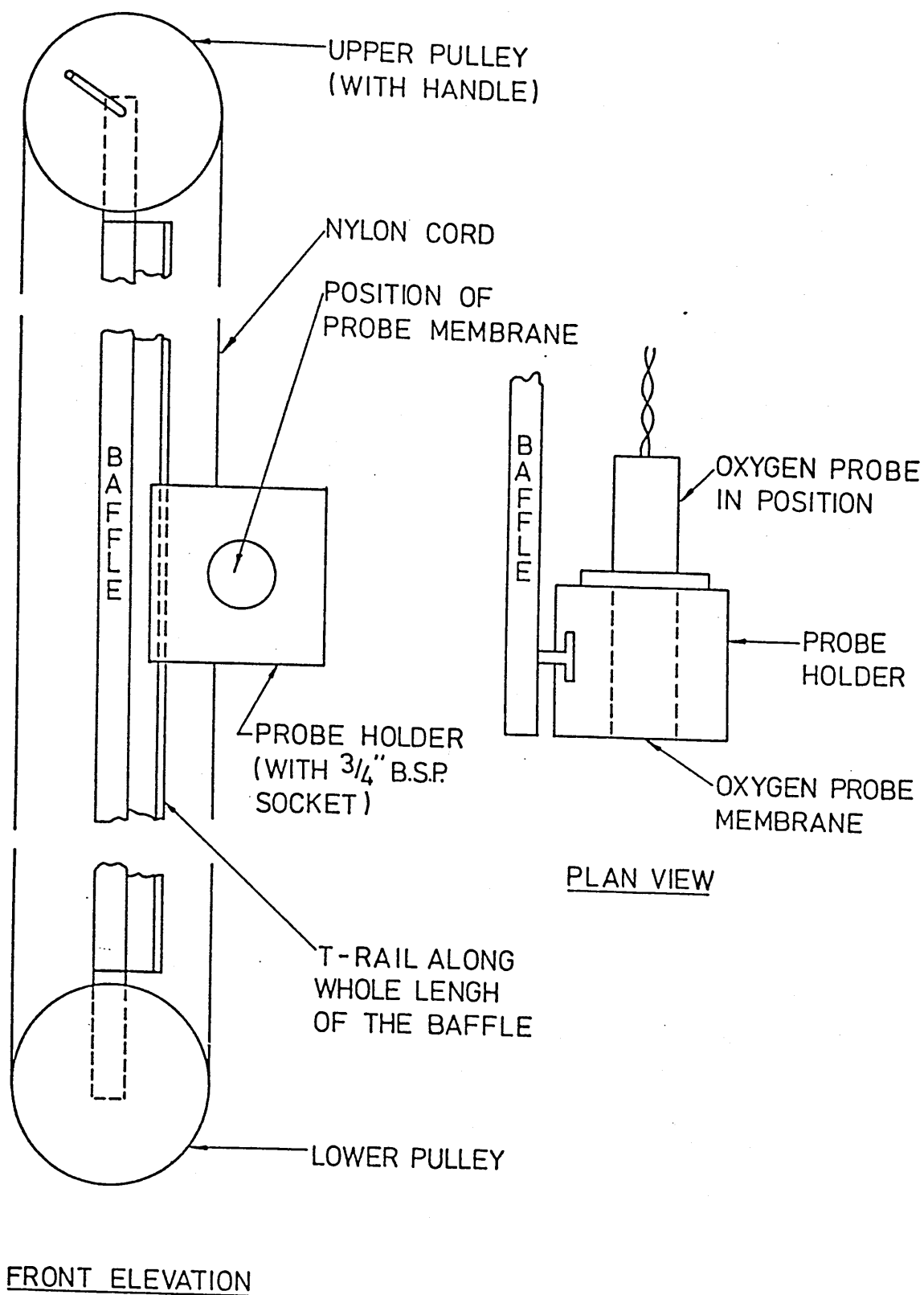
6.4

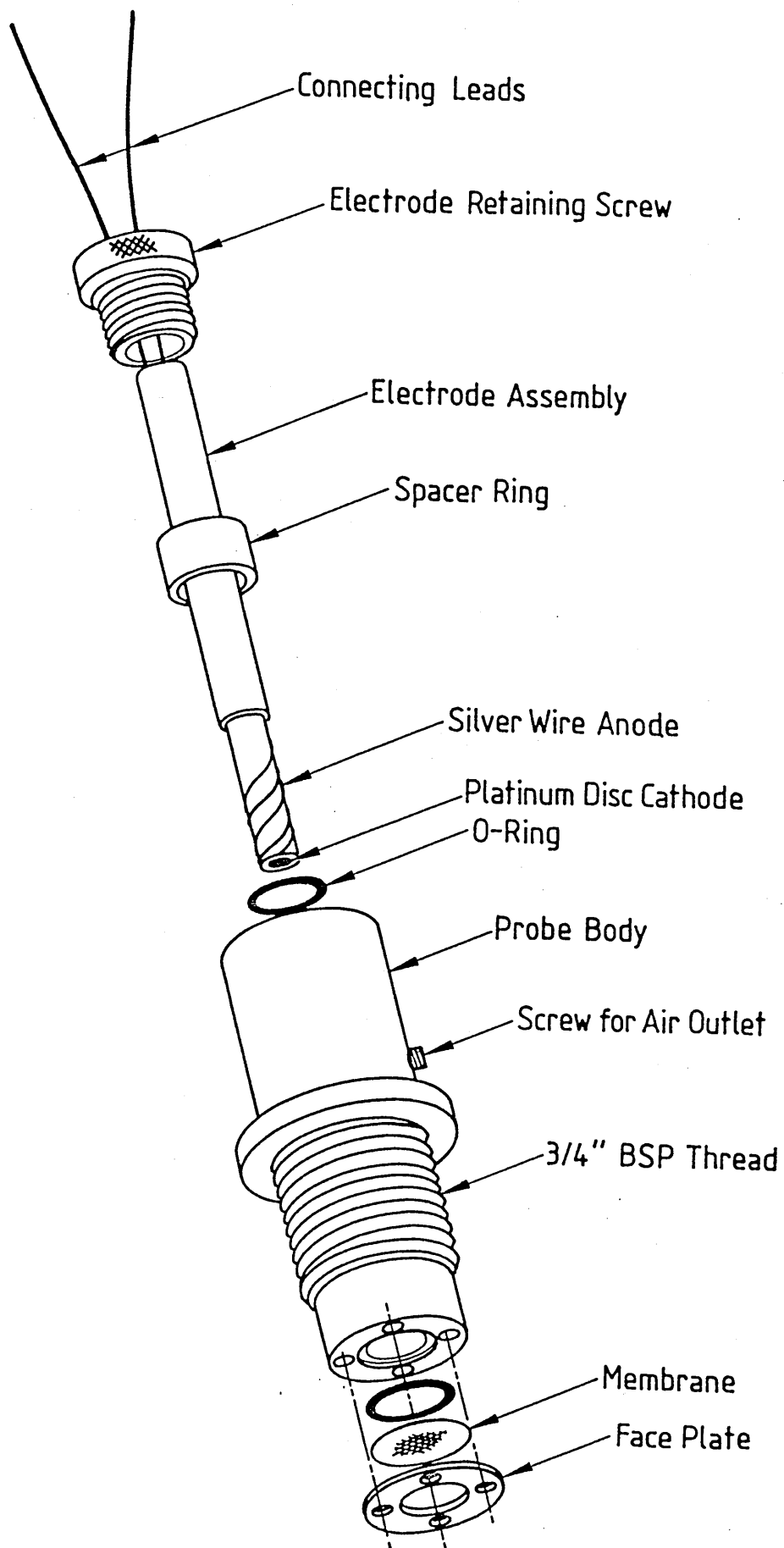
Instrument chain for oxygen concentration measurement

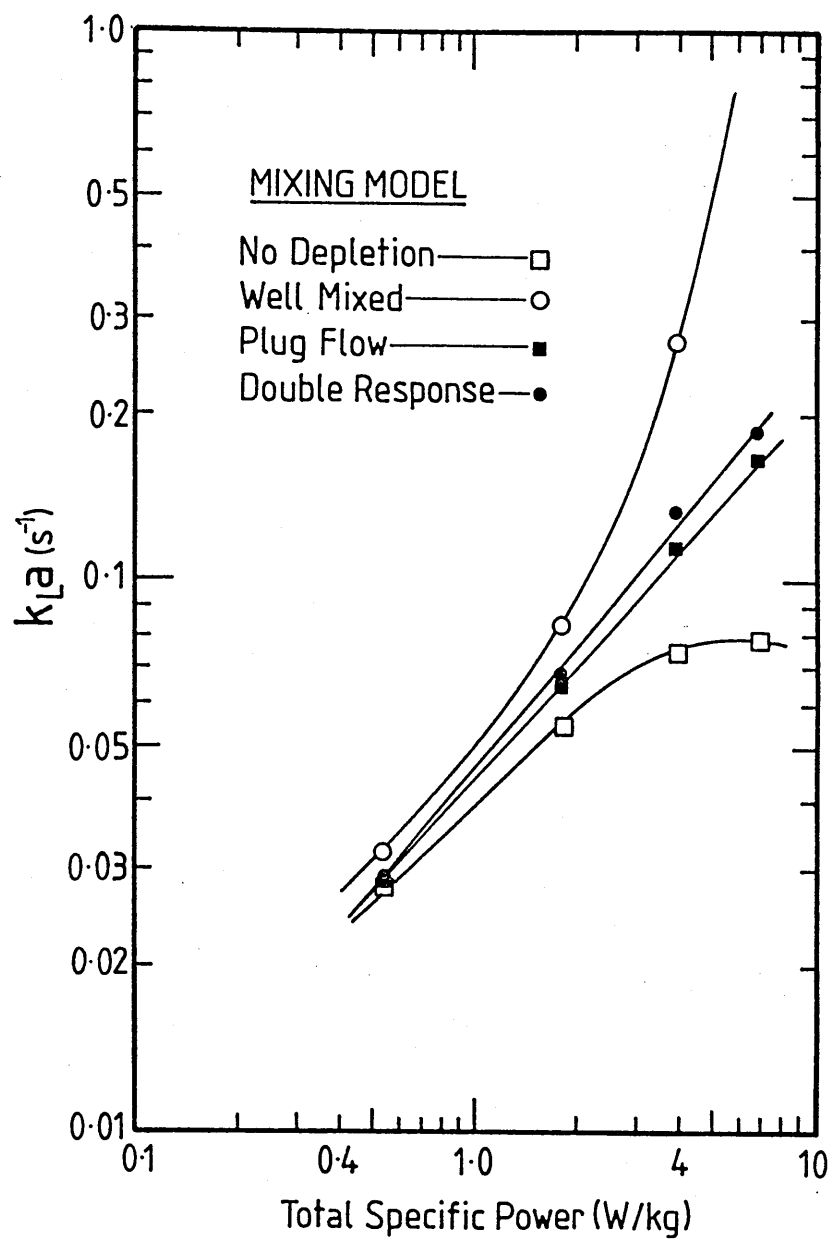


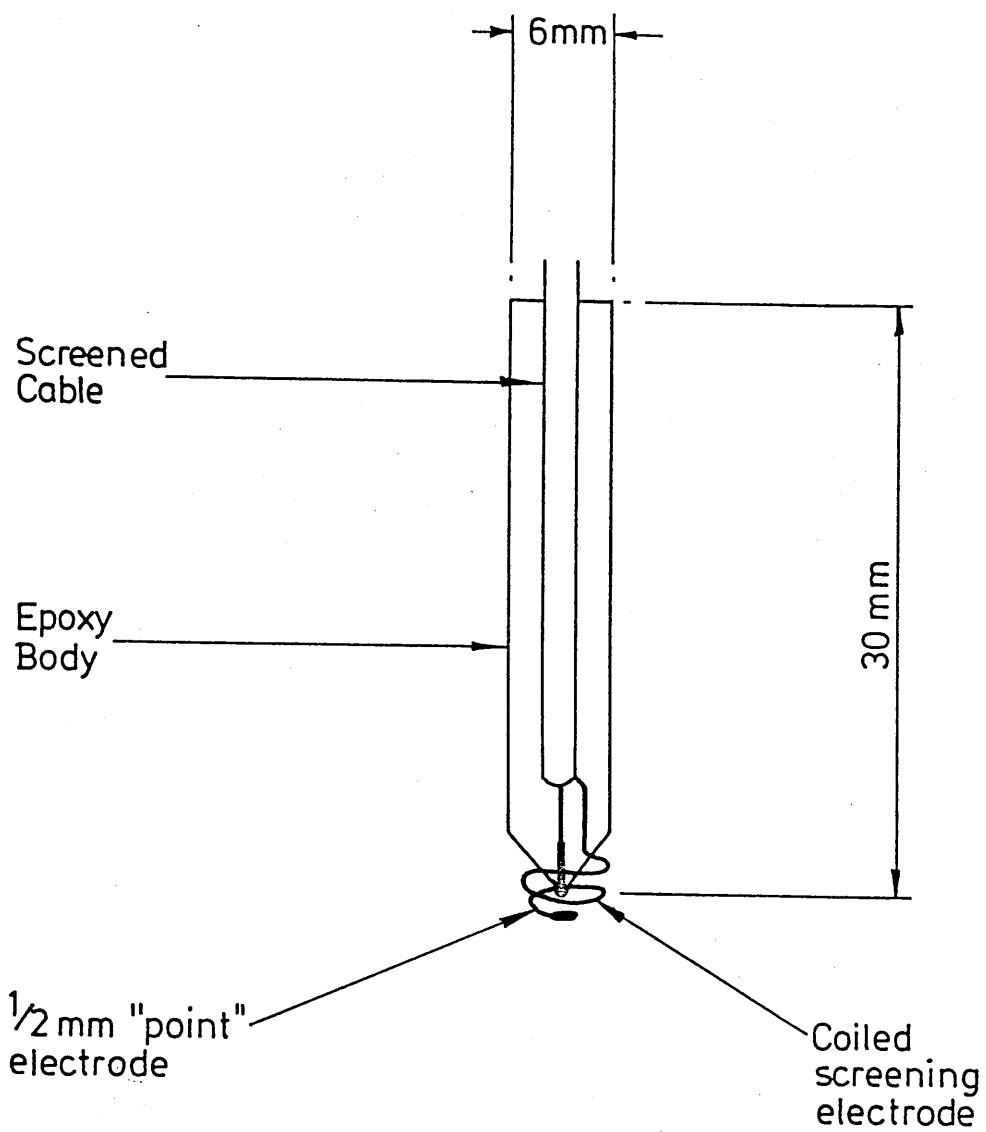


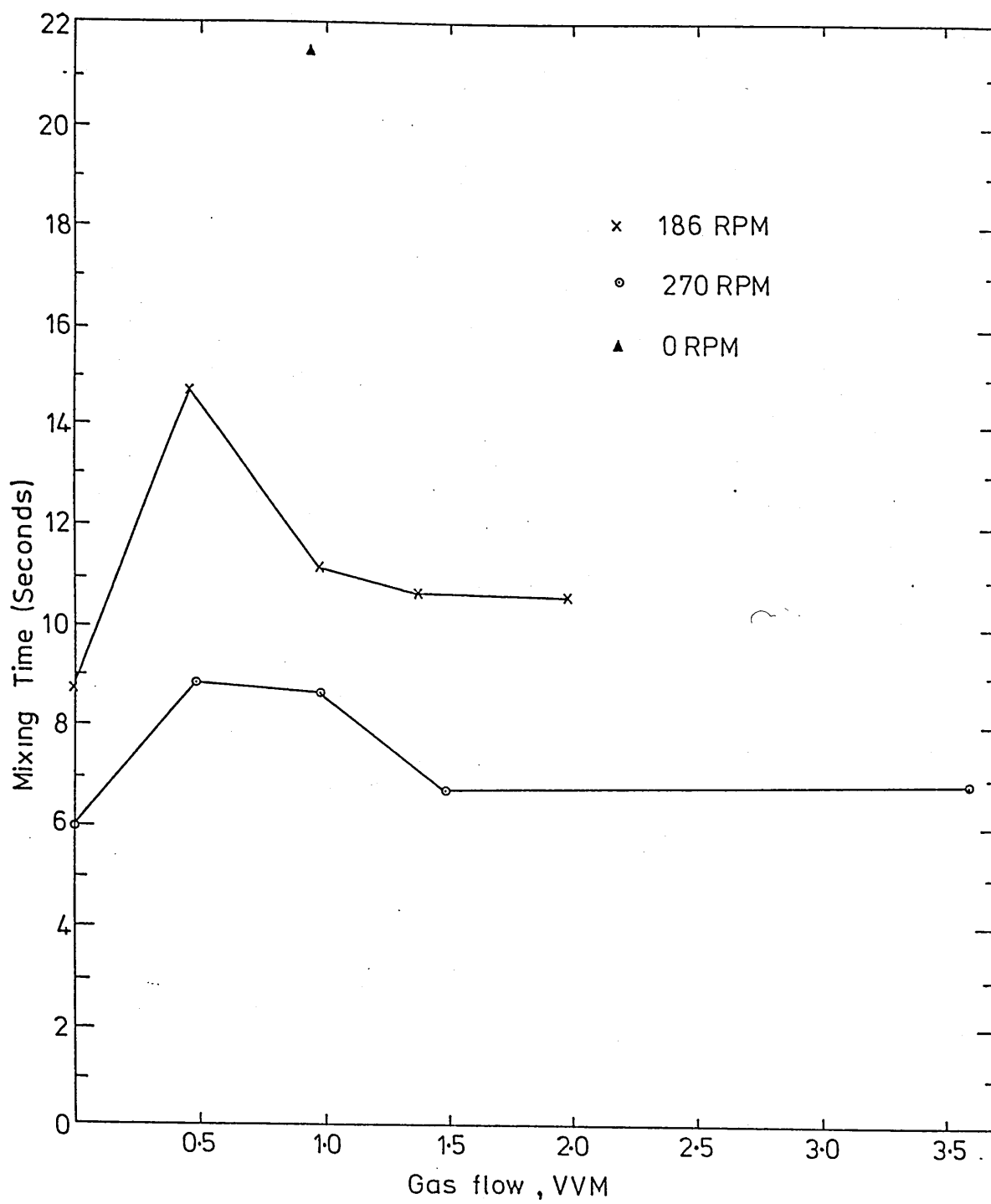
Gas supply lay-out: 2.67 m tank





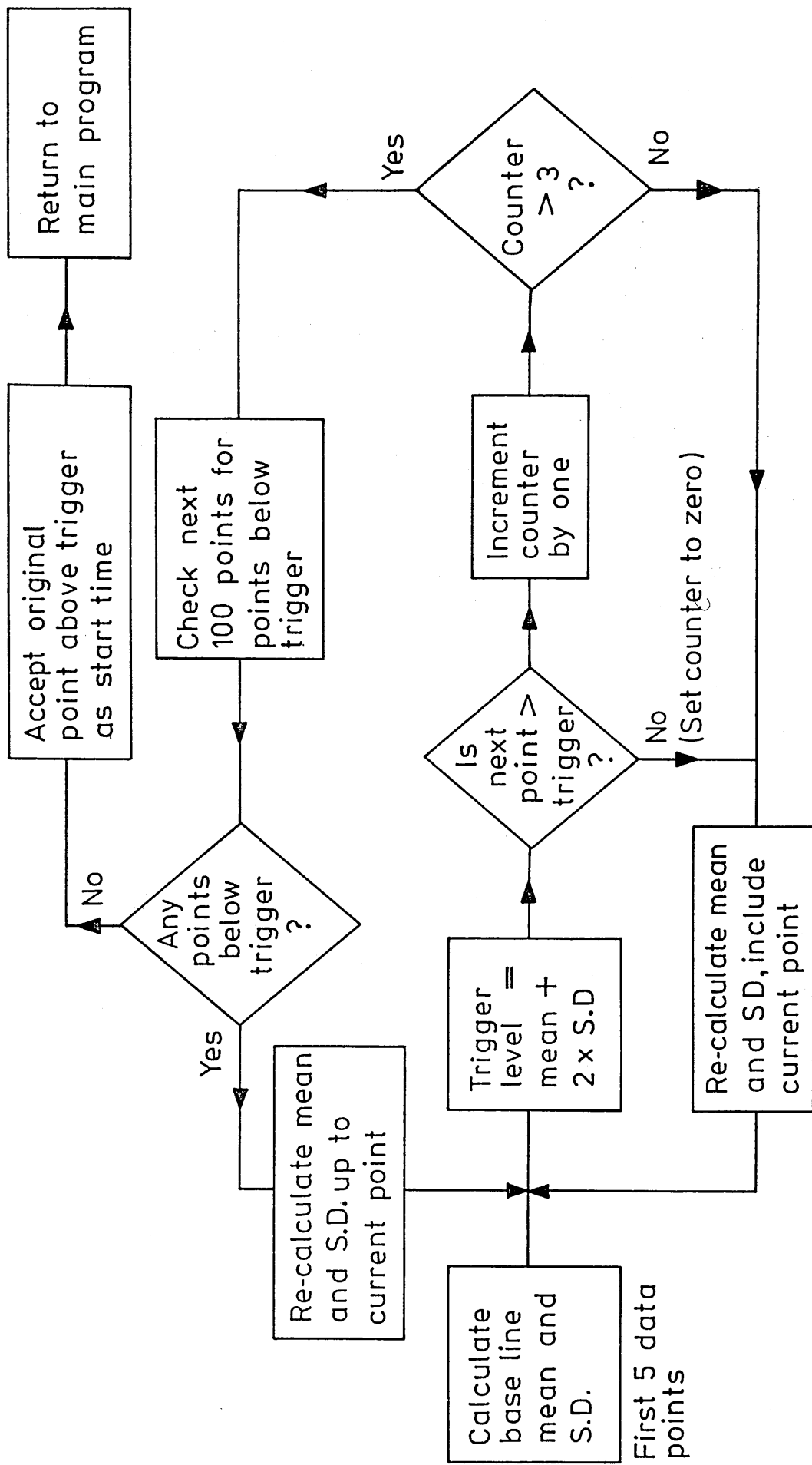


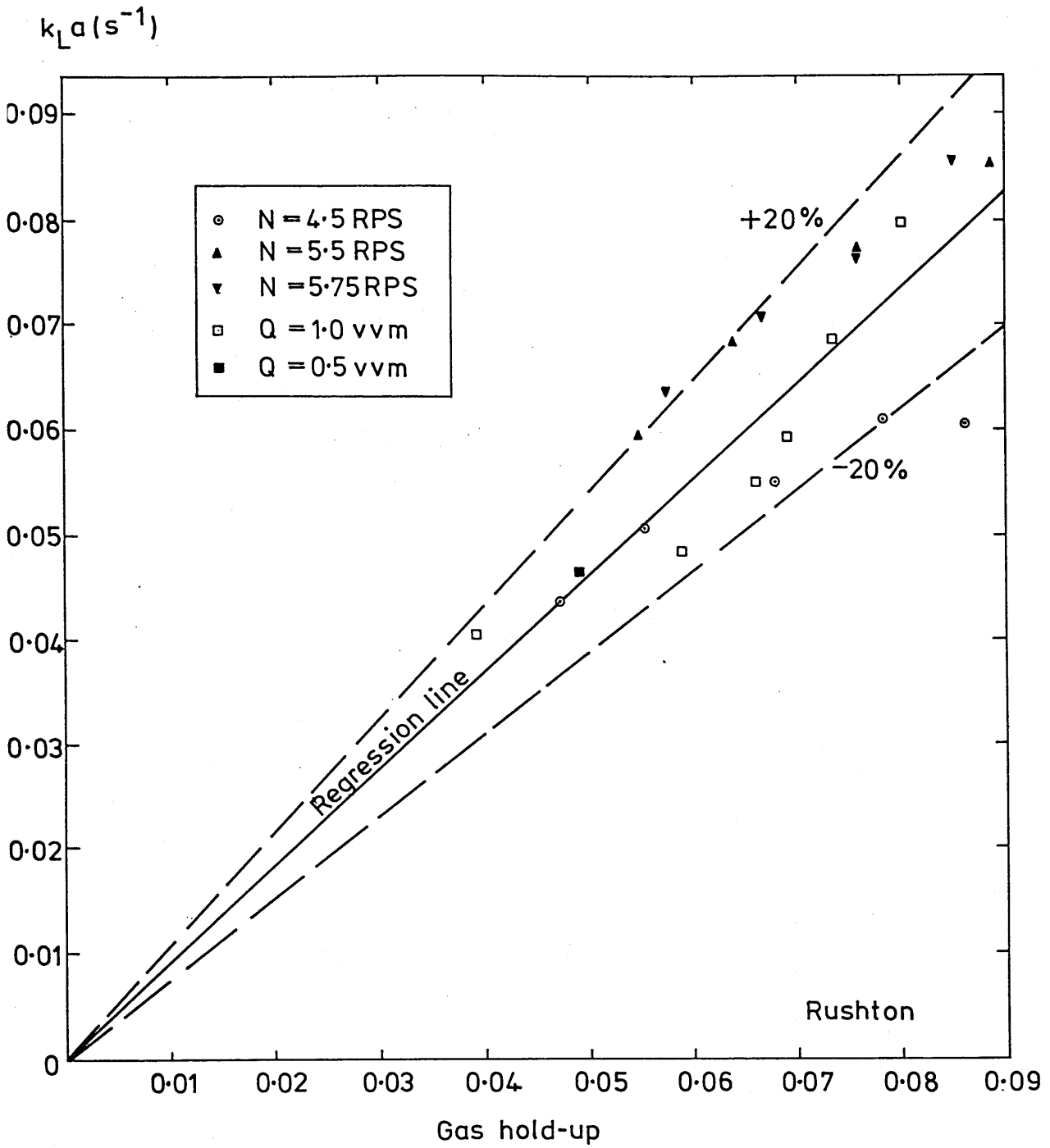


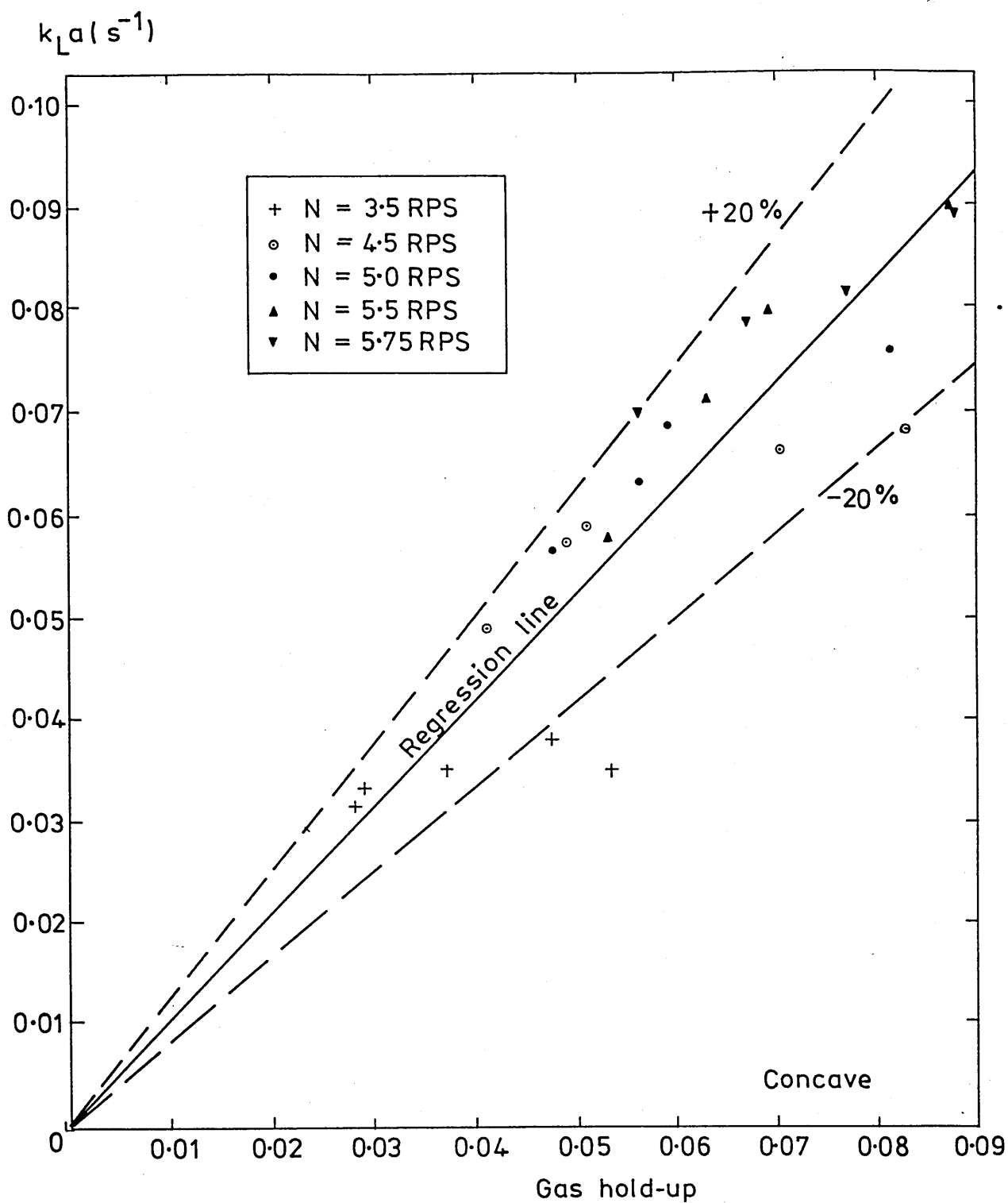


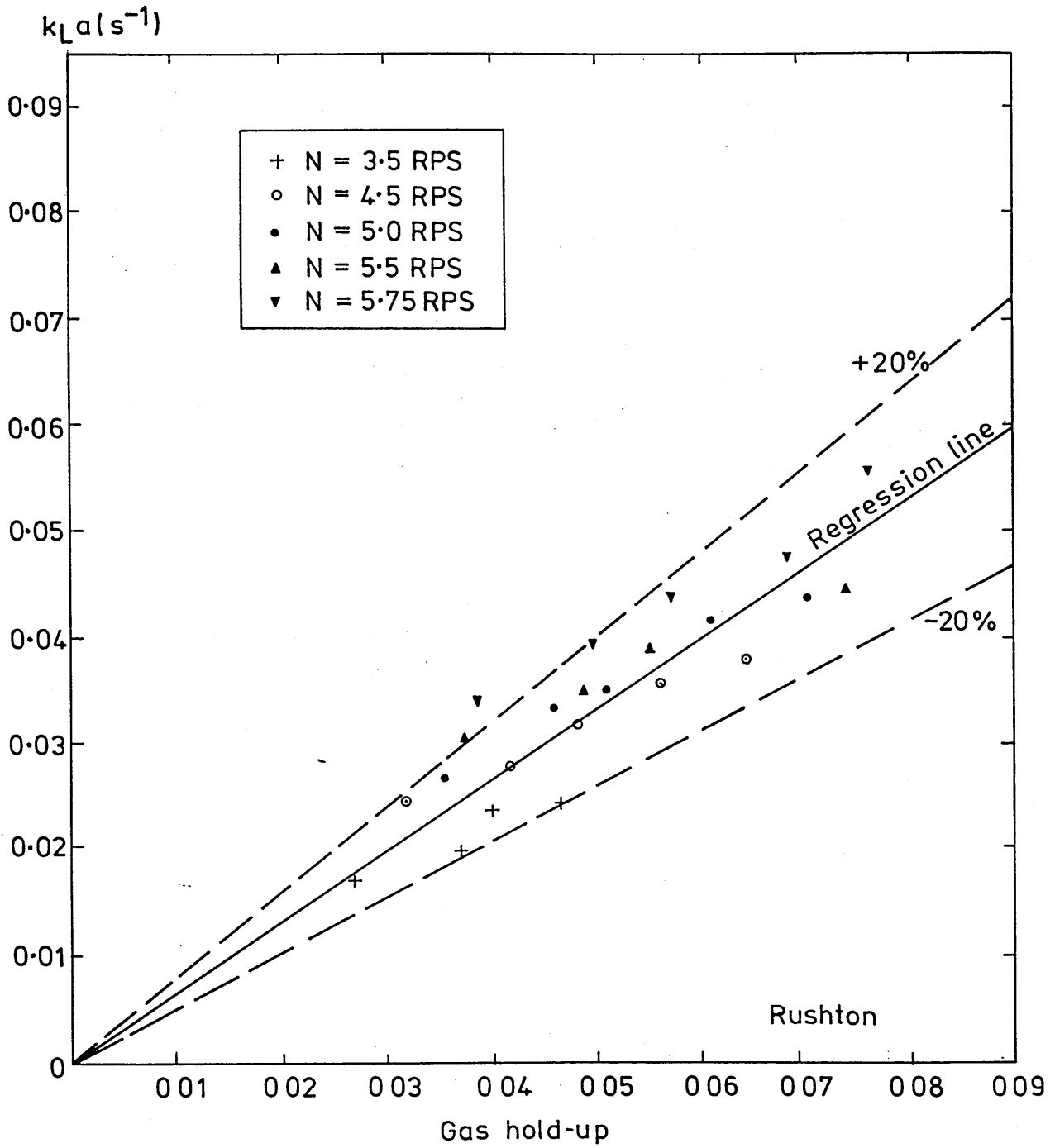
6.11

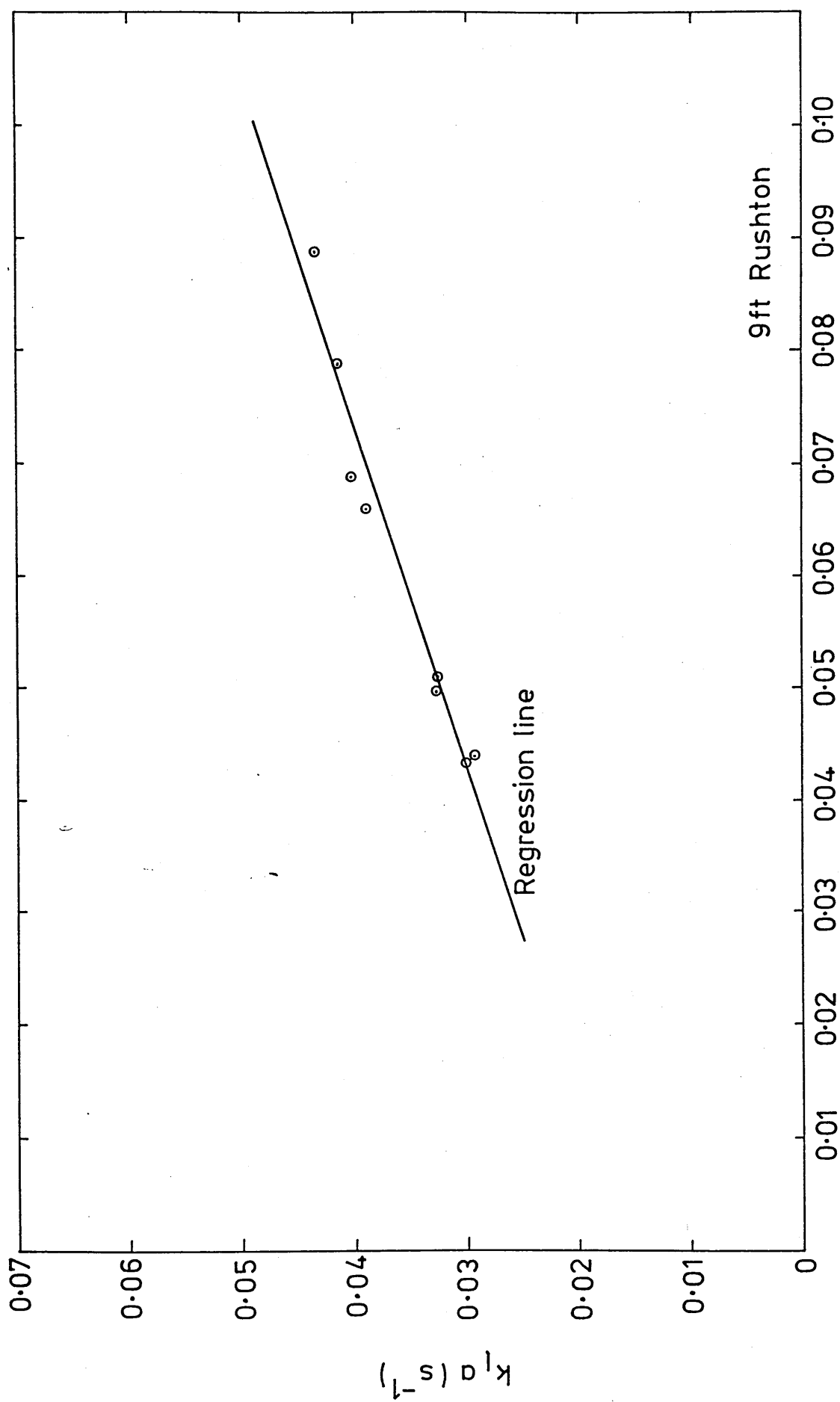
Mixing time as a function of gas flow rate







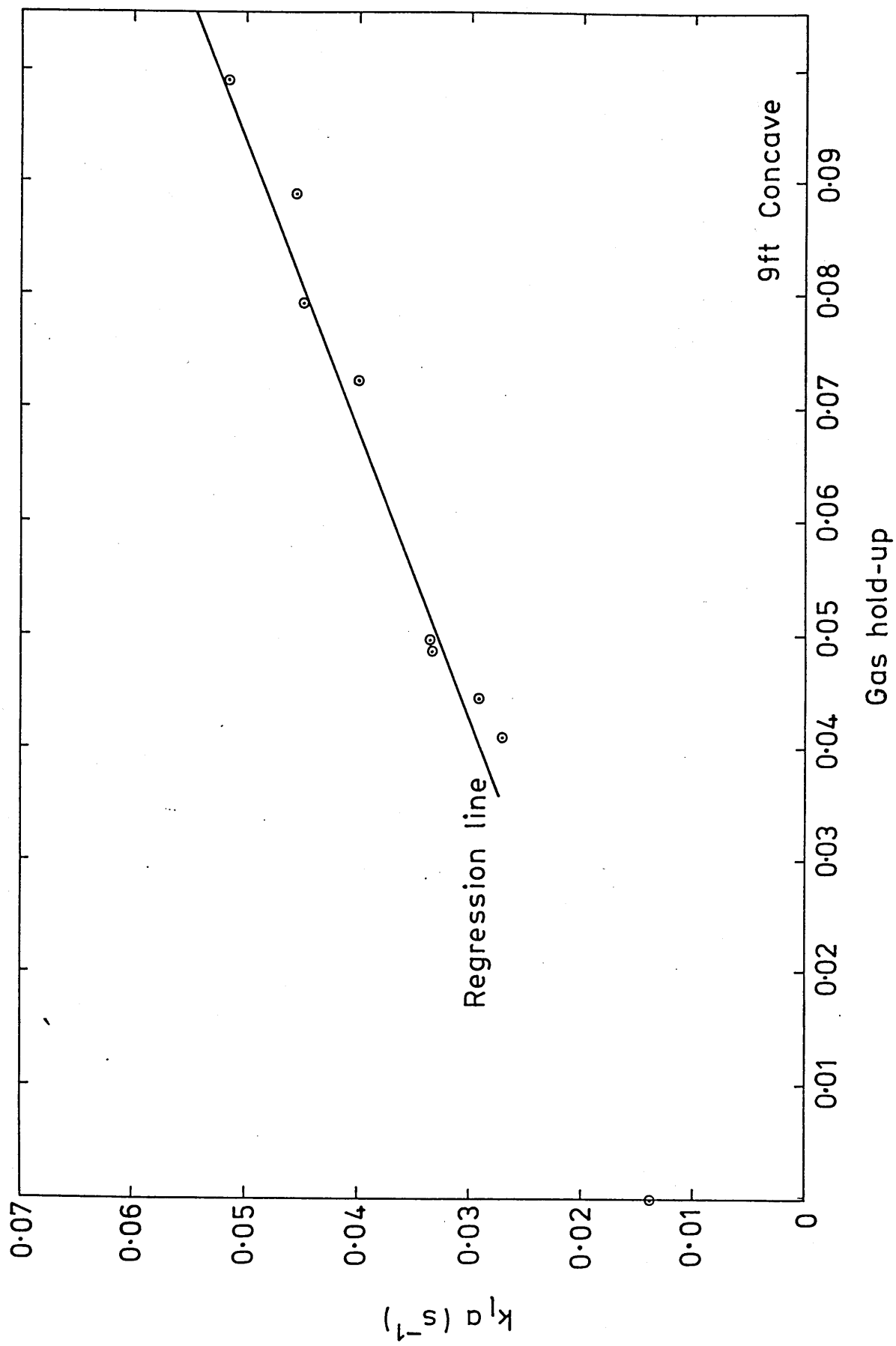




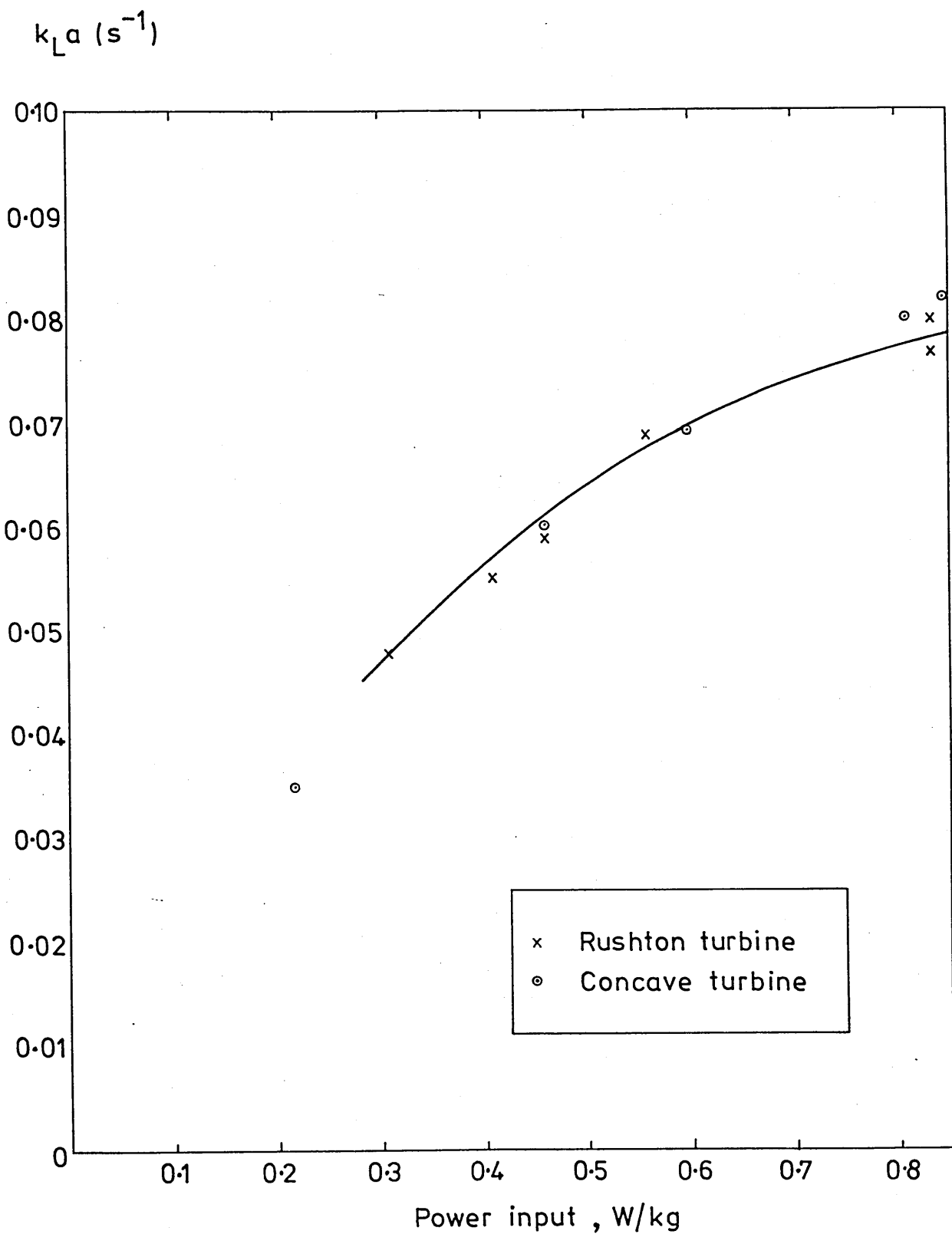
Gas hold-up

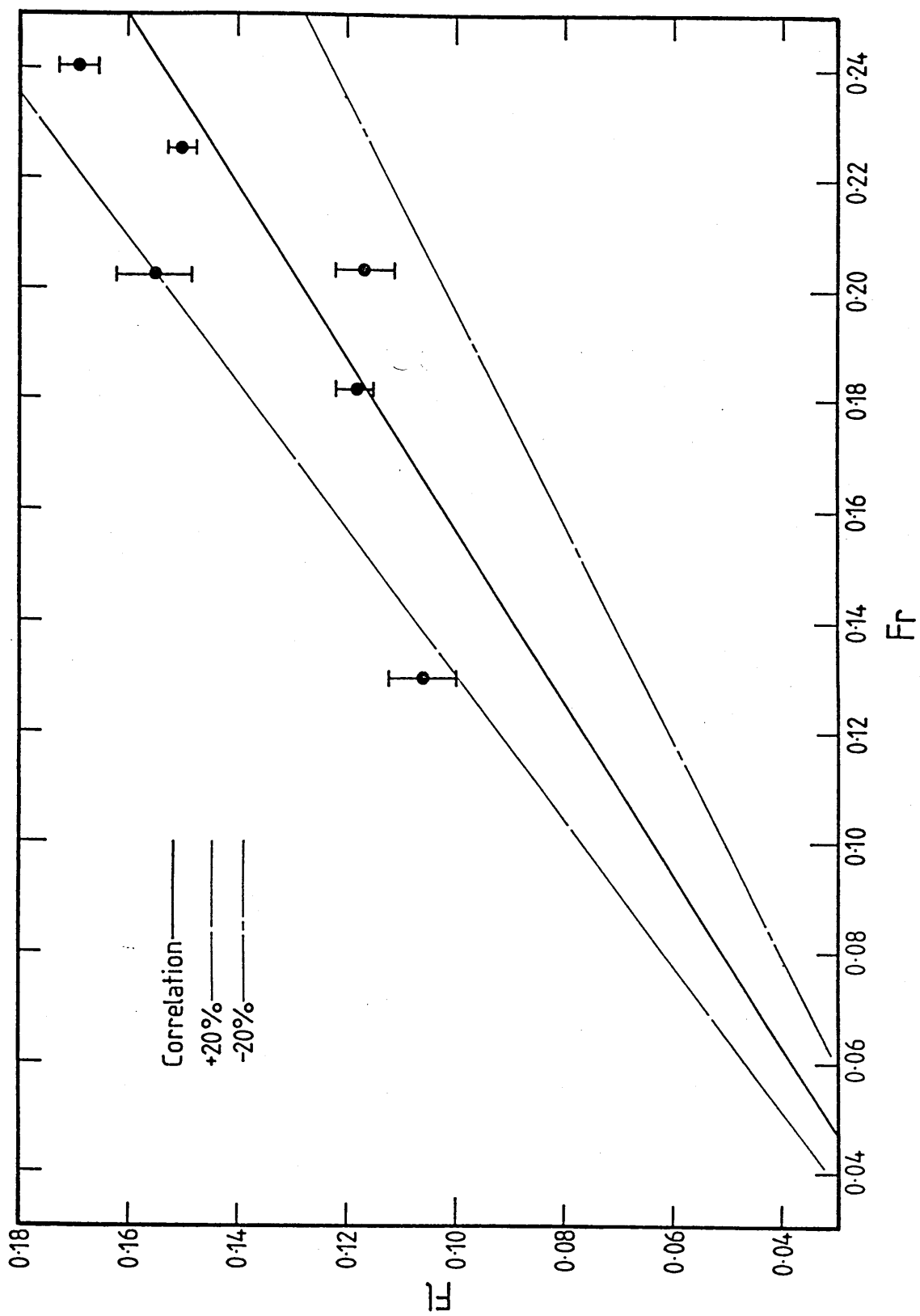
k_{La} as a function of hold-up: 2.67 m tank: Rushton: tap water

6.16

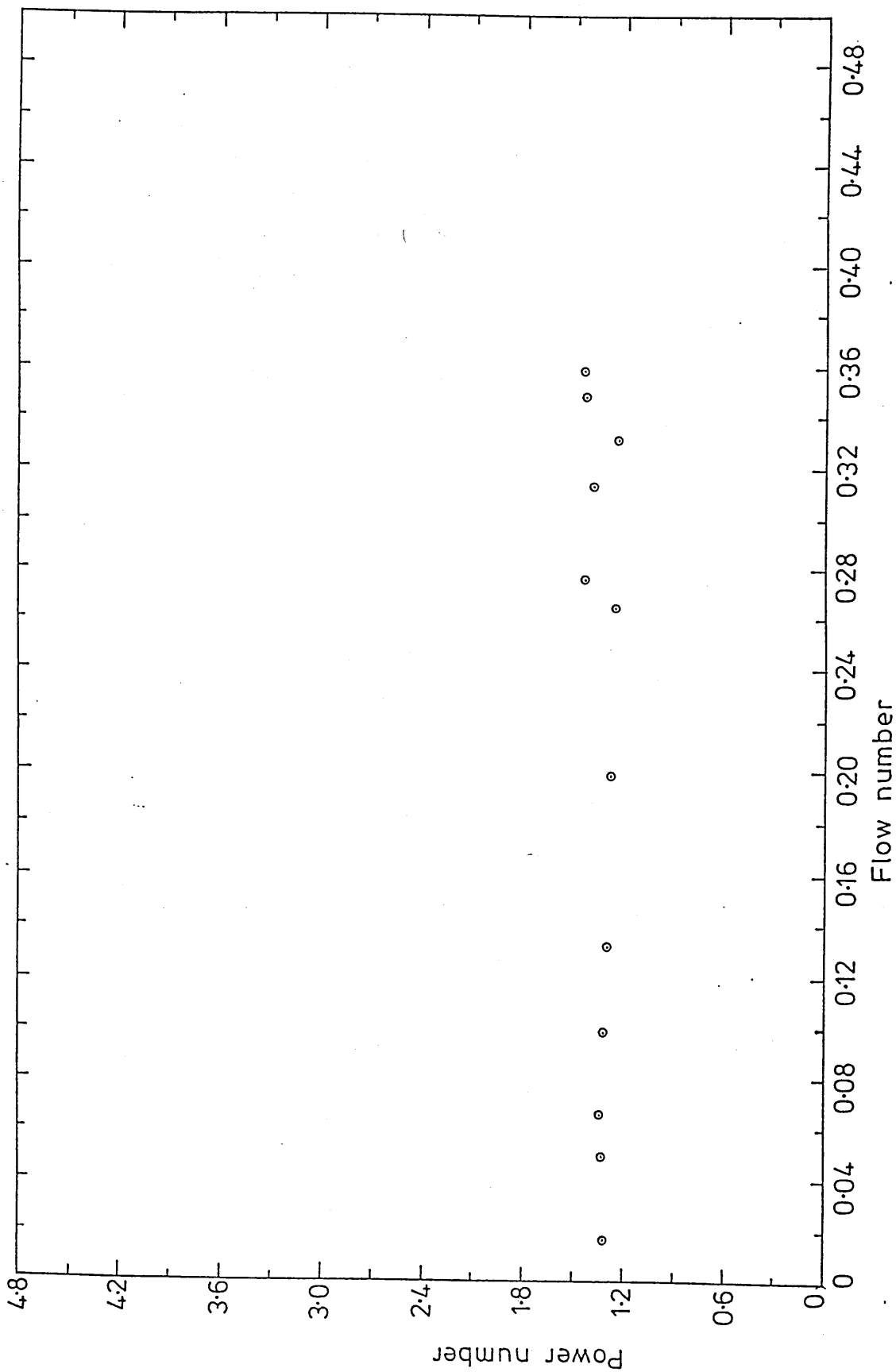


6.17 k_{La} as a function of hold-up: 2.67 m tank: concave: tap water

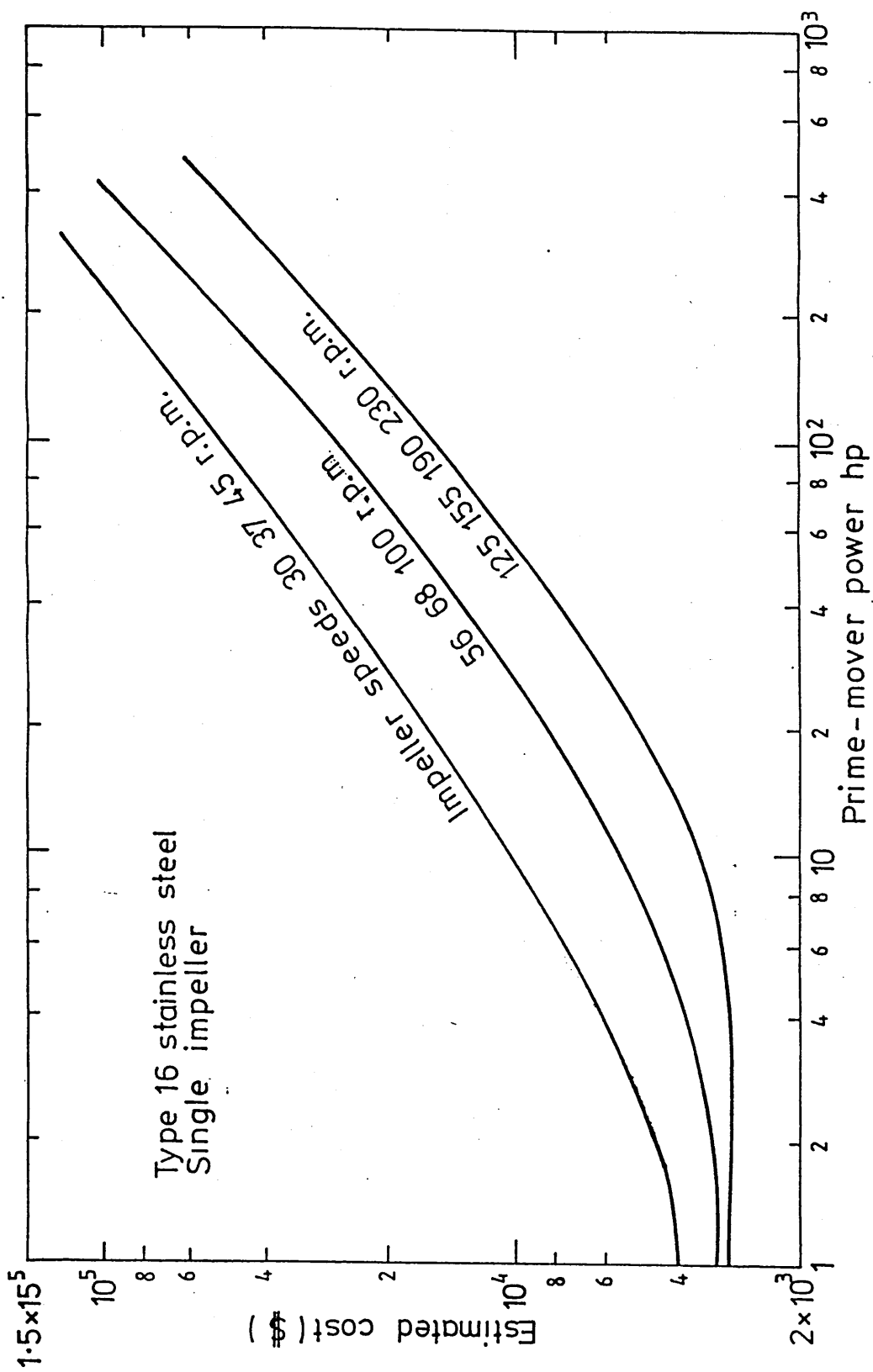




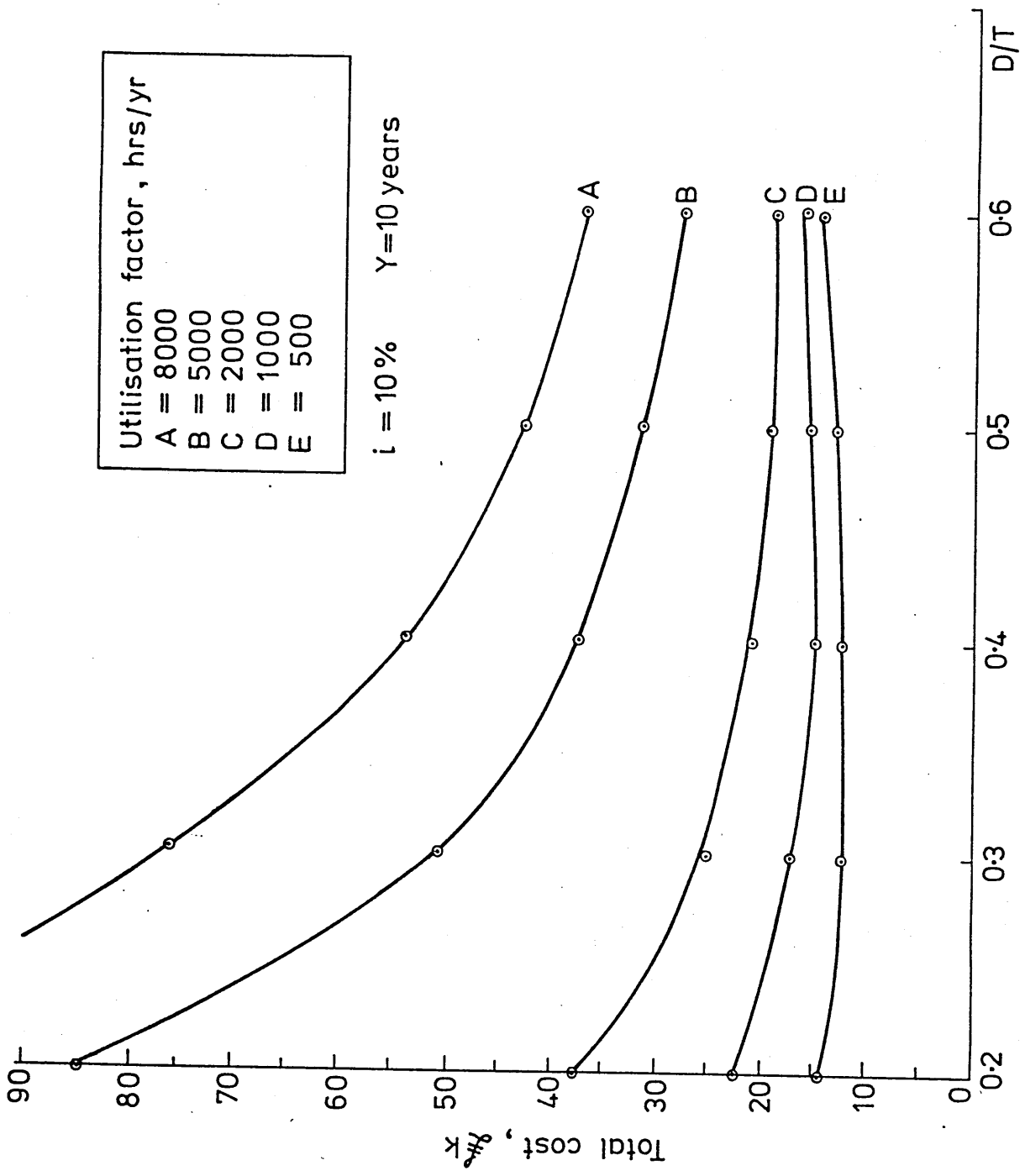
7.1 Flooding data



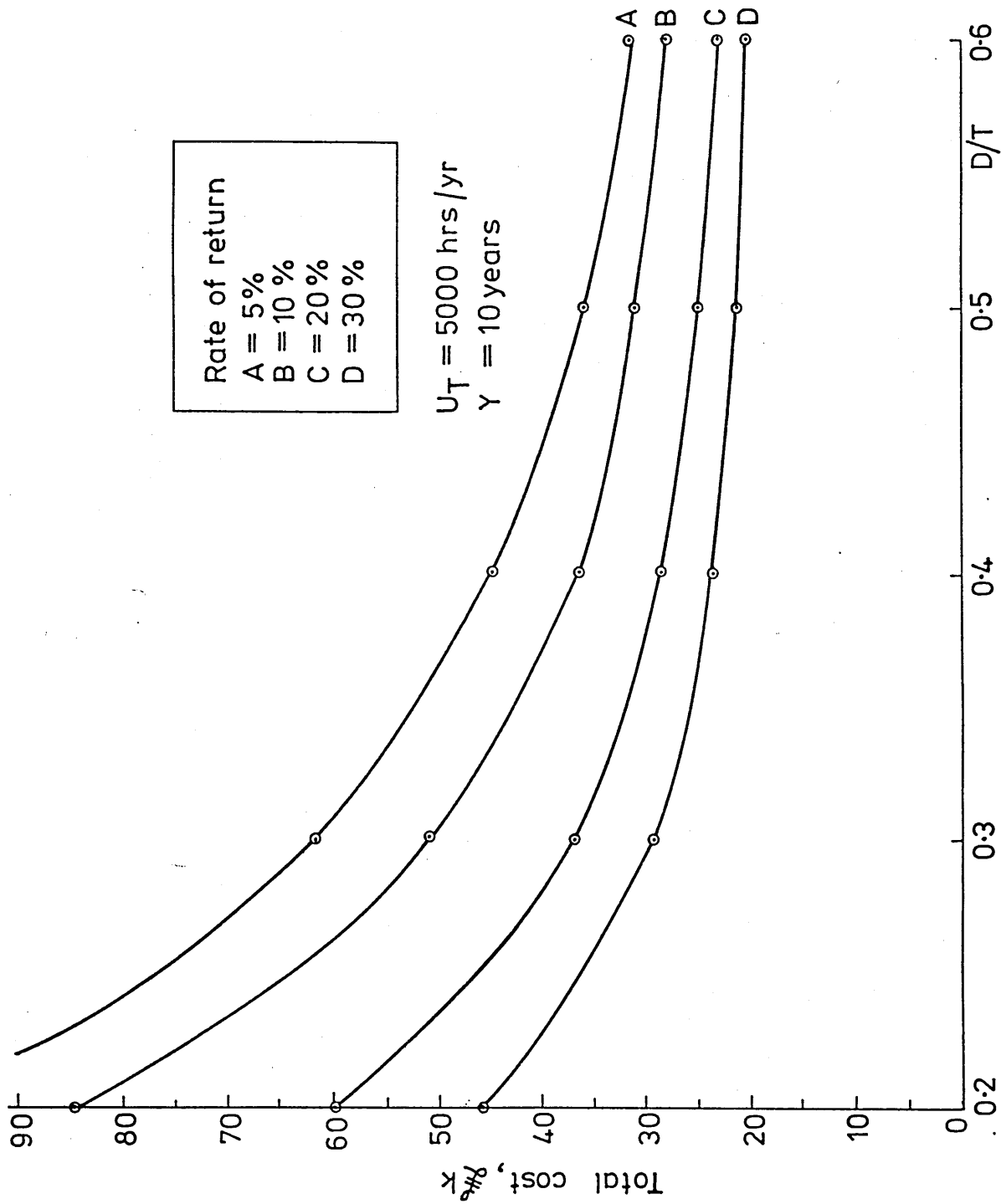
7.2 Gassed power number against flow number: CMC solution



8.1 Estimated agitator capital cost based on 1976 dollar prices

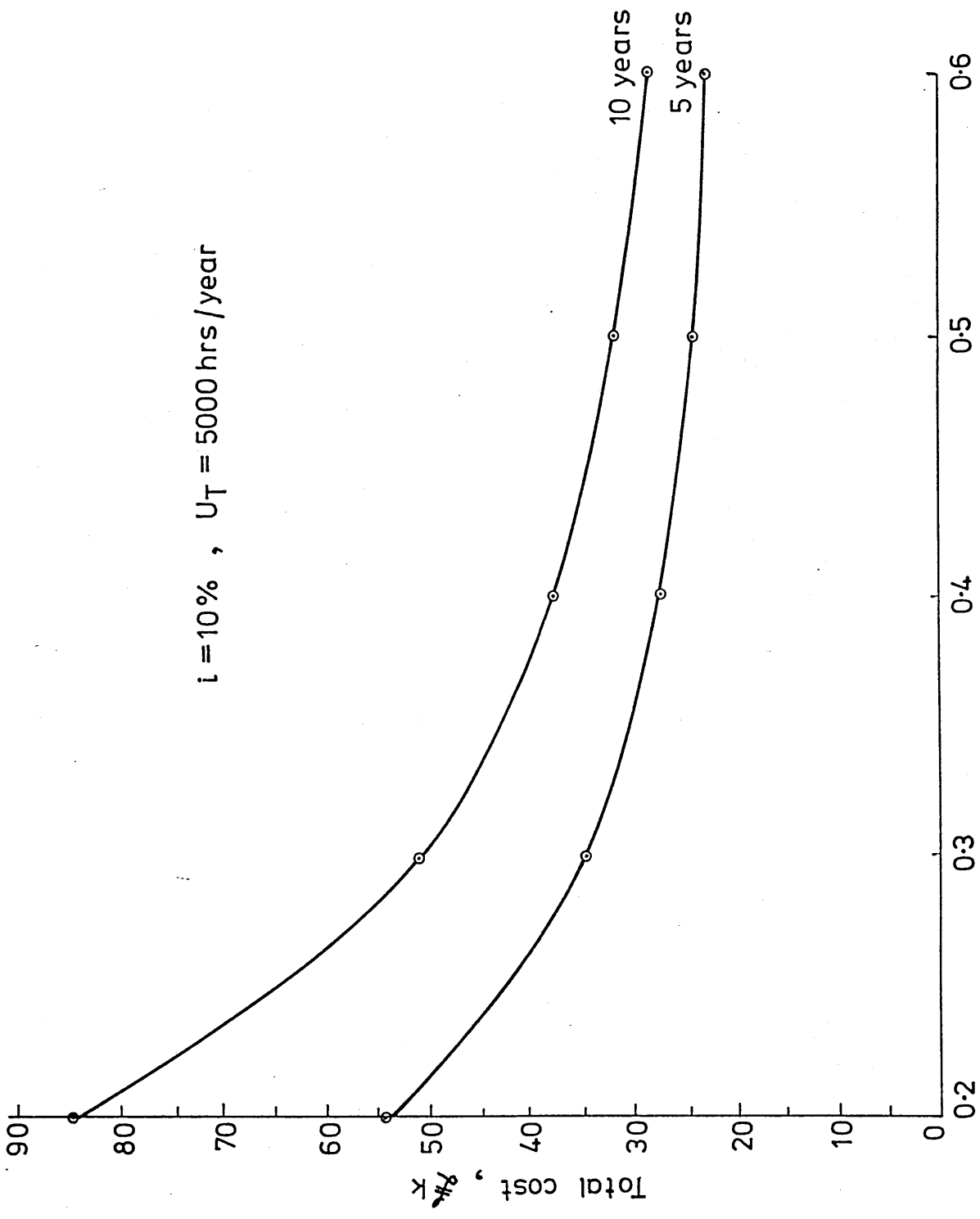


8.2 Influence of utilisation on total cost

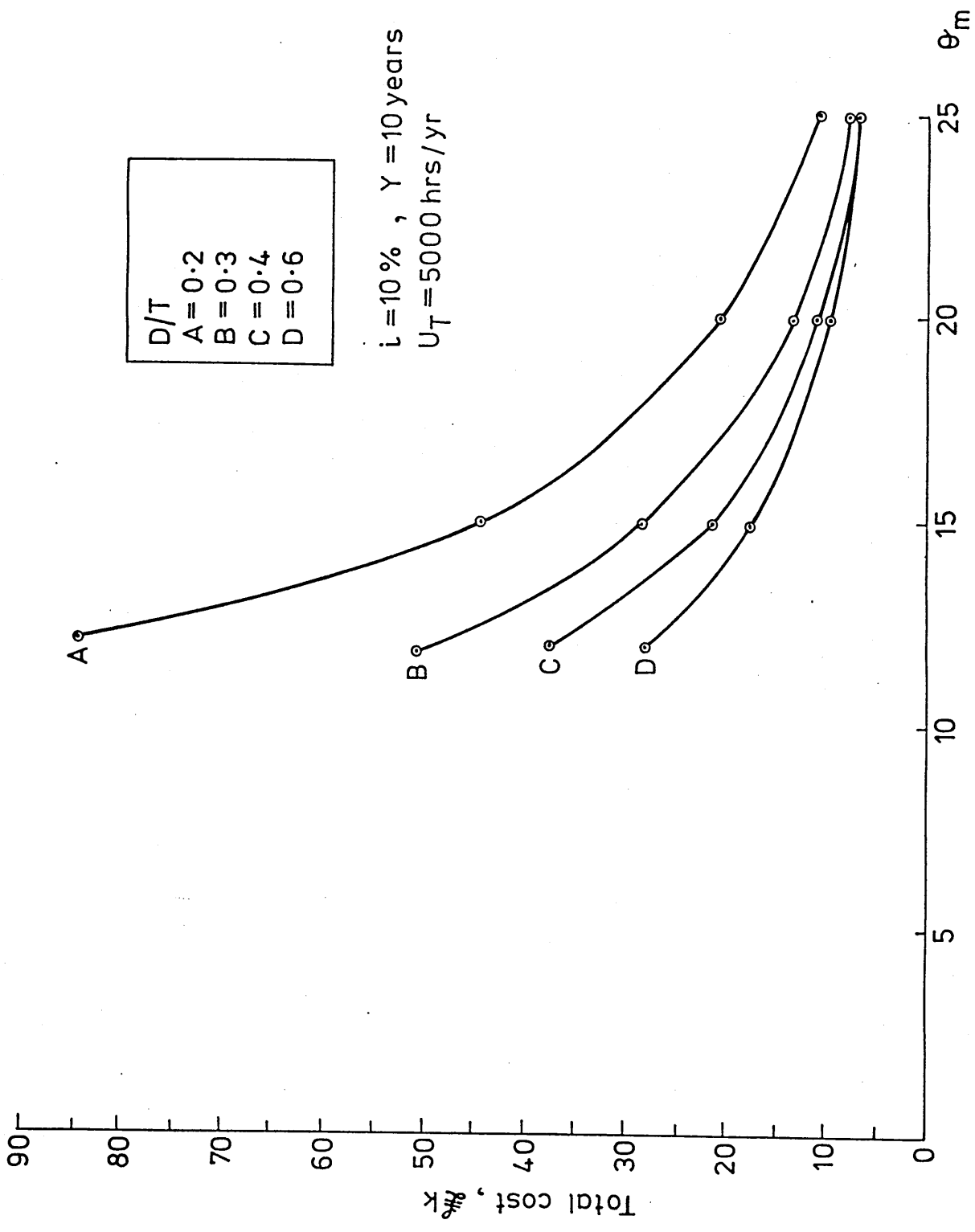


8.3 Influence of rate of return on total cost

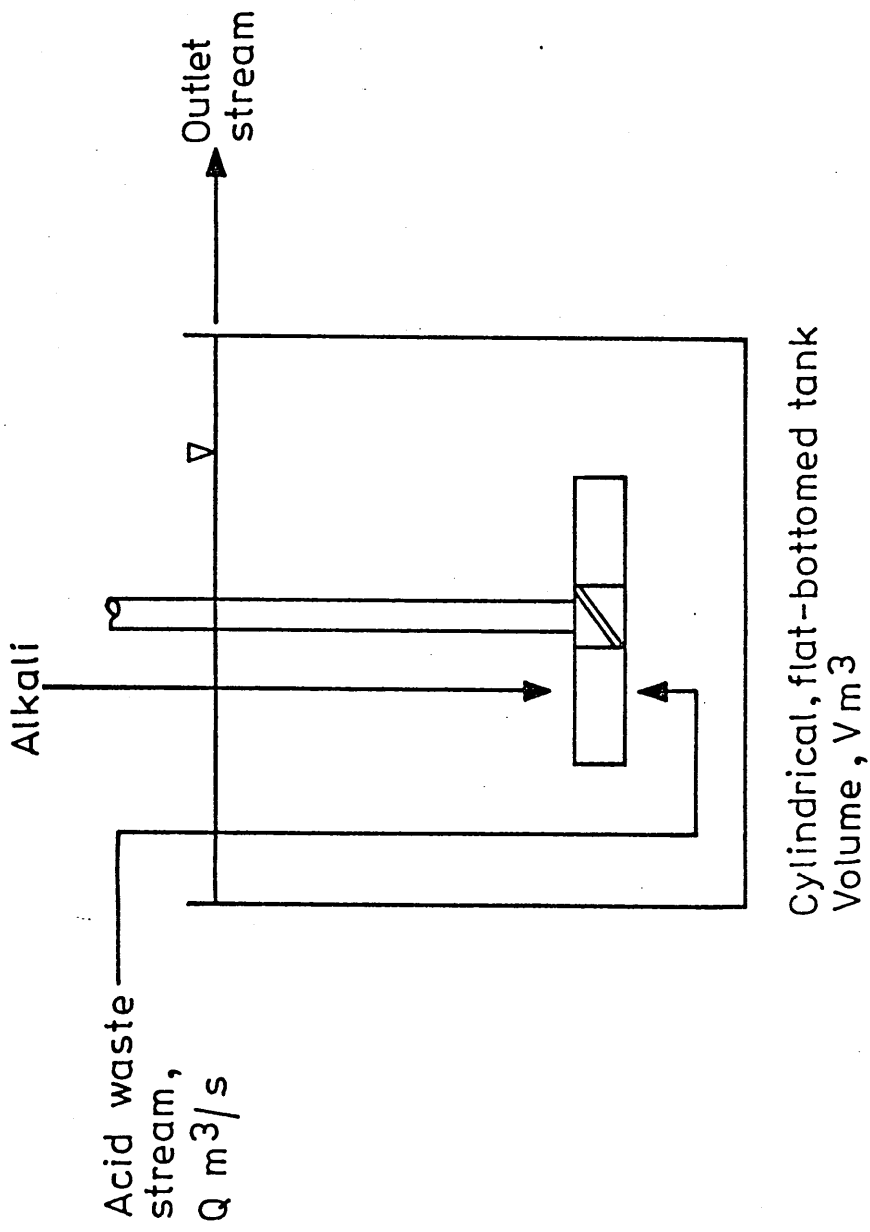
$i = 10\%$, $U_T = 5000 \text{ hrs / year}$



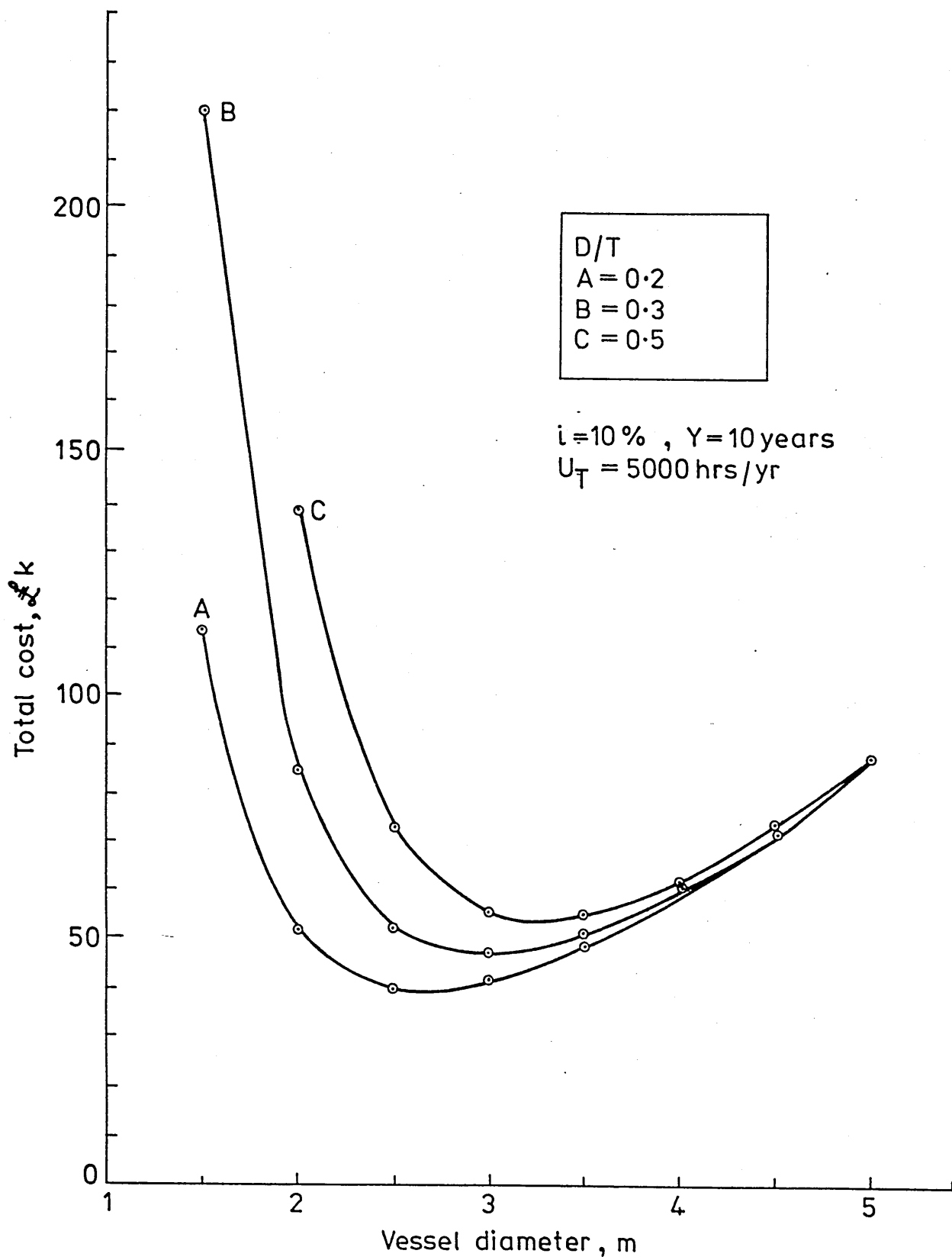
8.4 Influence of estimated life-time on total cost

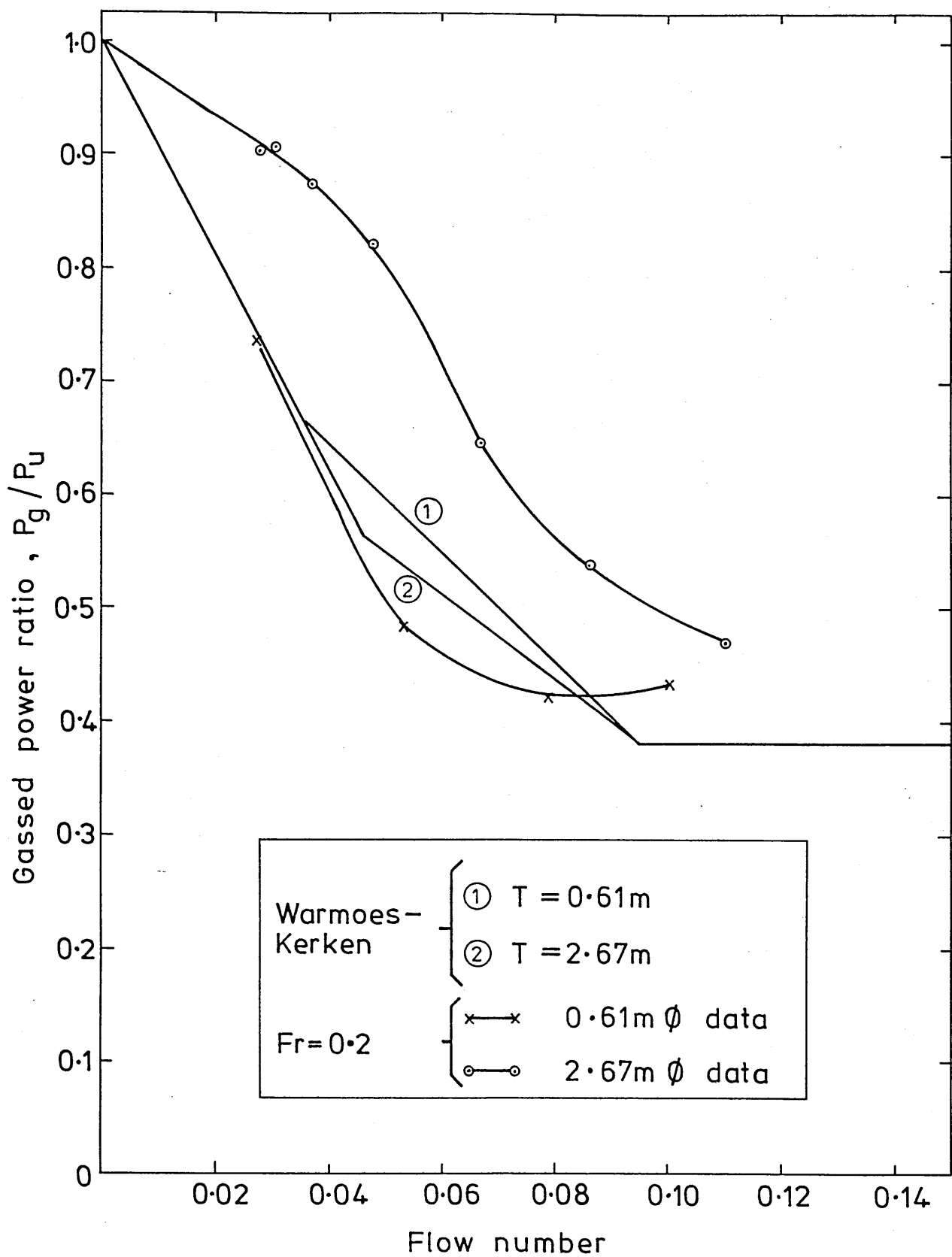


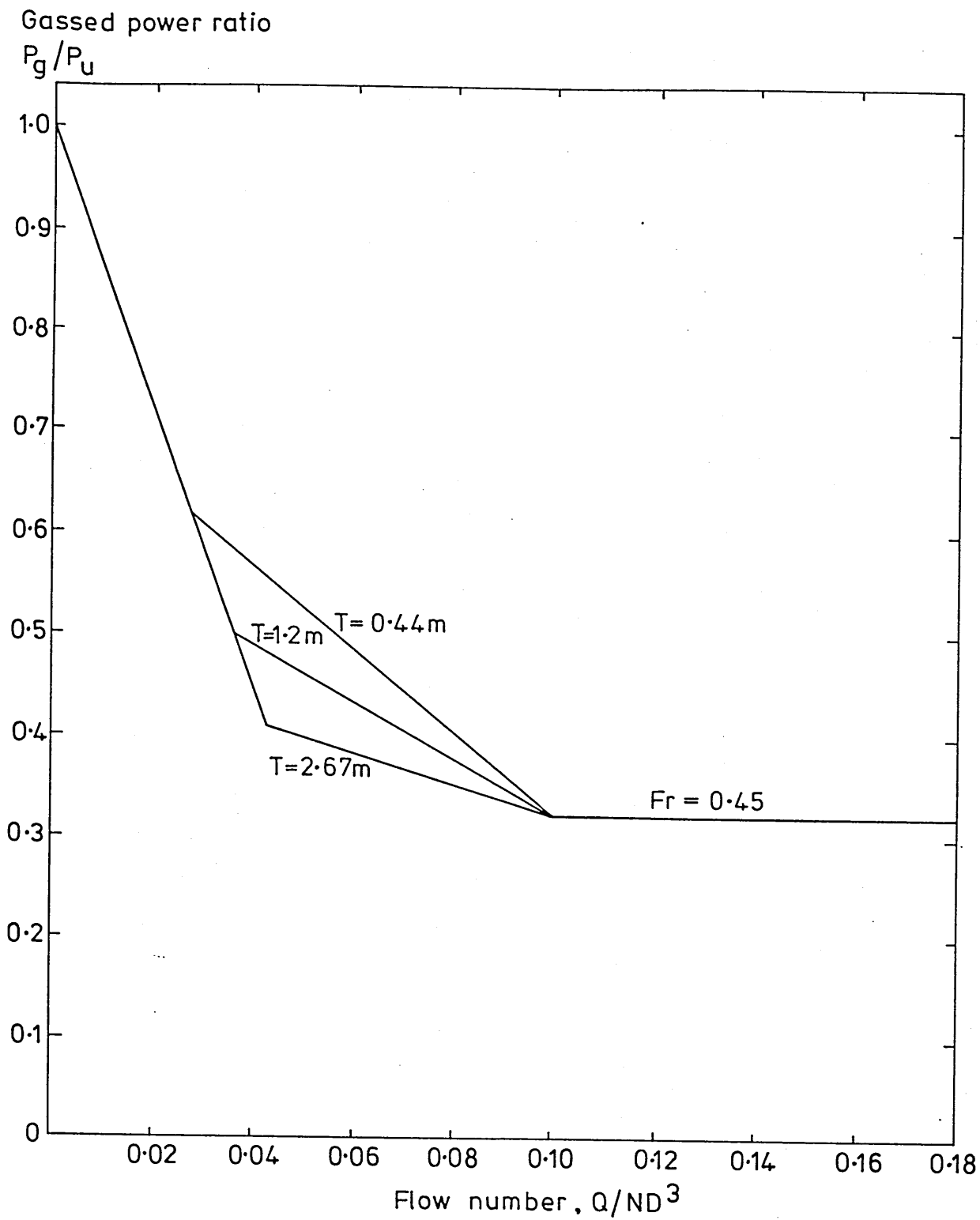
8.5 Influence of mixing time on total cost

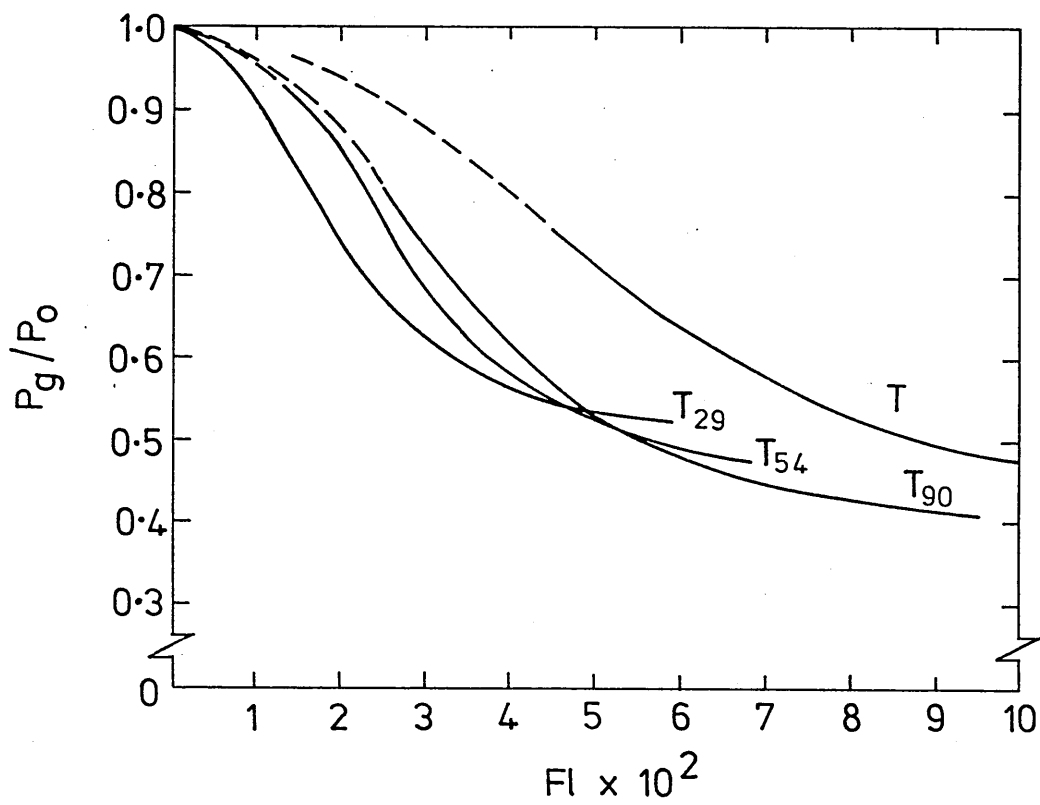


8.6 Continuous mixer example



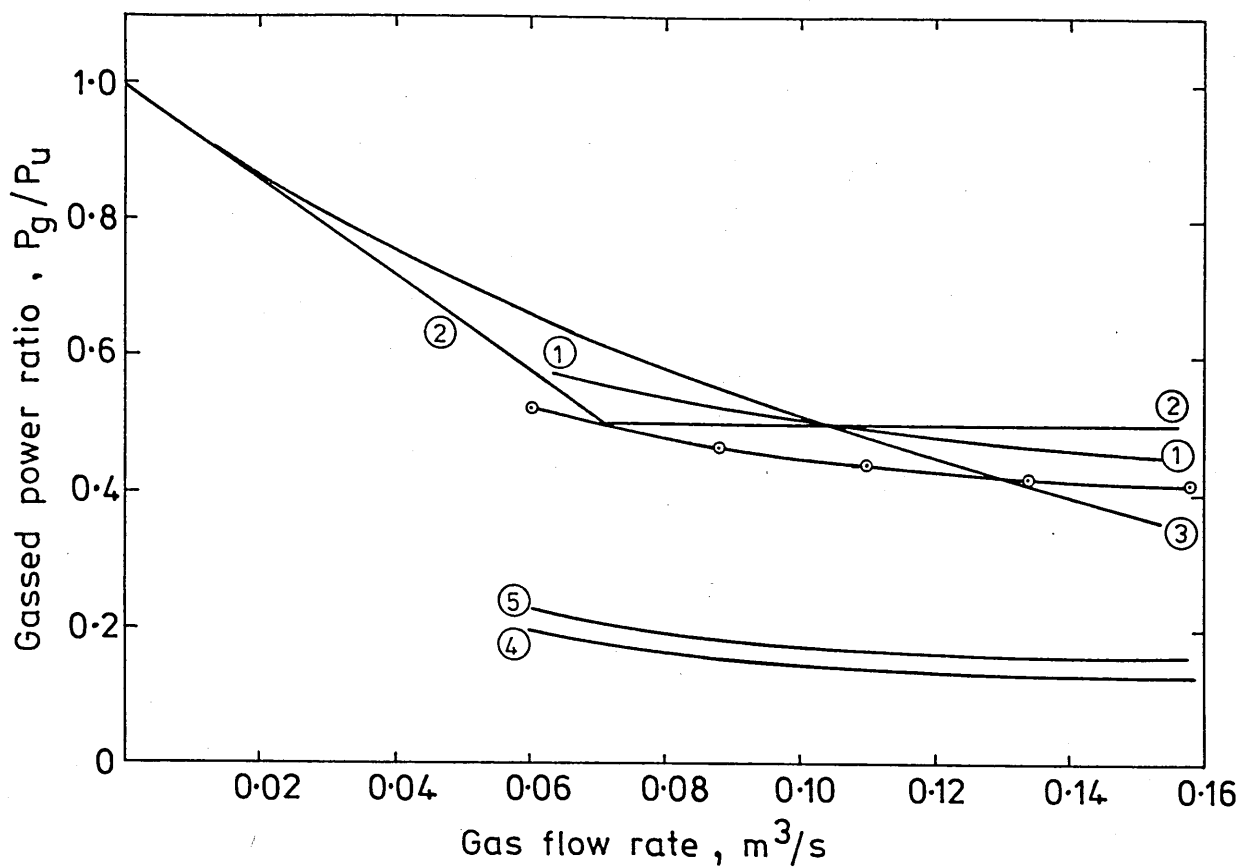






Best lines

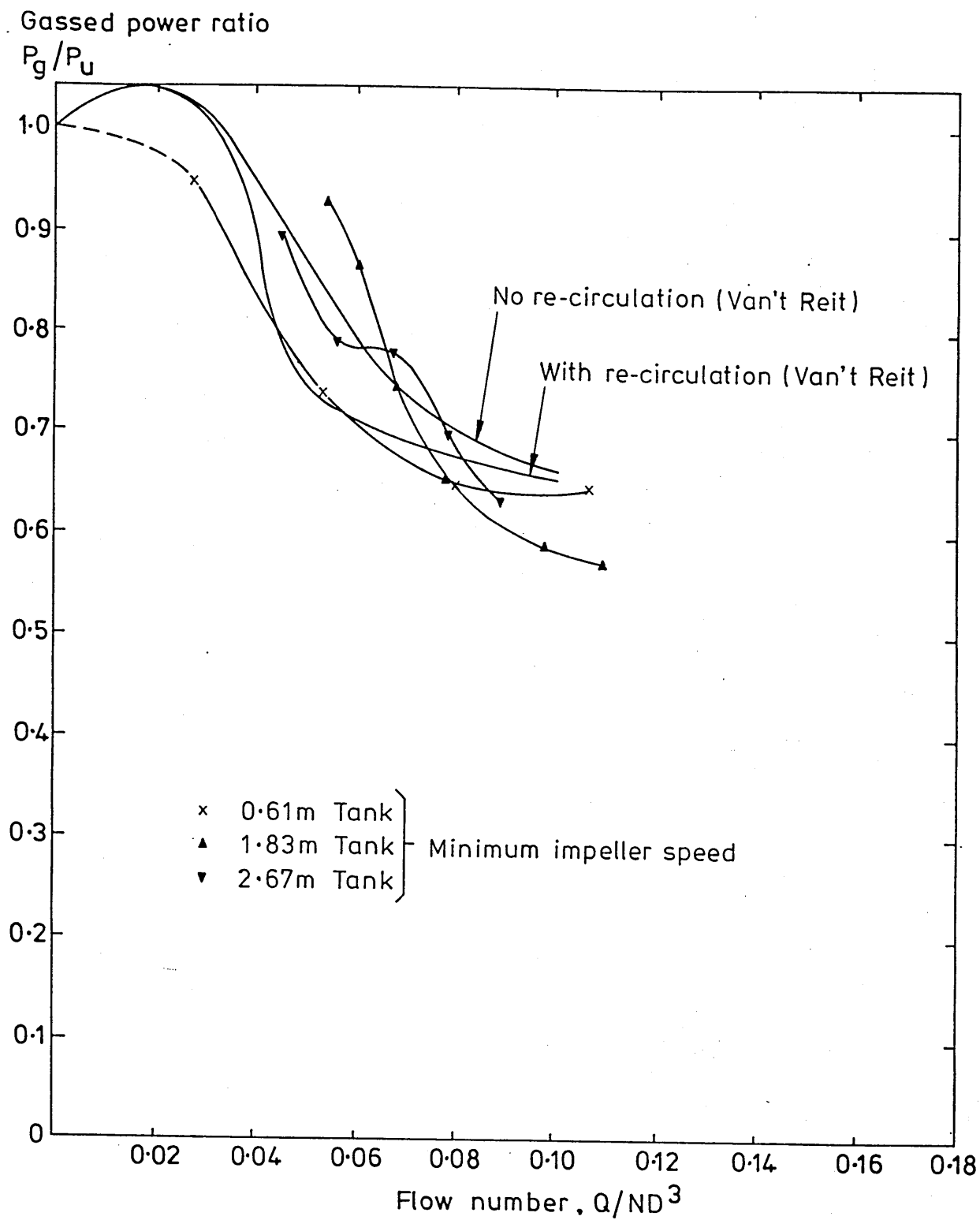
All data $(FI)_R < FI < (FI)_F$



- ① Michael and Miller, 1962
- ② Pharamond et al, 1975
- ③ Nagata, 1975
- ④ Hassan and Robinson, 1977
- ⑤ Luong and Volesky, 1979
- This project

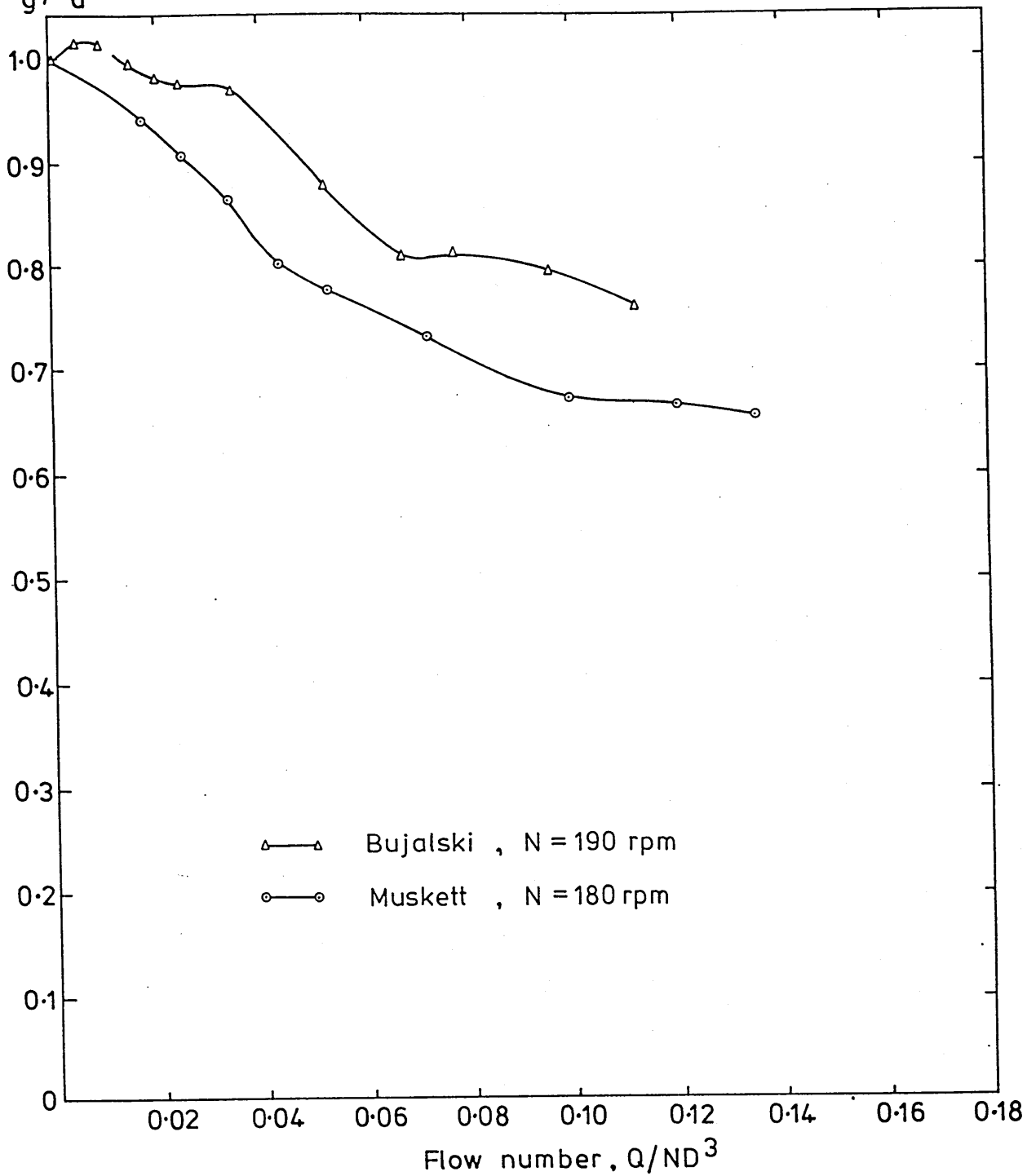
9.4

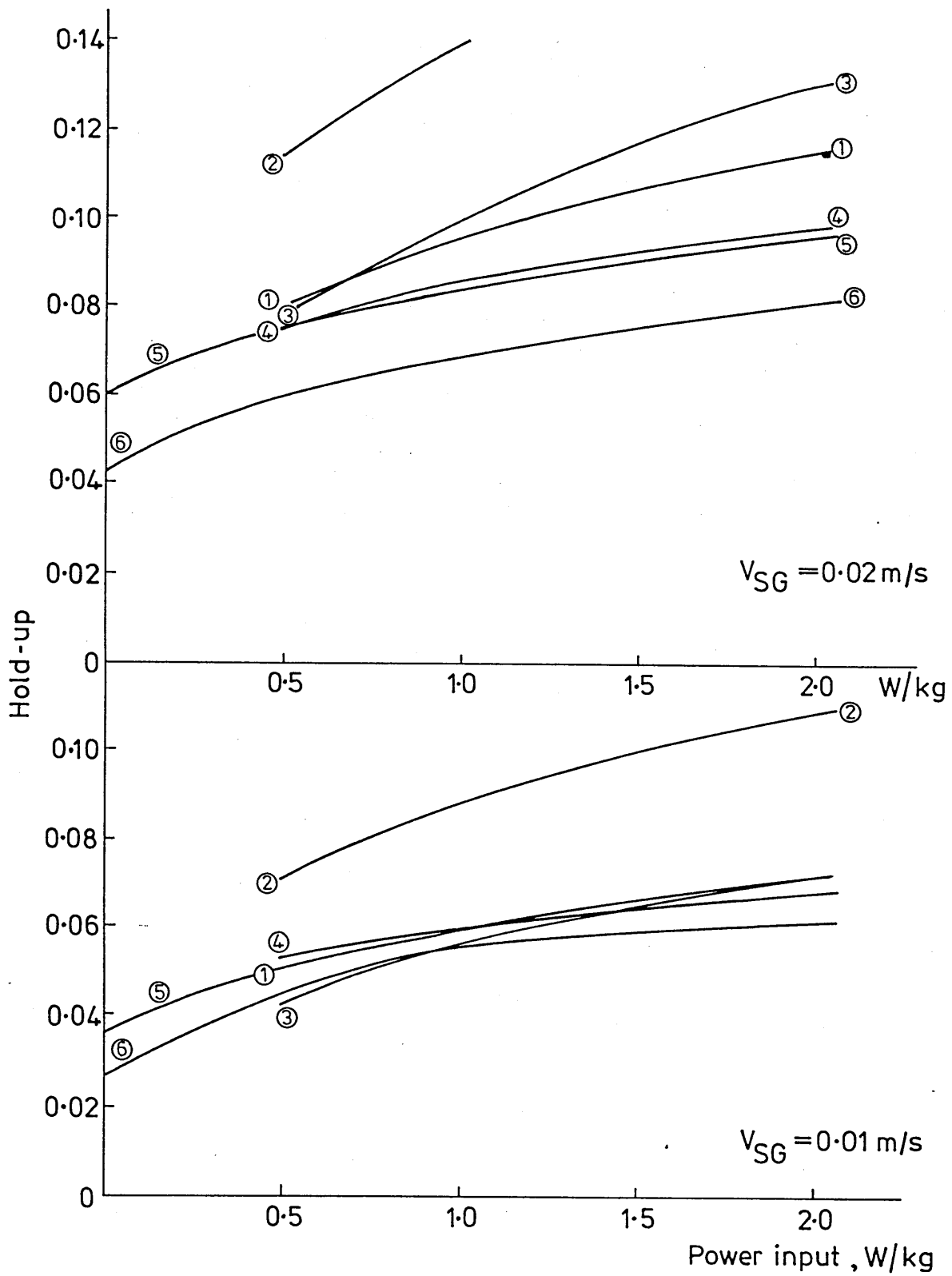
Comparison of literature correlations for gassed power: 2.67 m tank

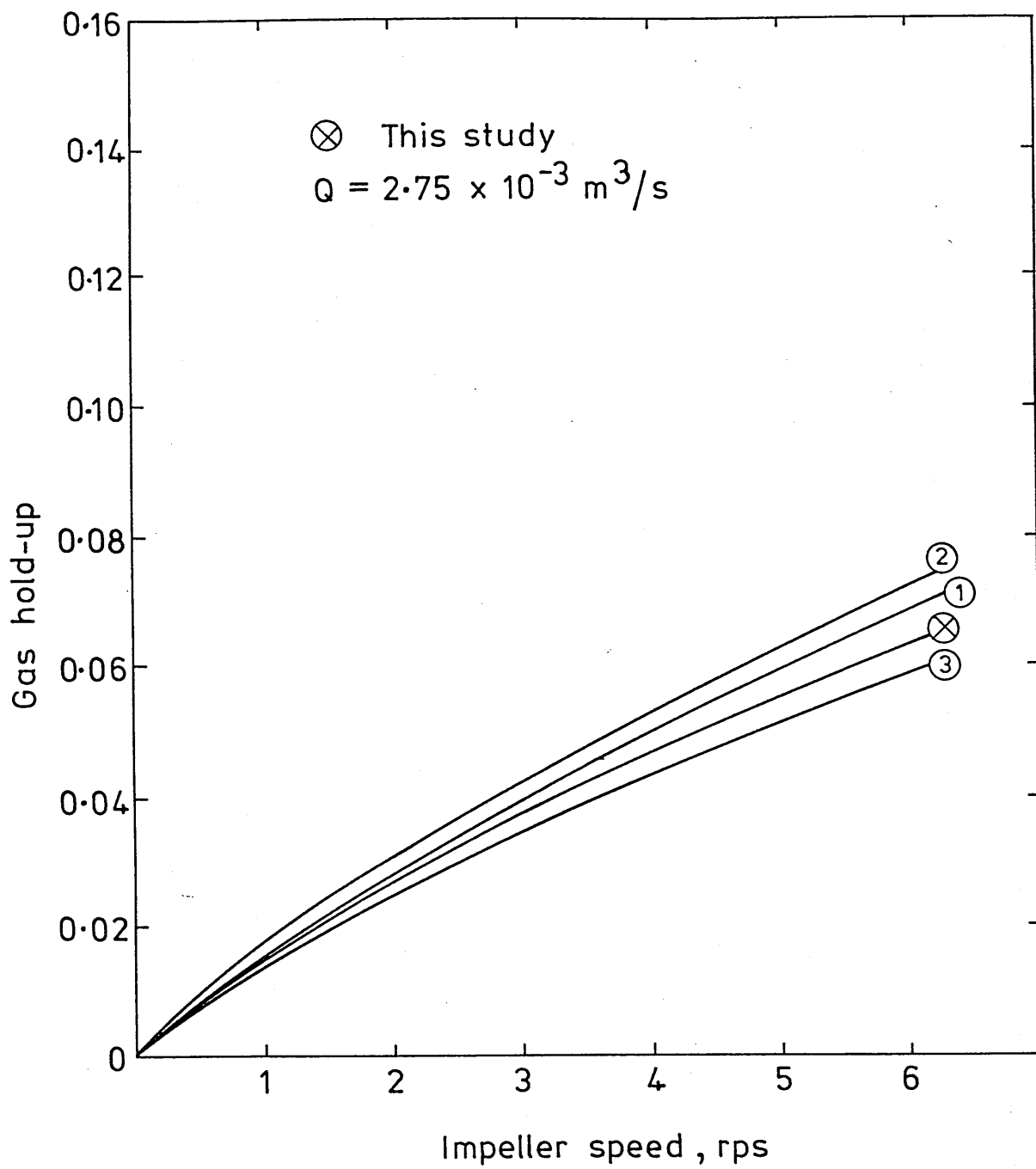


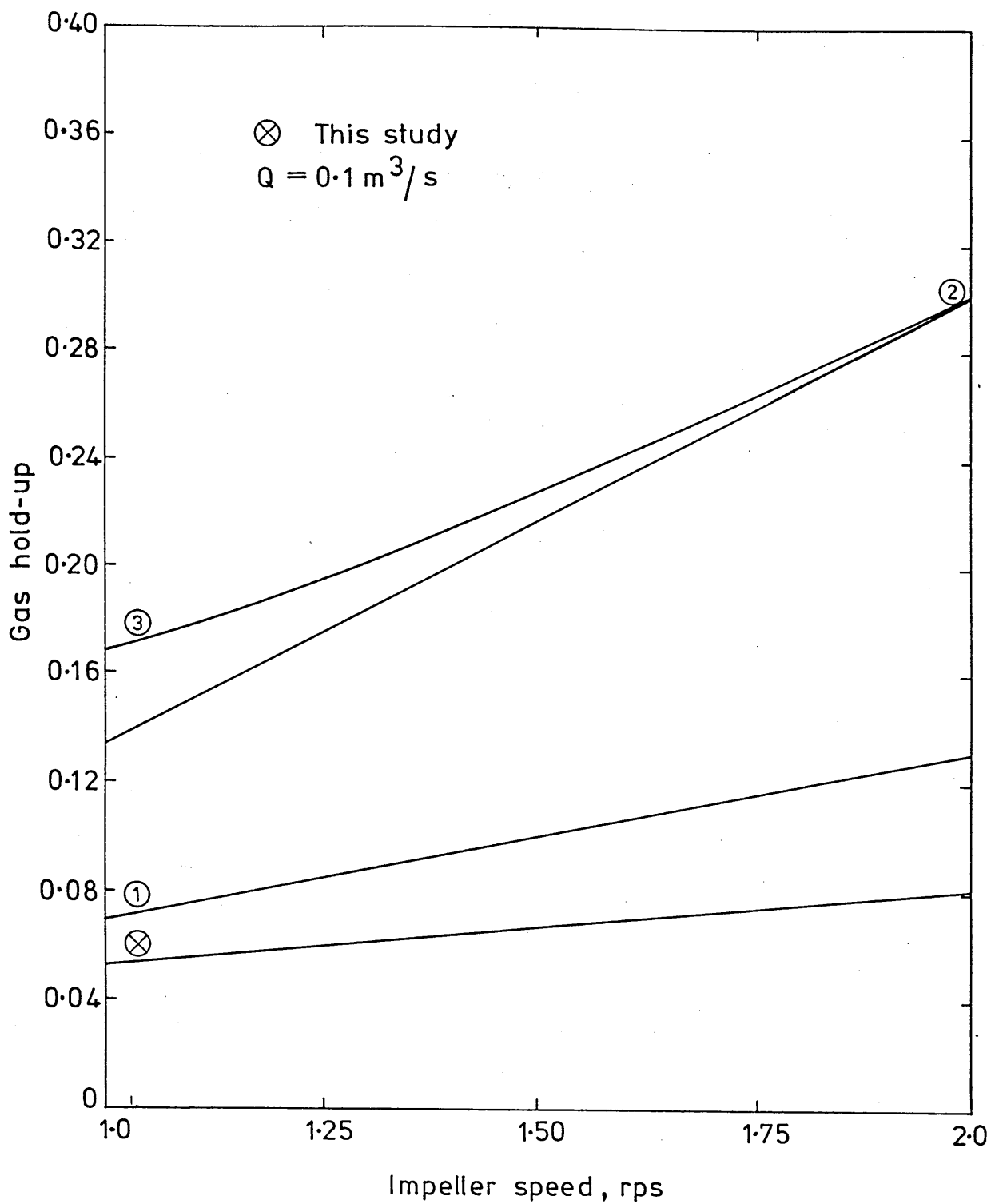
Gassed power ratio

P_g/P_u









$k_L a \text{ (s}^{-1}\text{)}$

