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An Experimental Study of Oil-Water and Oil/Water/Gas Flow in Horizontal and near-Horizontal Pipes

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ABSTRACT

This thesis describes an experimental study and discussion of gas/liquid/liquid and liquid-liquid pipe flow, where the principal objective was to collect three-phase flow data and treat this data by modified gas-liquid methods. The particular application involved is oil/water/gas pipe flow in the oil production industry.

A brief economics study highlights the potential importance of developing new technology to marginal oilfield exploitation, but a literature search revealed very little quality information to be available to industry, in contrast to gas/liquid pipe flow.

A test facility was constructed to investigate oil-water and oil/water/gas flow in small-diameter horizontal and near-horizontal pipes at low operating pressures.

It was found that in many cases modified gas-liquid methods provided satisfactory prediction of three-phase flow regime,liquid holdup and pressure drop characteristics. This result shows variable agreement when compared to the sparse existing oil/water/gas data. Several gas-liquid methods are reviewed for their applicability to systems where the liquid is an oil-water combination. It is suggested that in some cases the chemistry of the fluids can affect the success of using modified two-phase methods.

CONTENTS

CONTENTS - cont'd **Page**

CONTENTS - cont'd

CONTENTS - cont'd

Page

CONTENTS - cont'd **Page**

FIGURES

FIGURES - cont'd

FIGURES - cont'd

FIGURES - cont'd

FIGURES - cont'd

FIGURES - cont'd

FIGURES - cont'd

NOTATION

Symbol	Description	Dimensions
D	Droplet Diameter	L
d	Pipe Diameter	L
e	Correlation Error	-
f	Friction Factor	-
g	Grav. Acceleration	LT^{-2}
h	Liq. Layer Height	L
H	Fluid Holdup	-
L	Pipe Length	L
p	Pressure	$ML^{-1}T^{-2}$
Q	Vol. Flow Rate	L^3T^{-1}
V	Velocity	LT^{-1}
WC	Water Fraction	-
x	Horizontal Distance	L
ϵ	Mixing Intensity	L^2T^{-3}
λ	Input Fluid Fraction	-
μ	Fluid Viscosity	$ML^{-1}T^{-1}$
γ	Slug Frequency	T^{-1}
ω	Wave Frequency	T^{-1}
ρ	Fluid Density	ML^{-3}
σ	Surface/Interfacial Tension	MT^{-2}
θ	Pipe Deviation	-
τ	Shear Stress	$ML^{-1}T^{-2}$
ϕ	Volume Fraction	-

Subscripts	Description
a	In-situ
c	Continuous Phase
d	Dispersed Phase
e	Emulsion
g	Gas
i1	Upper Fluid Interface
i2	Lower Fluid Interface
L	Liquid
L1	Upper Liquid
L2	Lower Liquid

NOTATION - cont'd

Subscripts	Description
m	Superficial Mixture
max	Maximum
O	Oil
S	Slug
Sg	Superficial Gas
SL	Superficial Liquid
SO	Superficial Oil
SW	Superficial Water
W	Water

Dimensionless Groups

Symbol	Description	Definition
N_d	Duns and Ros (1963)	$d\left(\frac{\rho_L g}{\sigma}\right)^{1/2}$
N_{FR}	Froude Number	$\left(\frac{V_m^2}{g d}\right)$
N_{GV}	Duns and Ros (1963)	$V_{sg}\left(\frac{\rho_L}{g \sigma}\right)^{1/4}$
N_{LV}	Duns and Ros (1963)	$V_{sL}\left(\frac{\rho_L}{g \sigma}\right)^{1/4}$
N_L	Duns and Ros (1963)	$\mu_L\left(\frac{g}{\rho_L \sigma^3}\right)^{1/4}$
N_{RE}	Reynolds Number	$\left(\frac{\rho V d}{\mu}\right)$
N_{We}	Weber Number	$\left(\frac{\rho V^2 d}{\sigma}\right)$
X^2	Lockhart-Martinelli (1949)	$\left(\frac{(dp/dx)_{sL}}{(dp/dx)_{sg}}\right)$

CHAPTER 1

INTRODUCTION

The study of multi-phase flow in pipes was subject to investigation much later than it's single phase predecessor, and although the subject was encountered from the earliest scientific developments, particularly those involving steam systems, it's significance increased in the 1960s. The driving force for this surge in interest lay largely with the nuclear generation industry, but since that time the field of multi-phase flow has taken on a much wider significance. This includes the gas/liquid pipe flow of mixtures in process plants and large-diameter, long transportation pipelines. In several cases the second phase does not exist at the pipe inlet, but is a product of heat or mass transfer through the system. This study concerns the application of multi-phase flow to the oil and gas production industry.

In many situations multi-phase flow has little or no application to the transportation of exploited gas and oil reserves. Such cases arise where the reserve may be onshore and/or very large, and as such processing of the produced fluids can be economically undertaken and the fluids transported single-phase to the market in separate flowlines. Increasingly, however, the oil companies are being forced to adopt novel techniques to exploit their reserves for several reasons. Firstly, in several areas, notably the UK North Sea, the oil reserves which remain to be produced are of a volume which are a very small proportion of the average field size of 10-15 years ago. Secondly, the fields may be in difficult geological zones or in very deep water, where traditional technology would result in a prohibitive exploitation cost. In both cases the development of subsea technology has very attractive benefits in terms of field economic viability. This subsea approach is very often coupled to the use of a flowline to an existing production platform which may be 5-20 miles distant. This pipeline will then be transporting a multi-phase mixture, where water and in some cases, sand, will coexist with the gas and oil. It should also be mentioned that in relatively mature fields where water cut is increasing, the presence of water and its' effect on hydraulics and materials may affect the economics of continuing production and projected abandonment. Looking to the future, many of the future reserves are expected to consist of heavy, viscous oils where water can prove beneficial in terms of the transportability of the fluid. Much work has been undertaken in the last 20 years or so to improve the knowledge of oil/gas pipe flow, but very little has been investigated where the pipeline media are gas/liquid/liquid or gas/liquid/liquid/solid. A large bank of knowledge is available concerning liquid-liquid flow in reactors, with some studies of pipe flows also emerging. Work has also been performed on solid-liquid flow for slurry pipelining applications, and gas-solid flow for pneumatic transport applications. Therefore there is a gap in the knowledge which could yield useful information concerning both full wellstream transfer of marginal oilfields and the exploitation of future reserves.

Recognising this technology gap, in 1987 BHRA, Cranfield, proposed a new study to take an exploratory look at oil/water/gas pipe flow, which was a project set up after the initiation of a two-phase general research programme that had proved very popular with a large number of industrial sponsors since 1985. The first stage involved the design and construction of a laboratory-scale facility. The project was sponsored by the UK Department of Trade and Industry and 8 oil and gas operating companies. The original scope of work involving the investigation of dispersion effects was redrawn at the sponsors request to reflect what was deemed to be achievable within the time and budget available and in consideration of the lack of information and suitable instrumentation at the time.

The facility was designed and constructed within budget, and a considerable period of commissioning was undertaken due to the complexity of the apparatus. A series of data collection exercises were undertaken involving the measurement of pipeline pressure drop, liquids holdup, observation of flow regime and for a few flow rates, slug characteristics. The first and second series of tests involved the use of a low-viscosity (2cP) kerosene test oil: both oil-water and oil/water/gas flow were investigated, and later tests involved inclined flow of oil/water/gas mixtures. The study objective was to compare the collected data with predictions from two-phase flow correlations where the oil-water phase is treated as a pseudo-fluid. This was considered the logical approach in that there seemed little point in pursuing further complications if the data suggested it was unnecessary. However, it was judged important to view the results in the light of other data and to help identify where particular facets such as fluids physical chemistry would be of influence. A final series of oil-water and oil/water/gas tests were undertaken in the horizontal test loop where the facility oil was changed to a light lubricating oil of 4cP viscosity, and the experimental variables were modified. The same data was measured and, as for the earlier trials, this involved flow studies in the 50mm i.d., 56m long test loop.

An economics study was also necessary to fulfil the degree requirements, which was carried out under the guidance of the Cranfield School of Management. Chapter 2 presents this study, which involves the economics of marginal oilfield exploitation. Chapter 3 presents the findings of the literature review, where in addition to the review of existing oil/water/gas information, a number of related subjects are also covered. This includes the flow of liquid-liquid mixtures and the behaviour and properties of oil-water emulsions.
The details of the experimental facility design philosophy, equipment specifications and facility construction and commissioning are given in Chapter 4.
The experiments involving horizontal flow of oil-water mixtures are detailed in Chapter 5. This includes the measurement of pressure drop, in-situ liquid fractions and flow regime observations. Several comparisons with existing oil-water data are included in the discussion.
In Chapter 6, the data collected for the horizontal flow of oil/water/gas mixtures is presented and discussed, where 2 test oils are involved. Pressure drop, liquid holdup, in-situ water fractions, flow regimes and, for one oil only, slug

characteristics information is included.
For the lower viscosity oil, Chapter 7 then presents the findings of the inclined (uphill-downhill) oil/water/gas experiments, which includes comparisons with the respective horizontal data of Chapter 6 where appropriate.
The oil/water/gas data of Chapters 6 and 7 is then compared to the predictions from existing two-phase (gas-liquid) methods where the liquid physical properties have been modified using a simple mixing rule for the oil and water. These data and the discussion form the initial part of Chapter 8. Discussion of existing oil/water/gas experimental data is then given to compare earlier data trends with those of the new data. A discussion of the application of modified two-phase methods to three-phase flow is then given, with particular attention focussed on the behaviour of the liquid-liquid combination and it's interaction with the gas phase under different flow conditions.
Finally, Chapter 9 summarises the conclusions of the study, and in the light of information obtained and the review of existing information, presents some recommendations for future work.

CHAPTER 2

ECONOMICS AND CURRENT OFFSHORE PRACTICE

2.1 INTRODUCTION

Dwindling energy resources have meant that increasingly new technology is being utilised to the full to efficiently tap the available fossil fuels and to explore and produce from harsher environments where new reserves exist. The most important fossil fuel - crude oil, is being tapped from geographical locations where even exploration was not considered feasible only 20 years ago. The search and discovery of oil reserves in offshore, very deep waters poses an exacting problem for the oil industry: the development and economic draining of such reserves requires a departure from traditional on-shore, low cost oilfield engineering. On a world-wide scale, the tapping of reserves off Brazil, northern Canada and Alaska and the northern North Sea has required the utilisation and, so, proving, of new development philosophies which broadly share common problems - deep waters and harsh conditions. This report will concentrate on the North Sea scenario only, although many of the facets of the analysis will be applicable to the different scenarios.

The current drive for the production of oil from the UK North Sea has resulted in a significant turn-around in terms of economic feasibility. The older North Sea fields, such as Forties or Brent, contained reserves of around 2000 million barrels (mn bbls) when production commenced. Very few North Sea fields have recently been discovered with recoverable reserves of greater than 400mn bbls, and it is considered that most will lie in the 50-100 mn bbls reserves range. This then being coupled with a subsequent oil-price slump in the mid-1980s has meant a re-think on the technology necessary to tap such small resources. In most cases, the provision of a fixed production platform directly over the field will be uneconomic. The trend currently is to remotely complete a subsea well or wells, and manifold these wells at a subsea template and transport the well fluids to an existing production platform using a subsea pipeline. This then involves the technical feasibility of flowing oil, water, gas and sand simultaneously through a long pipeline. This is, at present, an unknown area of technology and is the basis for the current technical study. This philosophy will be highlighted with economic data provided by a North Sea operator. The impact of the multi-phase flow on the field economics will be examined by utilising a set of basic assumptions, a simple economic field model, and assessing field development economics for different operating scenarios and sensitivities to outside factors such as oil price and currency exchange rates.

2.2 OIL PRODUCTION SCENARIOS

With reference to Figs 2.1 and 2.2, the current method for developing marginal oil reserves in the North Sea consists principally of the following building blocks:

- i subsea completed wells
- ii subsea template/manifold
- iii subsea unprocessed multi-phase flow pipeline
- iv tie-in to an existing production platform.

The control of the production process is normally effected from the parent production platform. The subsea flowline is a full-well stream (FWS) system meaning that no separation/processing of the produced fluids takes place until the fluids arrive at the fixed production platform as shown in Fig. 2.2.

The well fluids usually consist of:

- i crude oil
- ii associated gas
- iii produced water
- iv sand
- v injected chemicals

An illustration of a typical oil reservoir model is shown in Fig 2.3.

The rocks in which the crude oil accumulates in the sedimentary pores is usually underlain by an aquifer which contains brines and is also associated with a gas 'cap' above the oil accumulation. This means that when a well is drilled to the accumulation, oil is driven out by reservoir pressure and, in widely varying quantities, water will also be produced with the oil. Also, as the oil passes up the well and so pressure is reduced, gas is liberated from the liquid hydrocarbon and so an oil/water/gas mixture necessarily passes through the subsea pipeline. Accurate reservoir investigations are vital to predict the production of these fluids, and any change with reservoir pressure maintenance methods such as water injection or gas injection. Sand is also produced with the gas and liquid, but usually is minimised with the use of gravel-packed completions.

Fig 2.4 shows the competing technical restraints during the life of a field. Initially, high reservoir pressure will mean that production may have to be choked back at the production platform. As the reservoir and subsequently well-head pressures slowly decline, the productivity is maintained by opening the platform choke to maintain a steady separator pressure. It is important to realise that the productivity is governed by:

i the pressure decline in the reservoir
ii the pressure drop in the pipeline
iii the amount of water produced from the wells.

Oil production is usually predicted to follow the trend depicted in Fig 2.4 where a rapid increase to plateau production from field start-up is desired. Water production is not usually significant for the first few years but then accelerates and eventually more water is produced than oil and field abandonment becomes a straightforward economic calculation. As suggested in Fig 2.4 the variation of gas production through the field life varies by much less than either the oil or water production.

One of the key issues in the UK North Sea field scenarios is the length of the FWS pipeline or 'step-out'. It is clear that the utilisation of subsea wells and a multiphase pipeline requires an existing fixed production platform to be suitably located. At present, no step-out of 20km or longer has been sanctioned: this is a direct result of the low level of confidence in the multiphase (oil/water/gas) fluid flow technology. Recent suggestions have been made that, should this be extended to 30km, over 90% of future fields would be within 'tie-in' distance of an existing production platform, as indicated on Fig 2.1. This, of course, assumes that the existing facility would have sufficient capacity to process this additional throughout. Fortunately, many of such facilities now operate with considerable spare capacity where tie-in from small fields could be accommodated with little alterations to the production platform. This proves to be one of the strongest incentives to development as it represents minimal capital expenditure, and additionally the platform can provide water injection and gas lift facilities which may be required for the new field.

Of the 3 considerations above, only the latter two will be considered in this study. It is fully recognised that the reservoir mechanics will be of great importance to the field economic study but in this context the influence of the pipeline flow hydraulics due to oil/water/gas flow will be a key variable in the economic analysis.

2.3 OIL PRODUCTION ECONOMICS

Much of what has been outlined above sets the scene for this introduction to oil production economics. In most production economics analysis the oil production forecasts will be assumed a deterministic factor and not open to variations in economic variables such as exchange rates, interest rates, inflation and oil prices.

Since the production life of most North Sea fields is of the order of 10-20 years, all economic forecasts must be undertaken in the form of a discounted cash flow analysis (DCFA) where the time value of money is of prime importance.

The basis for the analysis is the oil production forecast over time and the revenue which this will generate as compared to the expenditure involved in the production system and the amount which the Government gains in taxes and royalties. As this involves a good many considerations these will be dealt with separately.

2.3.1 Production Profile

The production profile (oil) of a typical marginal North Sea oil field is given in Fig 2.5 showing the relatively short field life envisaged. For comparison, the production envisaged from a Norwegian North Sea oil field is given on the same figure: the area under the graphs yields the field recoverable resources. The reason for differing production concepts then becomes clearer. As mentioned, the profile of the small field is the one relevant to this study. Fig 2.6 depicts the profile more clearly, and also shows the predicted water cuts through the life of the field.

2.3.2 Cash Flows and the Time Value of Money

The go/no-go decision for a field development is a corporate decision where risk analysis and careful economic appraisal is required. The generation of economic indicators is vital to allow management to decide which projects will provide the best return on the investment and rank projects against each other for the development queue. As mentioned, the duration of such projects requires that a DCFA be performed. This becomes particularly important as oil field development projects are prone to a severe amount of front-end loading, where return on the original investment will not be seen as a cash flow for several years after field start-up. Also, such an economic appraisal assumes that all of the costs of a project and all of the benefits of a project can be expressed in money terms.

To perform the financial analysis utilising a DCFA the performance of a project is usually compared against an interest which the investment would generate over the same project duration. The requirement then arises to correct all monies back to a common time base, which is usually a nominal present time. Incomes and expenditures are corrected or discounted to a common time base by the use of discount or deferment factors. If the discounted value is referred to as the present value then

present value (pv) = nominal sum x discount factor

When the income or expenditure occurs later than the timebase, then discount factors have values which are less than 1. In the foregoing analysis the income is considered to accrue at the midpoint of the year and the following discount formula is used:

$$\left[\frac{1}{1+r}\right]^{t-1/2}$$

where r is the interest rate for the period.
t is the time in years

2.3.3 UK Taxation Situation

A vital component of the financial equation is the Government tax take on the generated revenue. This is a complex situation which is the province of a taxation specialist, and significant changes have been made in the last 5 years in an effort to help encourage the development of smaller North Sea fields. The two major elements in the tax take are:

- i Corporation Tax (CT)
- ii Petroleum Revenue Tax (PRT)

The generation of gross revenue streams from production forecasts is then offset against operating expenditures arising from that development. In this analysis, the PRT amount is neglected in comparison to that of the CT, as PRT carries influence only when the field is somewhat larger than that involved in the later calculations. Capital allowances which can be offset against CT include:

- i exploration and appraisal wells
- ii plant and machinery.

In i, a 100% allowance is available as this pertains to expenditure incurred before a go/no-go decision, and can be set off as an expense against profits from a related trade. The majority of expenditure is on production facilities, where subsea systems are defined as plant and machinery, and a 25% (declining balance) annual allowance is available. Following these deductions, the operations CT rate is currently 35%. If appropriate, PRT monies are offset against gross revenue in the calculation of corporation tax. Corporation tax is generally treated as being due in the year after which the liability arises.

2.3.4 Project Profitability Analysis

Following the calculations of gross revenue streams, expenditure and tax takes economic calculations must then provide suitable economic indicators which assist in project ranking and senior management decisions on competing projects. One of the most common yard sticks of investment appraisal is the Internal Rate of Return (IRR). This is simply an interest rate, which, used in a discount factor, results in the present value of revenue being equal to the present value of expenditure. Therefore, a project IRR of 10% means that if, at project start, all monies had been invested at 10% interest rate, then the revenue generated would be the same as that generated by the particular project. Project IRR of, say, 40% is obviously a very attractive option to management as compared to an IRR of 10% for comparison. Another economic indicator is the Net Present Value (NPV). This is the present value of the project net cash flow discounted at the return on alternative investments. Therefore, if competing projects provide similar values of IRR, calculations of NPV, undertaken at strictly equivalent re-investment interest rates, provides another indicator of how robust the financial make-up of the competing projects are. Finally, another indicator used is the profit ratio which is the ratio of the present value profit to the present value investment. Only the IRR indicator will be used subsequently in the economic analysis.

2.4 ECONOMIC ANALYSIS

As mentioned earlier, much uncertainty surrounds marginal developments in the North Sea where tie-ins of up to 30km could prove very desirable both in sustaining the UK's position as a major oil producer and minimising reliance on oil imports. The basis of this study will be to examine the sensitivity of a marginal reserve to production uncertainties and uncertainties arising from financial considerations. The base data has been obtained from a major UK operator for a proposed typical marginal development. The field details are given in Table 2-1 for reference.

The projected oil and water production profiles are depicted in Fig 2.6. It is assumed that no revenue is derived from the produced gas; this is a reasonable assumption since for smaller fields produced gas is often used as fuel on the production platform or simply flared off. The production of water has three chief impacts:

- i no revenue is derived from the water
- ii water treatment facilities result in increased CapEx
- iii the formation of oil/water emulsions may lead to high flowline pressure losses.
- iv increased corrosion attack may require expensive exotic materials

To be able to perform analyses highlighting the effects of different parameters on the field viability, we start off from a base case as summarised in Table 2-1. This assumes that the given production profile is met throughout the life of the field, and that the £-$ exchange rate is fixed at a single value for the project duration. The simple tax model described earlier is utilised, and a discounted cash flow analysis (DCFA) performed using a simple routine on a Tulip AT Compact 2 personal computer. Three values of CapEx were used to calculate the oil prices necessary to generate IRR values of 10, 20 and 30%. In all cases, operating expenditure (OpEx) was taken as 10% CapEx/year, following discussions with economic advisers. The output from the economic analysis for Case 1 are given in Fig 2.7 and Table 2-2.

In Case 2, it is assumed that the production profile is identical to the end of year 4 but, thereafter, a production cutback of some 10% is necessary. The reason for this is that, after year 4, water production accounts for over 25% of the liquids produced. This may result in higher pressure losses in the flowline which means that, to maintain a minimum pressure on the production separator with the flowing wellhead pressure the flow rate should be lowered. The very limited information available to engineers suggests that such viscous emulsion problems do not arise until water cut has reached the magnitude above. Similar calculations to Case 1 were made and the output from these calculations are shown in Fig 2.8 and Table 2-2.

Case 3 is identical to Case 1 except for the value assumed for the US dollar-sterling exchange rate (ER). The prediction of exchange rate movements is an all-encompassing subject outside the scope of this study. The importance of this parameter is that US dollars is the international currency for oil trading, and so movement in exchange rates can affect the revenue obtained from oil production. The values chosen represent the maximum and minimum values which have occurred over the past 3 years. The economic analysis was repeated using these new input parameters, and results are shown in Fig 2.9 and Table 2-3.

Finally, Case 4 is identical to Case 2 with the modified ER inputs as mentioned previously. The data obtained are given in Fig 2.10 and Table 2-3.

2.5 DISCUSSION

2.5.1 Discussion of Test Cases 1 and 2

This discussion is based upon the simple economic analysis of the test cases highlighted in Table 2-2, Figs 2.7 and 2.8.

Examination of Fig 2.7 shows the effect of reduced field CapEx on the oil price required to generate different values of IRR. An internal rate of return of 10% would be generally considered a very marginal prospect, especially so when the prevailing financial climate is one of high interest rates. Even so, an oil price of 15$ would be required if the field CapEx was £180mn. A reduction in CapEx of £30mn requires an oil price of around 13$ which is a small though not insignificant improvement. To push the required oil price below 10$ would require a CapEx of £100mn which is a 55% reduction on the base case and would technically be a very tall order indeed. The required oil price for the higher IRRs is obviously constantly increasing.

One point to note here is that reducing the CapEx from £180mn to £150mn with a prevailing oil price of 15$ almost doubles the forecast IRR which is a very significant improvement. To generate an IRR of 30% would require an oil price of some 25$ if the CapEx was $180mn. If one examines the trend of crude prices over a long period, Fig 2.11, it is obvious that the mid-to-late 1970s saw a staggering increase in the price of crude oil. Oil prices of over 25$/bbl were sustained for over a decade, and this indeed coincides with the period in which North Sea exploration activity increased dramatically. However, the relative volatility of crude prices was aptly demonstrated in the mid 1980s when the price crashed to levels of less than 10$/bbl, and subsequently reached over 35$/bbl during the Gulf conflict of late 1990/early 1991. It is obvious that Case 1 is indeed marginal if early 1990 prices were maintained throughout the field life. Reductions in field CapEx, however, can help to swing the economics in favour of development.

The outputs from Case 2, which is the small reduction in production following increasing water cut generally show that an increase of 0.5 - 1.0 $/bbl oil price would be required to give the same values of IRR. This is probably not significant mainly because the oil production is not of a massive scale: early North Sea giant fields would be expected to lose more revenue if such a production cutback was required. The increased water production from large facilities could also result in increased facilities CapEx just at the time when production, i.e. revenue, is starting to fall off from a significant plateau level.

It is interesting to note some historical trends from the North Sea offshore industry which relate field size and the field CapEx via the CapEx/boe (barrel of oil equivalent) parameter. Fig 2.12 shows this parameter for the fields commissioned before 1976. The required $/boe is less than 10 for all cases for which information was available and was under 5 in many cases. This period represents the 'glory' days of North Sea production when all the largest fields were on-stream and when oil prices were around the 25$/bbl (mod) mark. It is also prudent to mention that in many ways the older technology employed at the time to produce the fields would mean that if the same fields were to be developed presently even lower values of $/boe would be expected.

Fig 2.13 illustrates however that in the period 1981-85 the discovered field were generally much smaller and the $/boe had risen somewhat. Two main effects are at work in this case, the first of which is the continuous fall in oil price from 1980 onwards. Also of importance is that the smaller nature of the fields had meant that the late 1970s technology employed to exploit them had resulted in a less efficient operation and a need to cut costs and exploit the fields in a more economic, innovative manner. This is reflected in Fig 2.14 which represents the period post 1985 which shows the continuing trend of smaller discoveries.

However, in the examples in this figure, it does appear that the $/boe parameter has been significantly reduced to generally less than 5. Note that the value of $/boe in the current example, given the upper limit of CapEx, is very close to 5.0. However, one must remember that this parameter does not necessarily reflect possible increased cost in terms of operating expenditure and abandonment.

2.5.2 Cases 3 and 4

Cases 3 and 4 have identical production profiles to cases 1 and 2 respectively, as shown on Table 2-3. The only parameter which has been varied in these cases is the £-$ exchange rate value. Discussion of Figs 2.9 and 2.10 now follows.

Fig 2.9 illustrates that, at the upper limit of CapEx involved, an IRR of 10% would require an oil price of 11.5$/bbl stable through the field life. This is a reduction of some 4$/bbl over the equivalent calculation involving the previous exchange rate value. This is clearly a considerable benefit, and for an IRR of 30% the required oil price is then almost 6$/bbl less than the earlier calculation. The same trend is obtained for the lower CapEx values, as would be expected. Again, the reduction in production by the same amount and at the same time i.e. Case 4, as in Case 2, results in a higher required oil price of 1$/bbl or less as CapEx is lowered. However, these examples illustrate the effect that the exchange rage can exert on the performance of potential oil fields. Unfortunately, the general volatility of the oil price is repeated somewhat in the fluctuation of exchange rates. Over 15 years the £-$ rate has seen values over 2.0 to a dip to close parity between the two currencies. The prediction of the exchange rate is of course part of a very much wider financial and political equation which cannot be explored here. However, it does serve to illustrate the dramatic effect extraneous factors have on the outcome of economic forecasts. The engineer may be able to reduce CapEx by innovative design, but there is nothing he can do about the wider financial and political factors.

2.5.3 Improving the Economics

The previous sections have highlighted the change in North Sea oil economics and the continual drive to improve performance in a climate that is much harsher both from technical and financial standpoints. This section will discuss the ways in which experience, technical input and technical management of resources can help effect cost reductions and so boost revenue.

2.5.3.1 Pipelines/Platforms Installations

As mentioned previously, major savings in capital expenditure may be expected if it is technically feasible to develop the field using subsea facilities. However, this does not mean that the use of fixed platforms is likely to come to an end in the near future. Indeed, many of the small-to-medium prospects being considered for development involve a platform of some kind: it is the function of the platform and efficient design that are crucial.

The use of fixed platforms may be expected to have a considerable impact on future North Sea developments. As mentioned, it is likely that many fields, especially in the 5-50mn bbl (barrel) range, will be developed indirectly through the use of existing infrastructure. Small discoveries close to existing production platforms would look most attractive if a subsea template could be tied back to such a platform. The question then is whether the parent platform has sufficient spare capacity to cope with the increased throughput and, if so, can the probable minor modifications be carried out economically and conveniently? The answer to the first part is probably yes: many of the larger production platforms are over fields which are operating off-plateau where processing capacity is available. The answer to the latter is much more difficult and is likely to be very field-specific.

The role of fixed platforms yet to be built may also differ slightly from the recent past. Whilst large, fixed installations will probably cater for the 100mn bbls plus future fields, the provision of well-head platforms or minimum facilities platforms (MFP) over smaller prospects may increase. In each case the design of the platform topsides is a very important aspect, mainly due to the requirement to reduce weight and have safety considerations at a premium. Many of the small platforms may be unmanned for most of the time, with control from an existing platform. This approach is already used on BP's S.E Forties Development, where the small MFP (Forties 'E') is unmanned and controlled from Forties 'A'. So in this case the result is similar to a remote subsea template, although a platform does provide more operational flexibility. It should also be mentioned that significant improvements have been made in offshore lift capability over the last ten years. Coupled with minimum design and smaller numbers of modules/platform, this results in less time i.e. less expense regarding the cost of installation and so reduced CapEx.

The development of future fields, anticipated to be many in number, will result in considerable activity in terms of pipeline fabrication and installation. Subsea tie-backs, export lines to the main oil/gas trunk lines and water injection requirements place considerable pressure on the pipeline contractors. Extensive advances have been made in pipe laying techniques meaning that less time is required to install the pipeline benefitting both the operator and the contractor. There has also been an increase in the use of flexible pipe subsea and technology may be expected to maintain this.

2.5.3.2 Drilling Technology

The area of drilling has been neglected somewhat in the past due to experience being viewed as a more important asset than technical innovation. In recent years this has been reversed mainly due to the possible benefits arising from extended reach drilling (ERD). Drilling can account for 30% of a field CapEx, therefore any improvement of the efficiency of the drilling operation will be beneficial.

Referring to Figs 2.2 and 2.4, it is apparent that in the early days where platform-completed wells were dominant the size of the oil reservoirs allowed many wells to be drilled vertically from the production platform. The possibility of installing highly-deviated wells on production platforms and subsea templates alike gives a better drainage zone for the well and may mean that wells originally intended to be remote completions could in fact be drilled from the parent platform instead. The use of horizontal drilling is presently attaining field experience and may be expected to improve future production economics.

2.5.3.3 Subsea Solutions

As touched on earlier, the ultimate goal of subsea production and processing and transport directly to shore whilst being economically very attractive presents severe technical difficulties. This has resulted in unprocessed subsea tie-backs of less than 20 km in length.

Utilising natural flow where the well-head pressure is used to drive the fluids to the processing module may be expected to be viable only where the tie-back is short or a high well-head pressure is envisaged. The emphasis then passes to artificial lift methods where some energising of the flow is necessary in order to achieve the fields' production potential. The methods usually considered include gas-lift, electrical submersible pumps (ESP) and multi-phase pumps. Gas-lifting is a technique where gas is injected to the well in order to provide energy, lighten the oil column in the well and so enhance production. This technique has been much employed, although a source of gas and compression facilities are required. The use of ESPs and multi-phase pumps carries a good deal less experience, and may not be expected to have a major impact on economics in the short term. The reader is referred to Darley (1989) and Leggate et al (1989) for a detailed analysis of the potential benefits in utilising the above artificial lift methods in a subsea environment.

Proceeding alongside multi-phase pump hardware developments are similar projects involving subsea separation/processing. This technology requires separation of the oil/water/gas subsea, and subsequent boosting of each phase for single-phase transport to distant platforms or onshore terminals in several pipelines. Again the developed hardware remains unproven in a subsea environment at the present time.

2.5.3.4 Project Management

As well as cutting manpower/hardware costs by the use of more elegant technical solutions, future savings may be expected to be made by more efficient use of resources such as service companies and better collaboration between project partners.

Operators will require improved performance from all aspects of engineer contractual work, particularly in concept design studies where up-front man hours, if not spent effectively, may delay the project design phase and any anticipated future installation schedules. It is expected that platform manning levels will be reviewed and particularly with new developments and safety considerations, better definition of the proposed work may be expected to improve the client/contractor relationship and so progress the project more smoothly. Competition for contracts may be fierce with the oil companies taking a tougher stand than on earlier years, with emphasis on work quality, performance in terms of time and budget, and a determination that higher-than-necessary contract prices will be much reduced compared to previous years.

2.5.3.5 Government Tax Take

As noted previously, the taxation to which oil production activities are subject is complex and the application has altered somewhat as the UKCS became a relatively mature oil province. Whilst it is not possible to predict what form future tax measures will take, it is reasonable to assume that the government will take steps deemed necessary to sustain offshore activity whilst at the same time maximising the effect of oil royalties on the UK economy in general. The example considered in the economic analysis includes the up-to-date situation, and it is worth pointing out that if the field had been subject to previous taxation methods, the methods would have been unworkable principally due to the relatively small size of the oilfield.

2.6 CONCLUSIONS

The purpose of this chapter has been to demonstrate, at a simple level, the changing economic situations of the UKCS oil production and the requirement to match the harsher economics with technical solutions to economically drain the oil. The following comments are made regarding future UKCS oil exploration:

i Very few fields yet to be developed could profitably sustain a constant oil price of less than 10$/bbl.

ii In the given example, a drop of CapEx from £180m to £150m would almost double the IRR from 10 to 20% at an oil price of 15$/bbl.

iii Previous levels of CapEx to exploit fields is unsuitable for developing marginal prospects where more efficient development approaches are increasingly being used.

iv Oil price increases above 20$/bbl would have significant effects on the prospects of development of future small fields.

v Small production cutbacks on a marginal held can require up to an extra 1 $/bbl on the oil price to secure the same value of IRR.

vi Production economics of a small field can be significantly altered by the sterling-dollar exchange rate value.

vii The extended use of subsea tie-ins, extended reach drilling and the development of multi-phase production hardware will significantly benefit small oilfield developments on the UKCS.

CHAPTER 3

A LITERATURE REVIEW

3.1 INTRODUCTION

The flow of oil-water-gas mixtures in pipes involves a large number of technical topics including fluid mechanics and fluid physical chemistry aspects. Whilst a comprehensive review of the literature covering such a scope is inappropriate here, work from the literature which was felt to be of relevance to a pipe-flow situation has been included where possible. A very large number of studies have involved two-phase gas-liquid flow and several excellent books have been produced on the subject. However, it was judged inappropriate to include this information at this point due mainly to the volume available and a desire to focus on the particular oil-water and three-phase aspects at this point. This does not preclude the examination of relevant work in future chapters where necessary. This review aims to encompass aspects involving oil-water flow, oil-water dispersions/emulsions in pipelines and oil-water-gas pipe flow. As would be expected, many studies touch on several of the above areas, and so the attempted chapter breakdown will necessarily involve some overlap. These subjects are of potentially vital importance to the study, and the author is unaware of any works where a discussion of the above technical range is presently available.

3.2 OIL-WATER FLOW - MACROSCOPIC ASPECTS

A large base of knowledge is available concerning the flow of oil-water mixtures in a wide range of test apparatus. This section aims to examine work where experimental or modelling activities have essentially treated the problem globally in that no detailed investigations of, for example, dispersion characteristics have been undertaken.

The flow of liquid-liquid, primarily oil-water, mixtures in pipes has received much less attention from researchers than the related field of gas-liquid pipe flow. The flow of an oil-water-gas mixture may depend on how the oil and the water phases behave as the total liquid phase, and so this section is of interest to gas-liquid-liquid flows.

Many of the earlier studies were undertaken in Canada. Russell et al (1959) reported on experiments from the University of Alberta. A test pipeline of 1-inch i.d. and 28ft long, with water and a mineral oil of 18cP viscosity, formed the test system. Pressure losses and flowing fractions of oil and water were measured and flow patterns determined by visual observations. For the range of flow rates involved, the flow patterns were classified as bubble, stratified or mixed as shown in Fig 3.1. The boundary between bubble and stratified flows resulted in inflection points in the pressure drop characteristic, with the mixed flows producing wide data scatter: the authors attribute this to the transitional state between fully stratified and fully mixed flow. In-situ liquid holdup results suggested that in the laminar region the holdups were independent of superficial water velocity whilst in the turbulent regime some dependence on water superficial velocity was noted. The authors finally attempt to compare their results to those of gas-liquid studies which, at the time, were relatively few in number. Russell and Charles (1959) then presented a brief analysis of oil-water pressure loss, concerning particularly the influence of the less viscous liquid phase: only laminar flow is examined and no data was available for comparison.

Charles (1960) gave a further account of oil-water flow in the experimental system of Russell et al (1959). The oil in this case was a crude oil of typically 500-800cP viscosity at flowing conditions, with Newtonian behaviour reported. For input water contents of 35 to 60%, the pressure gradient was found to be up to about 10% of that occurring when oil-only was flowing in the pipe at the same volumetric throughput. Again, highly variable pressure losses were noted for water contents up to about 35%, which is attributed to flow instabilities. A number of field investigations in a 2.5-inch i.d., 565ft long pipeline revealed pressure drop trends which were very similar to those found in the laboratory pipeline, although the author does point out that laminar motion of the oil was encountered in all tests. This paper is also available in a different publication (Charles (1961)).

Charles et al (1961) reported experiments on oil-water flow where the oil and water had equal densities. Test oils of 6.29, 16.8 and 65cP were used with density modified with the addition of carbon tetrachloride. Pressure drop, holdup and flow pattern were recorded from the 1-inch i.d., 24ft long test pipeline. Flow regime was found to be largely independent of oil viscosity: the flow pattern map for the 16.8cP case is shown in Fig 3.2. Pressure loss data revealed a reduction in pressure gradient upon water addition to a minimum value which was considerably higher than predicted by earlier theory.

Glass (1961) reported tests involving "core" flow of oil and water in a 1cm i.d., 4ft long horizontal glass pipeline. Oil viscosities varied from 10 to 30cSt with oil specific gravities very close to unity. A minimum in pressure loss was obtained at 35% water fraction, although the lower viscosity oils were seen to form a less stable emulsion.

Gemmell and Epstein (1962) tackled the stratified oil-water flow problem by developing a numerical analysis employing a finite-difference technique where only laminar flow was examined. The model gave predictions of local velocity profiles, holdup ratios and pressure gradient, and comparisons were made with the data of Russell et al (1959). Satisfactory agreement was obtained where the flow was oil (laminar) - water (laminar). Any departure from water transitional superficial Reynolds numbers towards turbulent behaviour markedly reduced the accuracy of the prediction. The authors also note the opinion that the data of Charles (1960), reviewed previously, where pressure gradient reduction was obtained even for turbulent water behaviour, was actually not involving stratified flow at all but a concentric oil-water slug type of flow as obtained by Charles et al (1961).

Charles and Lilleheht (1965) gave an interesting account of the interfacial wave phenomena observed in co-current liquid/liquid flow. The test facility consisted of an 8-inch wide by 1-inch high rectangular conduit, with water and a refined mineral oil of 5cP viscosity as the test fluids. Laminar-turbulent transition was seen to be affected by the introduction of the second liquid phase. The effect of flow rates of each phase on the nature of the interfacial waves and the turbulent behaviour are described with the aid of photographs of a selection of the test runs.

Charles and Lilleheht (1966) then presented a correlation for the stratified oil-water pressure drop where one of the phases is in turbulent flow, thereby representing an advance on previous methods. The authors subsequently took the approach employed for gas-liquid flow which was developed by Lockhart and Martinelli (1949). Correlating curves were produced, using a wide range of earlier experimental data, of

$\varnothing^2_L$ - X^2 and
$\varnothing^2_M$ - X^2

where

$\varnothing^2_L$ is the ratio of the two-phase pressure loss to the pressure loss for the less viscous phase flowing alone.

$\varnothing^2_M$ is the ratio of the two-phase pressure loss to the pressure loss for the more viscous phase flowing alone.

and X^2 is the ratio of the pressure loss of the more viscous phase flowing alone to that of the less viscous phase flowing alone.

The data fall below the curves developed for gas-liquid flows, but the authors urge that it is a useful correlative method nonetheless.

Darby and Akers (1966) presented a brief experimental account of oil-water stratified flow. Flows were measured in a PVC channel 15ft long and 11½ x 2½-in in cross section, with water and kerosene as the working fluids. Measurement of velocity profiles and interfacial shear stresses were the experimental objectives. Interfacial shear stresses computed from velocity profiles in each liquid produced slightly different values for the interfacial stress: this is attributed to a small amount of interfacial wave action. Interfacial shear values were found to be an order of magnitude smaller than the wall shear values, and the authors tentatively suggest that the interfacial shear characteristic is primarily a function of the upper fluid flow behaviour and not that of the lower fluid.

Achutaramayya and Sleicher (1969) presented another analysis of stratified oil-water flow involving both circular and non-circular pipes. No experimental data is involved, but one interesting outcome is that for oil/water viscosity ratios in excess of 100, greater pressure maintenance is exhibited by a flat-bottomed pipe as opposed to a circular pipe.

Yu and Sparrow (1969) performed experiments on oil-water stratified flow in a horizontal duct. The duct was 25ft long, but only 9/16-in height and 1 1/8-in width. The presence of waves and their effects on pressure drop calculations supported the views put forward by earlier investigators. As to the onset of interfacial waves, the authors suggest that this is due to instabilities at the interface itself, rather than to turbulence within the component flows.

The experimental system used by Charles and Lilleheht (1965) also formed the basis for a study by Stellmach and Lilleheht (1969). Velocity fluctuations were measured with a hot-wire anemometer. A stability criterion was used to compute transition zones for each phase where the transition for one of the liquids is computed assuming laminar flow in the other liquid. Sinusoidal waves were observed at low Reynolds numbers, which grew in amplitude as turbulence was approached, but which were found to be of constant frequency. This work was then summarised in a later publication by Stellmach and Lilleheht (1972).

Guzhov and Medvedev (1971) presented a brief paper on the pressure loss from immiscible liquids pipe flow. The authors consider the three-layer concept i.e. pure oil and water components with an oil/water emulsion at the interface, to be applicable and the resulting equations for one-dimensional momentum conservation are presented and further equations proposed to close the problem. However, this restricts the analysis to stratified flow and the composition and thickness of the emulsion layer must be known *a priori*.

Guzhov et al (1973) presented findings of an experimental study of oil-water flow. Laminar and turbulent flow of water and a 22cP viscosity transformer oil in a 39.4 mm i.d, 18m long test loop were involved. The pipe was of steel construction where pressure loss was measured and flow regime observed at a transparent observation section. The flow regime map is displayed on Fig 3.3 with the associated key in Table 3-1. The pressure drop data was qualitatively supported by the flow regime observations, with the dense, unstable emulsion giving rise to an increased pressure drop. Finally, it is reported that an early turbulising of the more viscous liquid was obtained, at a Reynolds number of approximately 1500.

Wicks and Fraser (1975) studied oil-water entrainment in pipe flow with the objective of avoiding settled water and so the possibility of pipe corrosion. Experiments were conducted in a 1-inch i.d. pipe using kerosene and water, mainly to observe water film breakup at pipe upward inclines. An existing criterion was adapted for the entrainment and transport of water droplets where it was assumed that once entrained, water droplets behave like solid particles. The Hinze (1955) critical Weber number equation was used to predict the largest stable droplet (which is the most difficult to entrain). A simple graphical method is employed to give predictions of the oil velocity required to entrain settled water, and the authors support their method with reference to several qualitative field examples. Tsahalis (1977) conducted a study which was also aimed at predicting entrainment and stability of separated water films in oil-water pipelines. Key assumptions in the theory developed are that the ratio of the water to oil viscosity is small compared to unity and that the water film is thin enough for the interfacial velocity to be well represented by the oil-only velocity at the interface location. The theory first calculates when flow at the film becomes unstable for laminar, turbulent, and transition flow. Break-up of the oil-water interface utilises the theory of Hinze (1955), and dependencies on variables such as oil physical properties and pipe diameter are illustrated with examples. Prediction for a transformer oil of 15 cP viscosity was found satisfactory, whilst those for a 2 cP oil showed large deviations when compared to the experimental data.

Malinowsky (1975) reported on the first of a series of studies at the University of Tulsa on oil-water pipe flow. The test loop consisted of 97ft of 1.5-inch i.d. horizontal piping with working fluids water and a diesel oil. Pressure drop, flow regime and in-situ liquids fractions were measured. The pressure drops were examined against the prevailing flow regimes, and the author noted that, for several flow patterns, different assumptions of which liquid viscosity value to use (assuming no-slip/slip/linear-weighting) could dramatically affect the predictive accuracy of existing pressure drop correlations. Data from a previous study, which involved lower flow rates, was examined and revealed the same general trends. Laflin and Oglesby (1976) used the same apparatus to make further similar investigations. The observed flow regimes showed variable agreement with Malinowsky's map, and the back-calculated apparent liquid viscosity was found to be sensitive to the liquid mixture flow rate.

Mukhopadhyay (1977) investigated oil-water flow in inclined pipelines. The test system was 1.5-inch i.d. transparent pipe arranged in an inverted U-shape, where each leg was 45ft long. Uphill and downhill angles of 30-90 degrees were involved, with pressure gradient and water holdup measured in both legs simultaneously. The test fluids were water and a diesel oil of viscosity approximately 5cP at operating conditions. Frictional loss was found to be dependent on oil/water ratio and pipe inclination - see Fig 3.4 as an example. Analysis of the holdup data for slippage behaviour suggested that the most important slippage behaviour will take place between -30 and +30 degrees: this study did not examine this inclination range.

The water holdup data yielded empirical correlations which are restricted to the inclinations involved in the experiments. Some years later Mukherjee et al (1981) summarised the findings in a journal publication.

Oglesby (1979) also performed experiments on horizontal oil-water flow. The test loop was 1.5-inch i.d. with an 84ft long, acrylic working section with extensive oil and water clean-up equipment at the process section. The behaviour of 3 oils of viscosity 32-167cP was examined with particular attention to flow regime, pressure loss and emulsion inversion characteristics. The observed flow regimes were given a variety of classifications and the 32cP map and associated key are shown in Figs 3.5 and 3.6. The flow patterns were found to be a function of oil viscosity and interfacial tension with a weak dependence on mixture velocity being observed in some cases. The inversion phenomenon was found to dominate the pressure drop.

Wang and Charles (1981) performed more recent studies of stratified oil-water flow concerning laminar-laminar and laminar-turbulent behaviour. Experiments were performed in a horizontal duct 8.5m long, and width 10.2cm and depth 12.7mm. Some tests involved a lower phase of ethanol-water mixture at low interfacial tension, but pronounced waves were always present at the interface. The velocity distribution and pressure drop data lead the authors to state that the early Charles and Lilleheht (1966) pressure gradient correlation gives satisfactory predictions, but also that work should now focus on turbulent-turbulent flows.

Vigneaux et al (1988) presented an account of oil-water flow in a large-bore, inclined pipeline. The test line was 8-inch i.d., 14m long with water and kerosene as the working fluids. Water fractions of 30-95% and inclinations from the horizontal of 35 degrees or more were involved, and flow rates were kept sufficiently low to avoid oil/water emulsification. At low inclines and 10% or more input oil fraction, an intermittent flow nature was observed, where swarms of oil droplets, which did not coalesce, move at a greater velocity than the water. The oil droplets were of 2-6mm diameter. The workers found that total liquids flow rate had a weak influence on the relative oil-water velocity, but admit that higher flow rates will be likely to influence the slip velocity to a greater extent.

Zavareh et al (1988) also investigated inclined oil-water flow in a large diameter pipe. The facility was 42ft long and 7¼-inch i.d., and an oil of 2.46cP viscosity and distilled water formed the test fluids. The greatest deviation from the vertical reported is 15°, therefore the results are outside of the range of the present study. However, it is pertinent to note that these authors compared their flow patterns to the predictions of Taitel and Dukler (1976), developed for gas-liquid flow, and found that the model failed to properly account for the flow regime characteristics observed.

Arirachakaran et al (1989) presented a paper which utilised the wide range of oil-water flow data which has been collected at the University of Tulsa. A correlation is proposed for the prediction of inversion point in an oil-water dispersion (see later). Pressure gradient prediction models are also presented for the cases of stratified and fully mixed flow which compared favourably with a range of experimental data.

Recently, Brauner and Maron (1989) presented the first fully-theoretical model concerning oil-water pipe flow. The approach embodies the Taitel and Dukler (1976) separated-flow technique which was successfully applied to gas-liquid pipe flow, but in this case attention is restricted to stratified flow only. Momentum and continuity equations are developed for each layer, and the result is that for horizontal flow the equilibrium liquid depth is a function of 3 parameters, namely:

X^2 - Lockhart-Martinelli parameter
n - laminar or turbulent flow behaviour
$\emptyset$ - ratio of oil superficial velocity to water superficial velocity.

The authors point out that for the gas-liquid flow, the last parameter above does not appear : this is due to the larger difference between gas velocity and liquid velocity than is expected for stratified liquid-liquid phases. The model is tested against independent data only, involving both laminar and turbulent flow, and generally satisfactory comparisons for holdup and pressure gradient are obtained. The pressure gradient reduction which can be obtained, for moderate-to-high oil viscosities, is shown to be dependent upon whether the flow is laminar-laminar or turbulent-turbulent. The laminar-laminar model suggests higher pressure loss reduction factors than when turbulent flow is involved, which is consistent with available experimental data. At very high oil viscosities, the maximum pressure drop reduction was found to approach an asymptotic level.

3.3 OIL-WATER FLOW - MICROSCOPIC ASPECTS

The study of emulsions/dispersions is a science which has attracted a vast amount of research effort over the last 40 years or so, this is due to the importance of dispersions in a great many industrial processes. Sherman (1968) and Becher (1965) represent two of the classical textbooks concerning the study of emulsions. The review here is restricted to studies of emulsions pertinent to determination of their physical properties and their characteristics as regards flow in pipes.

3.3.1 Droplets and their Behaviour

Much work has concentrated on the stability of the largest droplet when subjected to different forms of shear flow, such as in a pipe or in a stirred vessel. The classic groundwork was laid by Hinze (1955), who investigated droplet break-up in an air stream and in turbulent pipe flow. Physical arguments are proposed as to the controlling mechanisms of inertia and viscous forces. Assuming non-coalescing conditions and isotropic turbulence across the flow field, equations are given for the maximum stable droplet diameter. The principal relation is:

$$D_{max} \left[\frac{\rho_c}{\sigma} \right]^{0.6} \epsilon^{0.4} = C \qquad (3.1)$$

Determination of the constant in Eq.(3.1) was made by examining the droplet breakup data of Clay (1940) using concentric cylider apparatus where the inner cylinder rotated. Correlation of the data resulted in the value of the constant C = 0.725, although for many cases this still resulted in appreciable scatter. Several years later Sleicher (1962) examined the equation and derived a relationship which differed markedly from that of Hinze. The author suggested that extrapolation of the Hinze equation to fit Clay's data had been questionable and commented that he had found droplet break-up to be most prevalent at the pipe wall, which conflicted with Hinze's assumption of isotropic turbulence. Collins and Knudsen (1970) later collected pipe flow data which was predicted very well by the Sleicher relation.
Swartz and Kessler (1970) presented an account of droplet break-up in a liquid-liquid system. The system involved dispersing a single droplet in a continuous water flow in a 1.5-inch i.d., 40ft long transparent pipe. The location and manner of break-up was noted, with the predominant mechanism being break-up near the pipe centre-line into two smaller droplets. Correlations for the increase in interfacial area accompanying the break-up are given. The authors report that the droplet size distribution was approximately normal, although the fragmentation process was observed for a very limited timespan.

Sevik and Park (1973) presented data on the break-up of gas bubbles in a water jet. The authors develop equations based on the resonance of gas bubbles in the flow which agreed excellently with their experimental data. The theory was then extended to the liquid-liquid break-up data of Clay (1940) and the result produced was very close to that previously obtained by Hinze. Kubie and Gardner (1977) made a study of droplet distributions in straight horizontal tubes and helical coils. Their droplet breakup data was also in very good agreement with that of Hinze, even for the flow of liquid/liquid in the helical coil where a modified friction factor was necessary. This system differed somewhat from previous tests not only in terms of the geometries involved, but also due to the fact that water was, in one case, more viscous than the other liquid (n-butyl acetate).

Karabelas (1978) also carried out a study of droplet size distributions in pipe flow of liquid-liquid mixtures. Water was dispersed in two hydrocarbons; a kerosene of 2 cP viscosity and a transformer oil of 20 cP viscosity. The test loop was 5cm i.d. and 32m long, and droplet size distribution was measured using a special droplet encapsulation and photographic technique; the water concentration was 0.2%

(vol). Maximum droplet sizes were correlated excellently by a modified Hinze equation, which also performed well when tested against an independent data set : over the range 1-3 m/s liquid velocity the more viscous oil resulted in smaller mean and maximum droplet sizes. All droplet size distributions were represented well by a fitted Rosin-Rammler distribution.

Hesketh et al (1987) gave an interesting account of droplet and bubble breakup phenomena. The authors comment that the classic Hinze (1955) criterion produce critical Weber numbers of close to unity in a variety of pipeline studies involving liquid-liquid phases, but when the system is one of gas-liquid, much higher critical Weber numbers are obtained using the Hinze approach. The authors state that the critical Weber number should depend only on the breakup mechanism and not on the fluid physical properties, and experimental observations (independent) support this statement. The critical Weber number approach of Levich (1962), which includes the density ratio of the dispersed to continuous phase, is scrutinised in a similar manner, with the result being that for liquid-liquid and gas-liquid systems similar critical Weber numbers are obtained of close to unity. A general equation is developed for mean droplet size, but this is restricted to the special case of non-coalescing conditions.

Hayes (1988) also performed a study of the breakup behaviour and motion of droplets in turbulent shear flows. An experimental rig was built, based on the design of Taylor (1934), where a suspended droplet is subjected to shear from either 2 parallel rotating bands or a set of 4 rotating rollers. Droplets of distilled water or seawater were sheared in an oil phase continuum, and observations of droplet deformation suggested that at small deformations a linear relationship between deformation and shear rate holds. Droplets were observed to break up in a 'tip-streaming' manner where the droplet became increasingly ellipsoidal as shear rate was increased, until at a critical shear rate fluid was ejected from the pointed ends. Comparison of maximum droplet size data with Eq (3.1) gave the constant C to be 0.68, which is very close to the Hinze (1955) result. A computational model was then developed to predict the movement of the water droplets in fully developed turbulent pipe flow.

Recently, Hanzevack and Demetriou (1989) presented a paper where the incentive was to find the extent of oil-water mixing in typical pipeline networks where horizontal flow, vertical flow, and flow after bends are involved. Water was dispersed, at low concentration, into a low-viscosity kerosene flow through an 8.2cm i.d. pipe with measurement via a pipe-imaging technique. Data revealed that the maximum droplet size was more dependent on the number of upstream interactive bends than on the mixture velocity, with an increase in the number of bends producing smaller droplets and an adequately dispersed system. However, for flow in the horizontal pipe alone, adequate dispersion was only obtained at velocities above 2.3 m/s.

The paper already encountered, by Arirachakaran et al (1989), includes a correlation for the water fraction required to invert an oil-in-water dispersion, and is depicted in Fig 3.7. The correlation was based on a variety of pipeline/mixer tests and involves a very wide range of oil viscosity, although it should be noted that laminar flow of the oil is involved in all the experiments which compose the correlation. Where an oil-water dispersion will invert is dependent upon many fluid-mechanical and fluids-chemistry aspects, but Fig 3.7 does display an interesting trend. It should also be noted that the interfacial tensions of the fluid systems are not all available.

Work reported by Hesketh et al (1991) shed more light on the aspects of particle breakup in a turbulent pipe flow. These authors injected air bubbles into a turbulent water flow in a 3.8cm i.d. pipeflow, with some tests also involving injection of silicone oil drops. A water superficial velocity of 2.2m/s was involved. Droplets and bubbles were observed to produce the same characteristic breakage mechanisms. The authors found bubble breakup to occur only at a dimensionless radii of greater than 1.0 i.e. outside the core. No coalescence effects were involved since at most only 2 droplets/bubbles were in the pipeline at any one time. Characteristic bubble and droplet breakage times were also recorded. The authors stress the existence of active and neutral zones in terms of having the required turbulence to sustain particle breakup.

A large number of studies have been made concerning the droplet sizes produced in liquid/liquid systems involving stirred vessels or similar apparatus, Chen and Middleman (1967) and Sprow (1967) are typical examples, where the latter is particularly of interest in that coalescing conditions are involved. Meijis and Mitchell (1974) also published an interesting study involving oil/water mixtures in several mixing apparatus and the definition of a universal mixing parameter which is not restricted to non-coalescing conditions. Arai et al (1977) have investigated the breakup of high-viscosity liquid/liquid mixtures in stirred vessels. The Hinze theory was modified to take account of the high dispersed phase viscosity, although again non-coalescing conditions are necessary.

3.3.2 Dispersions and their Bulk Properties

A great deal of work has centred on investigating the physical properties of emulsions, particularly their viscosities. The prediction of emulsion viscosity also necessitates that emulsion inversion characteristics be determined. Einstein (1906) conducted a theoretical analysis of the viscosity of very dilute suspensions of rigid spheres and concluded that the viscosity could be expressed by:

$$\frac{\mu_e}{\mu_c} = (1 + 2.5\phi_d) \qquad (3.2)$$

where

μ_e = mixture (emulsion) viscosity

μ_c = continuous phase viscosity

ϕ_d = volume fraction of dispersed phase

Several modifications to the Einstein equation have been made, mainly resulting in power series expansions for the second term. Taylor (1934) allowed for the deformation of the dispersed particles and modified the above equation accordingly.

Richardson (1950) reported early work on the properties of flowing emulsions. Emulsions were prepared in a co-axial cylinder apparatus using water and benzene. Depending on which emulsifying agent was used, water-in-benzene and benzene-in-water emulsions were formed. Emulsion viscosity was seen to increase rapidly with increasing dispersed phase concentration and the following equation was developed:

$$\mu_e = \mu_c e^{k\phi_d} \qquad (3.3)$$

where k is a constant to be determined by experiment. The author comments that both average droplet size and droplet size distribution may influence the dispersion viscosity, although no account of this is included in the above equation. In a later publication, Richardson (1953) stated that emulsion apparent viscosity is inversely proportional to the mean droplet diameter for emulsions of the same concentration and size distribution.

Vermuelan et al (1955) measured interfacial area in a wide variety of agitated gas-liquid and liquid-liquid systems. Droplet and bubble diameters were extracted from interfacial area data for different impeller geometries and rotational speeds. The viscosities of the dispersions were inferred from measurement of the mixer power consumption. The following equation was proposed for correlation:

$$\mu_e = \left[\frac{\mu_c}{(1-\phi_d)}\right]\left[1 + \frac{1.5\,\phi_d \mu_d}{(\mu_c + \mu_d)}\right] \qquad (3.4)$$

A recent paper by Guilinger et al (1988) dealt with kerosene-water dispersions in mixing vessels, studying inversion phenomena and dispersion viscosity. The material from which the impeller was constructed was found to affect the dispersion inversion point, although this effect was seen to diminish as mixer size was increased. Power input data resulted in μ_e values which were compared to predictions from a number of existing equations. The authors concluded that the equation above gave the best fit to their data.

Whilst the above serves to illustrate the work available concerning liquid-liquid systems in general, it is valid at this point to discuss the systems of most relevance to the present study i.e., hydrocarbon liquids including crude oils.

One of the earliest investigations of crude oil-water emulsions was by Monson (1938). Californian oils of viscosities ranging from 68-253 cP at 100°F were emulsified with water using a hand-operated piston-type emulsifier. Tests were conducted at 100,130 and 185°F with water cuts up to 40%. The viscosity ratios, defined as the viscosity of the emulsion to that of the clean oil, were found to increase as water cut was increased at constant temperature: a maximum value of 4.1 at 40% water cut was recorded. The author stresses that the emulsions were checked diligently for stability, to ensure that no emulsion breaking was taking place.

Woelflin (1947) published a paper to reinforce to the oil industry the importance of being able to predict the viscosity of a water (brine)-in-crude oil emulsion. The author states that emulsion viscosity is dependent upon the relative amounts of oil and water and the size and distribution of the dispersed droplets. Viscosities were measured in a funnel-type viscometer and droplet sizes were measured using a centrifuge. Curves for emulsion-to-clean oil viscosity ratio are given against water content, with droplet sizes of tight, medium and loose used for classification. The given curves assume emulsion inversion in the range 60-85% brine, with maximum viscosity attained at the inversion point - see Fig 3.8. The effect of temperature was studied using a different viscometer and test oil. The author reported that the viscosity ratio appeared to be independent of temperature over the range 75-180°F, and for water contents up to 30%. Tipman and Hodgson (1956) also measured the viscosities of water-in-crude oil emulsions; in this case using rate-of-fall solids as the measurement method. A range of oils was involved, including a heavy oil which was also diluted with kerosene, and two grades of lubricating oil. No emulsifiers were added, although the emulsions were reportedly very stable, and Newtonian character was obtained over the range 70-180°F. Simple relations were derived for the emulsion/clean oil viscosity ratio. These relations involve dispersed phase concentration and continuous phase viscosity as input parameters: the relation is only valid over the range 0-30% (vol) of water.

Abdurashitov and Avanesyan (1964) also performed experiments on the properties of petroleum emulsions. The density of emulsions of 35, 47 and 68% water was linear over the temperature range 15-70°C. Water content and droplet size were found to influence the emulsion viscosity, which was reportedly higher than that of either or the sum of the clean constituents.

An interesting study by Strassner (1968) was primarily concerned with emulsion stability. Emulsions of water in oil were stabilised by chemicals, which form films on the interfaces, and by adjustment of pH, giving variable surface charge to the solutions. Very low amounts of commercial surfactants were found to drastically alter the interfacial phenomena, and emulsion type and stability were affected by pH. The author suggests that adjustment of pH could be utilised in the field to break emulsions or to prevent their formation, and suggest a mixture of laboratory and field tests to obtain the optimal formula for a particular application.

Simon and Poynter (1968) outlined a method which uses emulsions to improve production rates from viscous oil deposits. Based upon their laboratory tests, the authors comment that emulsion viscosity was essentially independent of the oil viscosity. The Richardson equation (Eq 3.3) was examined and the authors suggest that the constant k be taken as 7.0 below 74% oil fraction and 8.0 above this value. The effect of temperature on emulsion stability is also discussed.

Rose and Marsden (1970) performed a similar study with emulsions of heavy Prudhoe Bay crude oil in water. The flow properties of stabilised emulsions were examined at temperatures of 20°F and above using a 0.25-inch i.d., 134-inch long coiled glass tube as the flow resistance. Tests revealed the emulsion viscosity to be markedly lower than that of the crude alone and assuming laminar Newtonian flow, the value k = 4.08 in the Richardson equation fitted the data satisfactorily. McAuliffe (1973) also reported on the properties of oil-in-water emulsions. Emulsions were prepared using up to 80% (vol) oil in water, with caustic soda as the surfactant: clean oil viscosities ranged from 3600-1,000,000 cP. For 50% oil fraction, the emulsions had viscosities of at most 20 cP and, for 70-80% oil, the viscosities ranged from 300-10000 cP at 40°F. Droplet size and distributions were observed to depend on the type and quantity of the surfactant used, as well as the oil itself. The emulsions with greater than 60% oil exhibited non-Newtonian behaviour, which the author attributes to droplet-droplet interaction.

Chen (1974) investigated the stability of crude oil-in-water emulsions from the viewpoint of dealing with oil spills. Oil/water volume ratios of less than 0.1% were involved and shifts in droplet size distribution were attributed to gravity separation with no coalescence being observed over a 70-day ageing period.

Camy et al (1975) performed experiments on the physical properties of crude oil-in-water emulsions. Oil-in-water emulsions were found to exist as far as 60% oil concentration, before inversion occurred. The Richardson equation (Eq. 3.3) suggested a k value for oil-in-water emulsions which was close to that obtained by earlier workers. The value obtained for water-in-oil emulsions, however, differed appreciably from previous work and the authors suggest that this may be due to the fact that different surfactants were employed.

Mao and Marsden (1977) performed a further study of the stabilities and properties of oil-water emulsions. A closed loop cell with tight temperature control was used, with the emulsion flowing through a 0.2-inch i.d. tube. Apparent viscosities were obtained for oil concentrations spanning 10-90% and temperatures in the range 75-180°F. For oil-in-water emulsions, the ratio emulsion viscosity/water viscosity was found to be temperature-dependent. However, for water-in-oil emulsions, the ratio emulsion viscosity/oil viscosity was observed to be independent of temperature, in agreement with Woelflin (1947). The sensitivity of emulsion inversion to temperature, composition and shear stress was also examined. The shear stress at which inversion occurred was determined by the pumping rate which gave the highest pressure drop, and a simple equation is given.The results were found to be qualitatively compatible with those of previous workers.

Recently, Hartley and Bin Jaidid (1989) also performed tests on emulsions as part of a programme to identify potential fluid-handling problems in the North Sea Troll field. Emulsions were prepared with water contents 0, 20 and 40% and viscosities measured in a thermostatted rotating viscometer. The clean oil viscosity was approximately 15 cP at 20°C and the emulsion apparent viscosity ranged from 250-1150 cP depending on shear rate. However, the authors comment that they regard their artificially-made emulsions to be much tighter than those expected in the field, from previous experience, and present curves from a field correlation, although the source of this correlation is not given.

As mentioned previously, a considerable body of literature has been amassed concerning the behaviour of liquid-liquid and gas-liquid dispersions in agitated process vessels. However, the flow of liquid/liquid concentrated dispersions in pipelines has received a good deal less attention.

Cengel et al (1962) investigated the flow of a dispersion of a petroleum solvent in water. The test pipe consisted of a 7/8 inch i.d. pipe with section lengths of 10.5 ft (horizontal) and 9.5 ft (vertical). Viscosity was measured using glass capillary tubes installed at right angles to the flow: no measurements of droplet size were made. The dispersed phase volume concentration range was 5-50% and for all concentrations the vertical data indicated Newtonian behaviour. For the horizontal tests, Newtonian behaviour was noted for dispersed phase concentrations up to 20%, which the authors attribute mainly to phase-separation effects. Ward and Knudsen (1967), used the same apparatus with three different oils, ranging in viscosity from 1-260 cP at 62°F, which were dispersed in tap water. Assuming homogeneous flow and Newtonian behaviour, the pressure loss data for the light oils was in excellent agreement with the Einstein relation, (Eq. 3.2): dispersed phase concentrations of 1-47% were involved. The same success was not observed for the higher viscosity oils. Droplet size distributions were measured at a single point using a photographic method. The light oil dispersions were found to produce two peaks in the size distribution, whereas the heavy oil produced a single peak. The authors suggest that a higher dispersed phase viscosity results in a larger average droplet diameter, as the three systems had similar interfacial tensions. The authors also suggest that droplet breakup was of a mechanism whereby a small drop breaks off a much larger droplet, leaving the parent droplet largely unchanged.

Arirachakaran (1983) collected data on the flow of oil-water emulsions at the

University of Tulsa: this differed from the previous Tulsa studies in that here stabilised emulsions rather than temporary dispersions were involved. The test loop was 1-inch i.d. 20 ft long with pressure drop the key measurement. The test oils had viscosities of 176 and 1446 cP at 80°F. Oil-in-water and water-in-oil emulsions were prepared in a batch mixer and observations at a visual section in the loop suggested the flow to be homogenous. The pressure drop for the oil-in-water emulsions was found to be very close to that for water-only flow, given the same set of conditions. Pressure loss at the inversion point of the water-in-oil emulsions was found to be very close to that of the single-phase oil. The effect of temperature on pressure loss was found to be far more pronounced when oil was the continuous phase, which is at odds with the trends observed by several previous workers.

Wyslouzil et al (1987) presented an experimental study of pipe flow of crude oil-in-water emulsions. A Cold Lake crude was emulsified using a 0.1% (wt) sodium hydroxide solution and the resulting oil-in-water emulsion was pumped around a 3/4 inch i.d., 5m long test pipe. Tests were run where the emulsion was pumped in closed-circuit over prolonged periods at constant flow rate and temperature, until samples from the line indicated emulsion breakdown. Pressure drop data showed that friction factors were below the Blasius laminar line, which was attributed to either visco-elastic effects or the formation of a thin water layer at the pipe wall. In the conclusions the authors comment that, whilst simple viscometry tests had indicated the suitability of an emulsion pipeline, the tests conducted on the pipe loop had shown that the behaviour of such a fluid system in terms of stability was less easily predicted.

3.3.3 The Rheology of Liquid-Liquid Systems

The science of the deformation of flowing fluids takes on a particularly important aspect when dealing with dispersions and especially emulsions. The behaviour of a liquid-liquid dispersion is very dependent on the chemistry between the fluids and in many cases (not only the oil production industry) the situation arises where the dispersion behaves as a non-Newtonian fluid when the constituent liquids are themselves perfectly Newtonian in nature. This section is intended to demonstrate features of such oil-water systems pertinent to the current problem of co-production of reservoir oil and water fluids.

Lamb and Simpson (1963) reported details of one of the earliest operational emulsion pipelines. The pipeline is 20-inch i.d., 238km long and with a 30% (vol) water addition the emulsion viscosity decreased from 100,000 to 400 cP at a pipe shear rate of 10 reciprocal seconds. The crude's high pour-point and the crystallisation of wax particles meant that under some conditions the fluid was non-Newtonian, although this is not expanded upon.

Uzoigwe and Marsden (1970) reported on the rheology of oil-in-water emulsions.

Emulsions were prepared using a 1.3 cP oil dispersed in water to a maximum concentration of 70% (vol) and viscosities were measured using a series of capillary tubes. Rheological parameters were correlated using the Metzner-Reed (1955) approach and Newtonian behaviour was noted up to 50% oil. Emulsions of higher oil content were pseudoplastic, with the highest deviation from Newtonian having n = 0.88. The maximum viscosity of 70% oil was found to be around fifty times that of the clean oil.

Alvarado and Marsden (1976) investigated the flow of oil-in-water emulsions in capillary tubes and porous media. Newtonian behaviour was noted up to 40% oil, with pseudoplastic character observed for oil concentrations of 50% or higher, for the investigated range of shear rate. The transition from Newtonian to non-Newtonian was observed to have a dependence on the emulsifier concentration. The authors concluded that the rheological behaviour of the emulsions flowing through porous media was the same as that observed through the capillary tubes. Camy et al (1975) also noted that above 50% oil their oil-in-water emulsions exhibited non-Newtonian behaviour, with most being pseudoplastic and a few showing dilatant behaviour. Mao and Marsden (1977) found non-Newtonian behaviour above 50% oil concentration, with the emulsions being pseudoplastic in nature.

Zakin et al (1979) published experimental work on high concentration oil-in-water emulsions. The test loop was of 1-inch i.d. with a recirculation pump system. Oil concentrations were 50-75% and capillary-tube data indicated non-Newtonian character with a lowest n = 0.62, using the power-law model. The experimental pressure losses were consistently lower than those predicted by the Dodge-Metzner (1959) relation, by some 26% at most. The effectiveness of several drag-reducing additives was also tested, with several polymer additives proving effective, but only up to the point where they are degraded by mechanical shear.

Sifferman and Greenkorn (1981) presented an interesting account of drag reduction in fluid systems, one of which involved oil-water flow. A test loop of 27mm i.d. and 10m long was used, where the oil was a 200 cP mineral oil. A concentration of 70% oil was seen to give considerable reduction in pressure drop compared to the oil-only value, with the pressure drops being slightly above that of single-phase water. At oil concentration 75%, when dosed with a non-Newtonian polymer solution drag reduction was obtained which in some cases gave pressure drops lower than that expected for single-phase water.

Wirasinge (1980) published a paper outlining a method for the calculation of pressure losses for water-in-oil emulsions. He assumes that up to 10% water fraction Newtonian equations apply and for higher water cuts pseudoplastic flow is assumed. The power-law, Metzner-Reed (1955) approach is employed, but the author assumes that no viscometry data for k and n is available. It is assumed that field tests in laminar flow will yield pressure drop - flow rate curves, and equations are given to back-calculate k and n from field units. This then allows the designer to predict the upper value of throughput of emulsion through the pipeline which was used to generate the pressure drop - flow rate curve.

Flock and Steinborn (1982) studied the rheology of heavy crudes and their stabilised emulsions. Crudes of viscosities 40,000 and 50,000 cP, at 21°C and shear rate 1 reciprocal second, were used. Oil-in-water emulsions were found to be Newtonians up to 20% oil and pseudoplastics for higher oil contents. Water-in-oil emulsions were found to display pseudoplastic behaviour also. The application of the Richardson equation (Eq. 3.3) resulted in variable success, and the authors comment that shear rate and temperature were of great influence, although these parameters are ignored in Eq. (3.3).

Fruman and Briant (1983) performed a similar study on oils ranging in viscosity from 46,000 to 180,000 cP which were found to be Newtonian. Emulsions of oil-in-water of 60% (vol) oil were prepared, using variable amounts of salt and emulsifying agent. Addition of surfactant caused a massive reduction in interfacial tensions, and huge reductions in viscosities were also displayed by the emulsions. Droplet size measurement were made, and droplet diameter was found to be weakly dependent upon the method of emulsion preparation.

Martinez (1985) studied oil-water flow in pipes where the rheology was thoroughly investigated. A refined oil of 49 cP viscosity at 100°F and tap water were the test fluids. The test pipeline, effectively employed as a viscometer, was 1-inch i.d. and 20 ft long and set at the horizontal position. The experimental variables were temperature, water/oil ratio, mass flow rate and emulsion droplet size. The latter was inferred from the mixer speed used to prepare the emulsions and only comparisons between different mixer speeds were made; no droplet sizes were measured. For all flows a homogeneous flow patterns was observed. In analysing the pressure drop data, the power-law model is assumed to hold and parameters n^1 and k' are calculated using the Metzner-Reed approach. The greatest degree of non-Newtonian behaviour was pseudoplastic with $n' = 0.76$. Input water fraction was observed to influence pressure loss to a greater extent than either temperature or n. When water was the continuous phase, pressure drop was only slightly affected by temperature, and no dependence of inversion point on mixture velocity was obtained.

Pal et al (1986) investigated the rheology of high concentration oil-in-water emulsions. Mineral oil and tap water were emulsified using a non-ionic stabiliser, to a maximum attainable oil concentration of 84%. For shear rates in the range 2-200 sec^{-1}, the power law model represented the viscometry data satisfactorily, with the lowest value of $n' = 0.263$. Shear rates higher than 200 sec^{-1} were found to be described best by a Bingham-type model. Emulsion viscosity was found to be dependent on droplet size and distribution, which themselves were found to be dependent on emulsifier addition.

Pal and Rhodes (1989) presented a brief account of stable emulsion flow in pipelines. Oil-in-water emulsions were prepared where the base oil was a mineral oil of viscosity 2.4 cP at 25°C, and these were flowed through a series of steel lines the largest of which was 26.5 mm i.d. The workers found maximum droplet size to increase as the dispersed phase concentration increased, and that up to a dispersed phase concentration of about 55% (vol) the emulsions were Newtonian. Pressure losses were converted to friction factors assuming single-phase behaviour, and data agreed well with the theoretical laminar and turbulent flow (Blasius equation) relationships. At dispersed phase concentrations 45-55% (vol), non-Newtonian (pseudoplastic) behaviour resulted. Data agreement with the Metzner-Reed (laminar) and Dodge-Metzner (turbulent) non-Newtonian friction factor concepts was deemed satisfactory, although in the latter less data is available to furnish the comparison.

3.4 OIL/WATER/GAS PIPE FLOW

The study of the flow of three-phase gas/liquid/liquid mixtures has received very little attention from researchers, both in the field and in the laboratory. One possible reason for this is that the last 20 years or so has seen major efforts to improve understanding of two-phase gas-liquid pipe flow and, to a much lower extent, liquid-liquid flows. Hence the assumption that, for design purposes, the two liquid components form a "pseudo-liquid" phase with averaged liquid fluid properties. Whilst this may be appropriate for several liquid systems, it may not be true of oil-water mixtures in many instances, as has been discussed previously. Moreover, the simultaneous flow of oil/water/gas in a pipe has not, historically, been a common practice and so has largely been ignored by researchers/operators. The existing literature is now reviewed, although it is well to mention that no comprehensive studies have yet been published.

Sobocinski (1955) undertook the earliest recorded study of oil/water/gas pipeline flow. The experimental system consisted of a 10m long transparent horizontal pipe of 3-inch i.d. where pressure loss and in-situ liquid fraction were measured. Tap water and air were used at near-atmospheric conditions, and a gas-oil of 3.38 cSt viscosity at 100°F was the hydrocarbon. A wide range of fluid flow rates were involved, with the typical upper limit of water/oil (wt) being 4.0. The observed flow regimes for several oil/water ratios are described and summarised in flow patterns maps - see Fig 3.9. Flow regime information and hold-up data revealed that for many cases the increase of gas rate results in a fluid phase which is not strictly an emulsion, but rather a foam-emulsion caused by mixing with the air. Pressure drop data showed maximum losses to be obtained at a water/oil mass ratio of 4.0. Fig 3.10 is an example, and the increase in pressure drop was accompanied by a flow structure described as sluggish emulsion. All three-phase pressure drops exceeded the two-phase pressure drop data which was also collected. The flow patterns were mainly of wave-annular types, and classic slug flow was not encountered.

Correlation of the three-phase pressure loss data was made with existing empirical equations developed for two-phase flow. Some success was achieved, given certain adjustments of the density and viscosity terms and fitting of the empirical constants. However, the author concedes that this is an approach which is best avoided. Soon after this study, a paper by Sobocinski and Huntington (1958) summarised the main findings.

Shakirov (1969) collected oil-water-gas data from a Russian oilfield. The oil/gas flow was from producing wells and different water cuts were achieved by pumping in water from the industrial supply. Test loops of diameters 40, 50, 62 and 102mm were involved. Dry oil/gas tests gave two-phase pressure drops and the flow regimes varied between stratified, wavy and annular flow. Water cuts of 42 and 56.7% produced the three-phase data. A series of curves for emulsion viscosity (to a maximum 70% water cut) are presented for the temperature range 0-30°C. Empirical equations from earlier studies were utilised and, together with the presented graphs, result in a single equation for calculation of the coefficient of friction during oil/water/gas flow. The author claims this equation has been applied to pipe sizes 40-200mm in diameter.

Schlichting (1970) presented data on oil/water/gas flows and compared results of various prediction methods. The data was obtained from field lines where high viscosities were encountered and the gas/oil ratio was low (quoted as about 100 Sm^3/m^3 at most). The field data is from an 8 5/8-inch i.d. line and a combined line of 3 1/2, 41/2 and 51/4-inch i.d. in a different producing system.The larger of these lines had a length of 4760m, a water cut of 7.2% (vol) and producing gas/oil ratio 38.1 Sm^3/m^3. The dry oil viscosity was 225 cP. In this paper, the water cut is assumed as free water, where this excludes any water droplets finely dispersed in the oil phase. The second test line had free water cuts of around 10%, oil viscosity 15 cP and producing GOR or 83 Sm^3/m^3. The measured pressure drops were compared against a range of correlations, including Lockhart-Martinelli (1949). Wide scatter in the predictive accuracies was obtained, with some correlations resulting in inconsistent predictions.

The author notes that liquid viscosity has a large dominance over the success of the correlations. The liquid viscosities are calculated by a Richardson-type relation, depending upon the free water content and whether the emulsion is subcritical or supercritical, as defined by the author. A pressure loss calculation is employed which, essentially, employs a Lockhart-Martinelli-type approach. The predicted accuracy of the Schlichting correlation is seen to be very much better than that of the other methods tested. Finally, the author suggests applicability ranges for these correlations in terms of liquid viscosity, assuming no free water is present.

A paper by Bocharov et al (1972) presented further field-based experiments involving oil/water/gas flow. Pipe loops of 1, 2, 3, 4 and 6-inch i.d. of 50m long horizontal instrumented pipe were involved. The fluids used were crude oil and gas from various production wells and water from injection wells. A series of curves of pressure loss against water content and gas content are presented. Inversion from water-in-oil to oil-in-water emulsion occurred at about 77% water, with this location being unaffected by gas fraction. However, the pressure drop was seen to reduce progressively as the gas content was increased over the range 17-94%. The authors note that samples of the fluid at 77% water fraction, when examined in the laboratory, suggested non-Newtonian behaviour.

A short paper by Guzhov et al (1974) gave further information on the pipeline flow of water-cut crudes with associated gas. The authors comment that it is incorrect to take correlations from emulsion tests, as emulsions formed in pipe flows are often unstable. Field data, involving low-viscosity, light crudes and slightly-acidic and saline formation water, is presented for a narrow range of gas fractions. A plug-dispersion type of flow was observed in all cases. Curves are presented which show that gas content can affect the water cut at which a large increase in pressure drop can occur, and also the position of the inversion point itself. Increased gas content was observed to lower the maximum pressure drop attained, although the large increase in pressure loss is experienced at a lower water cut. The authors comment that these effects should only be observed in dispersion-type flows.

The work of Malinowsky (1975) mentioned previously also included a series of experiments investigating oil/water/gas flow. The gaseous phase was air, and 33 air/oil/water tests were run at near-atmospheric conditions.The majority of the tests involved slug flow, with several runs at the highest gas rates exhibiting a slug-annular/mist character. The flow resembled two-phase flow regimes, with the oil and water flowing as a homogeneous dispersion and a foaminess at the slug leading edge was observed as gas velocity was increased. The Beggs and Brill (1973) flow pattern map satisfactorily predicted the observed flow regimes. Pressure loss data was compared with the predictions of Beggs and Brill (1973) and Dukler et al (1964) and poor agreement was obtained. Large underpredictions of pressure drop was observed at water cuts of less 50-60% and overpredictions at higher water cuts, with the same trends observed for both correlations. The fluid properties input assumed a weighted oil/water average, and the author suggests that the treatment of the liquid viscosity term in this manner has affected the correlation accuracy.

Laflin and Oglesby (1976) used the same equipment as the previous author to investigate oil-water (reviewed previously) and oil/water/air flows. The three-phase data was taken at superficial liquid velocities of 2.65 and 4.7 ft/s, with water fractions between 0.2 and 0.8, resulting in a total of 79 tests. Three ranges of in-situ gas/liquid ratio were examined, with the emphasis on examining the effects of small amounts of gas on the effective fluid viscosity. The apparent liquid viscosity was back-calculated from the pressure drop data using the friction factors of Beggs and Brill (1973) and Dukler et al (1964). The introduction of gas shifted the inversion point from approximately 40% input water fraction to about 50% water. The authors attribute this to gas bubbles being occluded in the emulsion and preventing coalescence of the dispersed droplets, thereby requiring a higher concentration of the dispersed phase to achieve the necessary coalescence and subsequent inversion. The flow pattern observed for all tests was slug flow and this was predicted by both the Beggs and Brill (1973) and Mandhane et al (1974) flow regime maps. Finally, from analysis of the pressure drop data, the Dukler et al (1964) correlation was reported to give a consistently higher apparent liquid viscosity prediction than that from the Beggs and Brill (1973) correlation.

Fayed and Otten (1983) presented an account of field tests involving oil/water/gas flow. The data was collected from several 6-inch i.d. flow lines, of length about 7000-19000 ft. The crude oil involved had a viscosity of 8.3 cP at 100°F; the maximum water cut studied was 50%. The Mandhane et al (1974) flow pattern map was used for flow pattern identification. Predictions from the Dukler et al (1964) pressure loss correlation, with the Eaton et al (1967) holdup correlation, were compared to the field data, with the fluid properties being taken as a pseudo-phase of oil-water volume-weighted averages. Several of the calculations also incorporated the Woelflin (1947) (medium) emulsion viscosity correlation. For the tests involving slug and froth flow, predicted pressure drops were higher than those measured. Tests in the bubble flow regime resulted in underpredictions for the pressure drop. Interestingly, the results for an 18% water cut gave a lower friction loss than was obtained for a zero water fraction. The results in general indicated that the effect of water on pressure drop was insignificant and the use of the Woelflin correlation was inappropriate for this system. A possible reason cited for the lower-than-anticipated pressure losses for many of the tests is that separation has resulted in water forming a film at the pipe wall, with a resulting drag-reducing effect.

Gregory and Fogarasi (1985) reported extensive collative work involving field data from numerous multi-phase field lines and their comparison with the predictions from a large number of existing design methods. Of particular interest here are the limited comparisons for oil/water/gas data. Pressure loss results for pipelines ranging from 114 to 406mm diameter and water cuts from 32-46% showed that prediction accuracy was very poor, with all models and hybrid methods giving under-predictions. Liquid volume fraction was also observed to be predicted very poorly in general. Further data from gas-condensate-water systems gave comparisons with liquid holdup only and, once again, all methods were seen to underpredict the liquid hold-up by a considerable margin. In their conclusions, the authors comment that it is important to investigate both the qualitative and quantitative aspects of oil/water/gas flow.

Very recently, an experimental study of oil/water/gas flow was undertaken by Stapelberg et al (1991). This study is unique in that slug flow only is involved. Two plexiglass pipes of 24mm and 59mm were involved, and the oil was 31 cP viscosity: air and water were the other test fluids. Oil-water pressure losses were measured, and correlated using a modified Lockhart-Martinelli (1947) approach:

unfortunately the raw data are not presented.Systematic studies of slug flow (at low-to-moderate V_{SL}) of air/water, air/oil and air/oil/water were then made. It was found that the wavy-slug transition for 3-phase flow was between that for air/water and air/oil. Slug frequency was correlated with the method of Tronconi (1990), although the approach depended on whether the oil volume fraction was greater or less than 25%. Pressure drop data revealed the 3-phase losses to be less than that for air/water in many cases, which is attributed to the different slug flow characteristics. This study will be referred to in a later chapter.

Nuland et al (1991) performed measurements of in-situ fractions in oil/water/gas flow which produced a small amount of experimental data. A nuclear densitometer method was used to determine the phase fractions, the development of which is the main thrust of the paper, and used in an experimental system. This consisted of a 6m long test section of 32mm i.d., configured horizontally, where the test oil had a viscosity of 1.75cP at 20°C. Data was taken for a flow V_{SL} of 0.024 m/s, input oil/water ratio of 10 and V_{SG} up to 11 m/s and so corresponding to conditions expected in gas/condensate systems. Holdup data using quick-closing valves was also collected. For these conditions, three-phase liquid holdups were higher than the oil/gas equivalents, and the in-situ water fractions were also found to be higher than the input values. The authors comment that, from the gamma densitometer data, it appears that even if the oil-gas interface is smooth, the oil-water interface can show a much wavier, rougher nature.

3.5 CONCLUSIONS

The review of the literature available encompassing the given technical scope enables the following observations to be made:

i Most experimental oil-water studies have been confined to either well-separated or fully homogeneous flows.

ii The available theoretical modelling of oil-water flow is very sparse but suggests that improvement upon the few existing empirical methods is obtained.

iii Work concerning the stability of droplets in a liquid-liquid flow is numerous, but has been confined to very low dispersed phase concentrations.

iv Investigations of the properties of concentrated dispersions have been mainly restricted to systems where the dispersion has been chemically stabilised or where a rotating cylinder or capillary viscometer has been the flow apparatus.

v No accounts have been found of the behaviour of a dispersed bubble in a liquid-liquid dispersion carrier fluid.

vi Oil/water/gas studies are very few in number, those available typically involving narrow flow rate ranges and fragmented data sets.

vii Few convenient comparisons exist to enable examination of particular oil/water/gas systems with their equivalent gas-free liquid-liquid behaviour.

viii A systematic database does not exist for the development and testing of three-phase design methods.

CHAPTER 4

DESIGN OF THE EXPERIMENTAL FACILITY

4.1 OVERALL DESIGN PHILOSOPHY

At the outset of the design stage many considerations were involved regarding the capability of the facility given the constraints of project budget, timescales and technical objectives. The following were the main points which influenced the facility design:

i The facility would operate indoors and at low pressures.

ii A significant flow visualisation capability was necessary.

iii Attention would be restricted to horizontal and near-horizontal pipes.

iv Standard instrumentation would be utilised throughout.

The extent to which each of the considerations above had a bearing on the final design is detailed in later sections.

4.2 CHOICE OF SYSTEM TEST FLUIDS

The choice of which fluids to use in the test system was a factor of greater-than-anticipated importance. There was an obvious health and safety requirement due mainly to the indoor siting of the test loops. The use of high flashpoint hydrocarbon was deemed vital, especially due to the fact that air was intended as the gas phase. It was also considered important that the oil exhibited a low aromatics content and that any contact with skin would be non-hazardous.

Technical considerations were made as to the suitability of certain oils in terms of fluid physical properties and any possible problems regarding oil/water dispersion formation and oil/gas foaming tendencies. The problem of how the oil and water will combine/separate was central to this subject, and instances of severe difficulties encountered in oil-water pipeline experiments have arisen in the past. The oils eventually chosen for the tests are detailed in the experimental chapters where appropriate. A low viscosity kerosene was chosen as the first test oil. This oil was colourless and it was decided that some form of marking of the oil (or the water) would be necessary to help visual observations of the oil/water/gas mixture. The oil is dyed with a commercial red tax powder to a concentration of approximately 10ppm (wt). This avoided the inconvenience of marking the water.

The choice of tap water as opposed to distilled water meant the former was used because of practical considerations. However, prolonged contact between oil/water/air necessitated that some form of action be taken to control any problems regarding the growth of bacteria in the system. Dosage of water with a low concentration of chlorine was considered difficult to control. A commercially available microbiocide (Oil-Aid-82) was added to the tap water to concentration 33ppm (wt). This chemical is active in both the water and hydrocarbon phases and, with regular monitoring of samples by a specialist laboratory, it was intended to dose the water with the chemical to control any bacteria problems.

Air was used as the gas phase principally due to the expense of using nitrogen in the rig. Nitrogen is desirable from the cleanliness point of view, but a once-through blowdown system or a buffer vessel plus recirculatory blower proved too expensive.

4.3 DETAILED FACILITY DESIGN

4.3.1 System Design Overview

Fig 4.1 illustrates the flow path for the fluids in the system. Oil and water are stored in separate storage tanks and pumped separately, through metering sections, to the liquid/liquid mixer. The gas is delivered to the gas/liquids mixer which is sited at the same elevation and close to the test loop inlet. After passing through the test loop it enters a 3-phase primary separator vessel. Gas is vented to atmosphere through a scrubber vessel. Each of the bulk-separated liquid streams then pass through a coalescer clean-up vessel and back to their respective storage tanks. The following sections describe the design of the key facility components.

4.3.2 Prime Movers and Metering Systems

The oil and water are stored in 2200 imp.gal. steel, well-insulated tanks, the oil tank being sited outdoors as shown in Fig 4.2. The oil tank is fitted with two 6kw thermostatically controlled immersion heaters; the water tank has a single 18kw heater. These heaters were intended to maintain the necessary temperature in the oil tank during cold periods outdoors, and for any emulsion-breaking which could be required in exceptional circumstances. The oil is drawn by a MONO positive displacement helical-rotor type pump with capacity $48m^3/hr$ at 3 bar head for all cases. The pump has a 4:1 speed control operated by means of a pulley arrangement. The oil suction valve is a pneumatically-activated shut-off safety valve which can be closed from a number of emergency switches indoors and outdoors. The pump discharges through a 4-inch i.d. heavily-insulated steel pipe with a valved 2-inch bypass back to the storage tank. This branch enters the tank through a simple jet-pump mixer. The bypass effects two chief purposes - to turn down the flow and keep a constant temperature in the tank by means of agitation.

The water pump, Fig 4.3 is of very similar design to the oil pump. The specified performance is the same as that of the oil pump, and again there is a valved 2-inch bypass back to the storage tank.

Air is supplied to the test loop by a screw compressor which discharges through a dryer/filter unit to a constant-pressure buffer vessel - see Fig 4.4. A valve/regulator combination is then used to tap air at the required rate and delivery pressure to the gas/liquids mixer.

The oil is metered by means of 3 parallel turbine flowmeters which provide the required turndown capability, only one flowmeter being used at any particular time. These meters have been supplied with calibrations for water flow (see Appendix B), although the manufacturers have advised that the use of the low viscosity (less than 20cP) oils should have a negligible effect on the meter factor. The water stream enters the water metering section which is below the oil metering runs - see Figs 4.5 and 4.6, where two Krohne electromagnetic flowmeters are used for water flow rate measurement. The gas metering system, shown in Fig 4.7 consists of 3 turbine flowmeters which have been calibrated using air flow. All initial calibration and subsequent re-calibration certificates are included in Appendix B.

The design of fluid mixers is a matter of some conjecture between researchers, although simple tee-pieces are commonly employed. In the liquid/liquid mixer, the oil enters horizontally (see Figs 4.5, 4.8) over the input water stream just downstream of the liquids metering section. This mixer is of steel construction.

The gas/liquids mixer is shown in Figs 4.9 and 4.10. This allows the gas to enter a void around the oil/water 'core' via the manual gas input manifold - see Fig 4.11. The unit has been machined from perspex with sealing of components by O-rings, which is squeezed between flanges using a set of tie-rods.

4.3.3 Separation System Design

It was considered that the design of the separation system was a vital exercise to maintain the controllability and quality of test data by tight control of fluids cleanliness. It was necessary to consider long term objectives for the facility given that major investment now could preclude design modifications in the future. Although near-atmospheric operating pressures were involved in this study, major capital items were designed around moderate operating pressures from a long-term standpoint. The separation and metering systems were designed to BS5500 Cat 2/ASA 300 codes representing maximum working pressures of 25 barg.

The first-stage 3-phase separator (Burgess Manning Ltd) is made of carbon steel with an epoxy internal coating and is 1.5m in diameter and 5m long. The vessel is equipped with a pressure relief valve, pressure controller, level controllers and sight glasses. The inlet element is an angled plate, and the gas flows out through a vane-pack and a 6-inch manual gate valve to atmosphere outside the building. The float type level controllers, one on each side of the weir, operate 4-inch control valves at the oil and water outlets, see Fig 4.5. The pressure controller operates a pressure control valve sited on a dog-leg around the main gas outlet - see Fig 4.12. All control valves are fed with air from the works compressor. The 3-phase separator was assumed to provide only bulk separation of the oil and water. The bulk-separated streams are processed by cartridge-type filter-coalescer vessels, 36-inch diameter, 3.5m high and of stainless steel construction. The cartridges are of fabric and woven to ensure 5 micron particles (minimum) should be removed. Each vessel is fitted with pressure relief valve, gas vent, sight glasses, and top and bottom pressure gauges to help indicate when cartridges require replacement.

The oil from the coalescer vessel flows to the oil tank which is some 30m remote and involves considerable elevation change. For this reason the primary separator operates at a small back-pressure of 0.6-0.7 barg to provide continuous facility operation. Collected oil (water) from the top (bottom) of each vessel are periodically passed through the other vessel before discharge to the respective storage tank.

4.3.4 Test Loop and Supporting Structure

As mentioned, it was considered valuable to have a good visual capability on the test pipes. The test loops were designed predominantly of glass section and components, with the test loops has meant operating pressures will be restricted to 3 barg.

The test loops are mounted on a pipe-bridge structure which is 4 x 7m long sections of a triangular space frame similar to a radio mast - see Figs 4.13 and 4.14. Using this structure it was possible to site a considerable amount of equipment on a single unit, and is also convenient for maintenance of the test pipes. The frame is centrally supported on a split bearing of capacity 5 tons and locked onto supporting towers at each end. With this arrangement, inclinations of $\pm 2°$ from the horizontal are possible. The frame and supporting structure are sited on a sealed concrete plinth which included a slope and sump to provide effective drainage of any spillages. The frame is electrically grounded at various positions along its' length by buried copper conductor rods. The test loops sit on aluminium supports along the frame which are secured using U-bolts.

The test loops have a length of 56m which provides 1100 diameters (2-inch) and 550 pipe diameters (4-inch). Instrumentation is sited on spacer pieces which are made of a high molecular weight polymer - see Fig 4.15.

Pigging facilities were installed to allow foam pigs through the test loops. The pig launcher is hung off the inlet tower and the pig receiver and bypass arrangement, Fig 4.15, is located between the loop outlet and primary separator. This is also the location where the test loop is isolated from the separator for static pressure testing of transducers on the loop.

For safety reasons, a graphite bursting disc has been placed on a tee-piece on the loop which was intended to protect the glass pipe from any overpressures. The 2-inch disc is rated to 2.5 barg. The entire pipe bridge structure is surrounded by a fence and transparent 'Lexan' safety shield.

4.4 FACILITY INSTRUMENTATION AND DATA ACQUISITION

4.4.1 Process Plant Instrumentation

Several sensors are placed on the process system to enable the operator to check operating conditions and help flag any mishaps. Six LED liquid level sensors are placed on each of the storage tanks which give visual displays both next to the tank and in the control cabin. There are also a number of thermocouples located as follows:

- i Oil coalescer
- ii Water coalescer
- iii Oil metering line
- iv Water metering line
- v Gas metering line
- vi Oil storage tank

These instruments had visual displays in the control cabin and are logged manually. A Druck PDCR10 3.5 bar dp pressure transducer is placed at the gas metering section to enable correction of in-situ gas superficial velocity V_{SG} in the test loop. Bourdon-type pressure gauges have also been placed on the primary separator, coalescers, oil and water metering lines and one on the test loop principally for start-up purposes.

4.4.2 Test Loop Data Acquisition

4.4.2.1 Overall Objectives

The primary objectives of the data collection exercise were measurement of pressure drop and liquids holdup. Measurement of in-situ fractions in a flowing three-phase mixture is a topic of great interest currently, but it was decided that a reliable, economic and accurate instrument was not available for this work. Observation of flow regime was also viewed as important.

Of secondary interest was the measurement of slug flow characteristics, namely slug frequency, velocity, length and holdup information.

4.4.2.2 Pressure Measurement

Test loop pressure is measured using Druck PDCR10 pressure transducers located on the loop as shown in Fig 4.16. The transducers are rated at 1 bar dp and are wet/dry: the dry side was either vented to atmosphere or, for most tests, backed up by nitrogen from a storage bottle and regulator. This permitted use of the instruments at local pressures of greater than 1 barg. The nitrogen back-pressure line contains a similar transducer of 2 bar dp. All pressure transducers read to signal conditioners in the control cabin for conversion to computer inputs.

4.4.2.3 In-situ Liquids Fraction Measurement

Measurement of three-phase holdup is, as mentioned, attracting much research and development attention. For this study, holdup measurement is by means of quick-closing valves with a bypass for subsequent analysis. The selection of a suitable valve meant that the following requirements had to be met:

- i valves to be full-bore
- ii valves to have light weight
- iii closure time 0.25s or better

It was found that no 3-way valves were suitable on consideration of i and ii above. The chosen set-up was by coupling 2-way ball valves as shown in Figs 4.16 and 4.17. The valves are mounted via screwed steel flanges on units between the glass components, and the bypass line is of galvanised steel and restrained to the pipe bridge structure. The valves are placed just before the bend in order that when the pipe is inclined, measurement upstream of the bend will mean that in inclined flow the measurement will be unaffected by the change in inclination at the bend. The capture length for average holdup is 10m which is 200 diameters of the 50mm pipe. A capture length of 1m for slug body holdup is also provided. The valves are pneumatically actuated by air at 100psi from a clean mains supply and solenoids powered by mains electricity. It was desirable to be able to delay the switching of valves by very small amounts to help minimise any pressure-kick problems. A programmable sequential controller was built to achieve this where delays are set by the input program with a 10ms step being the best increment available. Mode A refers to average holdup measurement, B to slug body holdup measurement. The trapped fluids are drained by a vacuum sampler which consists of calibrated perspex settling pots connected to an ejector - see Fig 4.18. The oil and water then settle out and the volumes of each recorded manually. The fluids are then transferred back into the system at the pigging bypass using a small pump.

The measurement of holdup using this method is not wholly satisfactory when compared to continuous monitoring if the flow is time-varying or locally variable. The collected steady-state data is then only subject to inaccuracies if infrequent slug flow is occurring in the pipe. This is discussed in a later chapter.

4.4.2.4 Temperature Measurement

Temperature is measured at 5 locations on the test loop as shown in Fig 4.16. Type K thermocouples are connected in the top of instrumentation spacers and signal readout is in the control cabin and values are recorded manually. The thermocouples are rated to a maximum temperature of 100°C.

4.4.2.5 Measurement of Slug Flow Parameters

The measurement of slug length, velocity and frequency in a small-bore pipe permits few economic and simple options. The use of liquid level sensors (LEDs), as used on the storage tanks, has proved very successful for measuring slug parameters in large diameter pipelines provided due care is taken to the manner in which the sensors are mounted. The probes have an extremely fast response and operate on the principle of total internal reflection of a light ray focussed on the sensor head which is either reflected or refracted depending upon which phase is in contact with the head. The digital responses can then be interpreted by in-house software and time histories given and slug characteristics calculated. Six of these probes were mounted on the 2-inch pipe in two banks of 3 as shown in Fig 4.16. It was not clear how these would perform in the small-diameter pipe. The two sets of 3 also limited the collection of any data to only 2 locations in the test loop. The signals are displayed visually in the control cabin and conditioned for entry to computer.

4.4.2.6 Computerised Data Acquisition

Initially a PDP 11/23 computer was used for data acquisition, with data storage on a hard disk. The sampling rates for the pressure transducer flow meter and LED probes were 50Hz in all cases. Raw data was copied to magnetic tape for subsequent reduction via a VAX 11/750 mainframe, and final data recording was made manually in a controlled file, with raw data filed in a magnetic tape library.

Subsequently, the above system was replaced by a more convenient and efficient data acquisition package. A Tulip-AT Compact 2 minicomputer with 40mB hard disk was purchased for the remaining tests, and is depicted in Fig 4.19. The same sampling frequencies were used as for the earlier system, but in this case, it was possible to perform on-line data analysis. Following data acquisition, ASYST software was used to perform data interrogation, tabulated output of flow conditions and graphical output . This method then greatly reduced the time required to analyse the data and provided the added benefit of output data while the facility was still operating. Each test day data was transferred from the hard disk to floppy diskettes and stored in a library for future inspection and further analysis if required.

4.5 COMMISSIONING WORK

4.5.1 Start-Up and Operation Trials

Following assembly of all rig components the 2-inch glass test loop was tested to 3 barg to check for any leaks, before the loop bursting disc was installed. There then followed a period of air/water runs to check the performance of flowmeters, separator control valves and to charge the oily-water coalescer vessel with fresh water.

A similar procedure was followed with single-phase oil. Oil/water tests were then run to establish operating procedures for the separation system. It was found that, as expected, a minimum of 0.6 barg separator pressure was required to drive the fluids (particularly the oil) through the coalescers to the storage tanks.

A series of oil/water/air slug flow runs were then undertaken chiefly to test system stability and operability. At this point, separator pressure control was by means of manual adjustment of the gas outlet valve. These tests indicated the requirement for automatic separator pressure control, which was subsequently installed.

Sampling of the oil and water tank return streams was made to assess the performance of the coalescer clean-up vessels. The water return indicated an oil concentration of less than 30ppm using a Wilks-Miran infra-red spectrometer. This value was within the specified performance of the coalescer. Oil samples indicated no visible water layer, although no detailed measurement was possible.

The operation of the quick-closing pneumatic valves on the test loop was checked with water-only flow (i.e. worst case). Measurement of average holdup indicated that the control inputs were satisfactory. However, slug holdup measurement caused the bursting disc just upstream of the first pneumatic valve tee to burst from the pressure kick which resulted. The program for Mode B was consequently modified to avoid this problem and regular switching then produced no such undesirable effects.

The construction of the test loop supporting frame had meant that, to maximise the loop length within the available space, a return bend on the loop was unavoidable. During commissioning (and during all subsequent experiments), observations were made of the flow behaviour around the bend. In low-velocity slug flows it is possible that the slug slows down slightly as it encounters the bend. However, in no cases was the holdup measurement observed to be affected by bend effects. Also, it is considered that the location of the pressure drop measurement downstream is sufficiently far from the bend to be unaffected by any influence from the bend. In high-velocity conditions the bend will be expected to provide additional mixing to the flow, but in no case was an oil-water separated flow changed to a mixed flow simply by flow around the bend. In very high energy flows, the bend could contribute to a change in droplet size distribution, but the existance or extent of such an effect could not be quantified.

4.5.2 Instrumentation Checks

A number of single-phase oil and water runs were made to measure pressure drop in the pipe and compare with the prediction from the Moody (1944) chart. Measured pressure losses were considerably higher than the Moody smooth-pipe calculation, and much effort was expelled in the checking of instrumentation and associated equipment, particularly the pressure transducers. It was concluded that the instruments were performing satisfactorily and attention was then focussed on the way in which the transducers were mounted and the glass pipe. It appeared that the large number of joints were possibly contributing to a higher pressure drop, although it is fair to say that it was a contribution to what was, in real terms, a very small pressure loss in the first place. Also, the effect of pipe roughness is sometimes regarded as much less influential in a gas/liquid flow than in single-phase flows. Nonetheless, it was desirable to improve the measurement. A 6m section on the return leg of the 2-inch loop was replaced with a steel section with pressure tappings close to each end. The resulting single-phase losses were found to be within 5% of the equivalent smooth-pipe value. This was considered acceptable and the steel section was retained.

During the aforementioned slug flow tests data was collected by the loop LED liquid level sensors to examine their performance in the small-diameter pipe. Data was sampled at 50Hz frequency - this was considered the minimum satisfactory sampling rate - and time-histories of the LED data were inspected. The probes appeared to switch satisfactorily on contact with the slug front but it was the slug tail which indicated potential problems. The probes had been sited at approximately the 2 o'clock position as it was thought this would minimise problems with liquid droplets settling on the tip. The probe tip was about 1/2-inch inside the pipe bore: this was a compromise resulting from blockage and slug film height/wave production considerations. Results generally indicated inconsistency in the performance of the sensors, and such inconsistencies would be impossible to eradicate sufficiently for automatic data production from the slug characteristics FORTRAN program. Also, the tests were run at the lowest anticipated V_{SG} of around 1m/s. This should produce the easiest slugs for the LEDs to read; they are relatively slow and have a very low void fraction. Problems would multiply as V_{SG} was increased so unfortunately the use of the probes had to be suspended. As mentioned, slug data was a secondary objective of the study. However, it was decided that any collection of slug characteristics data would be made using a video technique, the nature of which will be described later.

4.6 DISCUSSION OF EXPERIMENTAL ERRORS

4.6.1 Flow Rate Measurement

The details of the instruments used to measure oil,water and gas volumetric flow rates have already been discussed. When considering the degree of accuracy and repeatability which can be ascribed to the raw experimental data, several points are of importance :

- The accuracy and repeatability of the meters
- The sensitivity of the signal conditioning
- The accuracy of manual monitoring (water only)
- Correction for pressure changes (gas only)

Considering the first point above, as no on-line meter-prover loop was available on the facility, the fluids employed in the experiments were the same as or similar to those for which manufacturer calibrations were made. Meter accuracy of better than 1% measurement error is expected, and examination of the meter factors before and after each calibration check showed very little difference. The sampling frequency of 50 Hz is considered adequate for the moderate mixture velocities which were encountered. A careful watch was made of the water flow rate for the duration of each run, and experience indicates that, due to the fact that the LCD outputs were normally very stable and the high robustness of this type of meter, the confidence in the measurement is high. It should also be underlined that due to the fact that the metering systems were designed for a very large flow rate span, meaning a number of meters were required for each fluid, each meter was operating comfortably within its acceptable flow rate range. The final point above indicates the reliance on the accuracy of the pressure measurement, discussed below. It would be difficult to accurately quote the experimental errors, but it is considered that a deviation of approximately 2% of measurement, for each phase flow rate, is conservative.

4.6.2 Liquids Holdup Measurement

At the instigation of the study no method had been proven as practicable for three-phase holdup determination. The use of the quick-closing valve bypass method does mean the following deserves mention:

- The representativeness of the sample
- The number and size of each sample
- The effectiveness of the fluids evacuation system

The former points above are particularly important when intermittent flow is the dominant flow pattern. Three 10m samples were taken in each test, where the flow was allowed to stabilise between each measurement. This is equivalent to trapping over 50% of the pipeline length. The reader is referred to 6.4 where measurements were made to obtain an estimate for the suitability of the method in intermittent flows. The final point above is to acknowledge that for the horizontal pipe it was impossible to completely drain the fluids using the suction method. For low holdup tests of 20% liquid or less, the efficiency is considered to be 90-95%. At the higher holdup conditions it is estimated that approximately 98% of the in-situ liquids were removed. For the inclined-pipe tests, it is estimated that liquid removal efficiency exceeds 98% due to the effect of gravity on drainage.

4.6.3 Pressure Measurement

From the outset a great deal of time and care was taken to provide a satisfactory pressure measurement. The accuracy of the measurement relies on:

- The accuracy of the instruments
- The sensitivity of the data sampling system

It is the first point above which is considered to be of key importance here. The instrument outputs were checked at least twice during each morning or afternoon session. Additionally, the instruments were checked against a traceable deadweight tester very frequently. The information collected from these activities indicated that a reasonable estimate for the accuracy of the pressure measurement is +/- 0.2 mB/m.

4.6.4 Slug Characteristics Measurement

The accuracy of the slug velocity, frequency and length data was governed by the capability of the video editing suite. The machine enabled resolution to 25 frames/second, and results in the following ranges in the error produced:

- Slug front velocity 4% - 12%
- Slug length 2% - 12%
- Slug frequency < 0.1%

The larger errors will obviously be obtained where high mixture velocities and short slugs are involved. Due to the averaging of 100 slugs which has been made, the error in the average values will then lie between the extremes quoted above.

CHAPTER 5

OIL-WATER FLOW IN A HORIZONTAL PIPE

5.1 INTRODUCTION

As noted in Chapter 3, studies relating to the pipe flow of a liquid-liquid mixture are considerably less numerous than those concerning gas/liquid pipe flow. In this study the collection of oil-water flow information was viewed as providing data which could be used to compare with some of the existing liquid-liquid data, but principally the exercise was intended as an important precursor to the oil/water/gas tests. This would enable one to investigate at a later stage, how the addition of a gas phase to the oil-water flow affects the behaviour of the oil-water mixture.

5.2 TEST OBJECTIVES

The technical objectives for the oil-water tests were as follows:

i Examination of oil-water flow regime

ii Measurement of in-situ oil/water ratios

iii Measurement of pipeline pressure drop.

The first point above involves the liquid phases distribution and, importantly, the superficial velocities necessary to establish a fully homogeneous oil-water mixture. The use of 2 oils of different viscosity provides the possibility of examining the effect of oil viscosity on mixture homogeneity, but more importantly it provides a benchmark of oil-water mixing prior to gas addition in the later 3-phase tests. The latter subjects above are necessary to investigate if any slippage was taking place between the oil and water and to examine the pressure loss characteristics of the liquid-liquid combination in a number of flow regimes.

5.3 EXPERIMENTAL PROCEDURE

Instrumentation and signal conditioning units were switched on early each test day and given at least 30 minutes to stabilise before any start-up procedures were initiated. For the test involving the lowest liquid throughputs the test loop pressure transducers were "zeroed" at atmospheric pressure and single-phase oil was then pumped into the loop. With single-phase oil in the system, water was then added to give the required oil/water ratio and total liquid superficial velocity V_{SL}. When water was observed to have progressed through the test loop and steady conditions achieved, which in some cases took a considerable period, sketches were made of the flow pattern at the inlet and outlet of the pipe. Pressure and flow rate data was collected by computer for a period of 3 minutes, and following this data acquisition the pneumatic fast-acting valves were actuated and trapped liquids drained to the measurement vessels.

The valves were then switched back to allow liquid back through the holdup section. Water input was then stopped and single-phase oil was swept through the loop to effectively "zero" the system water content before proceeding to the next set of oil and water flow rates. When sufficient volume of liquid had filled the separator to a high level fluid input was halted and the vessel isolated from the test pipe using the valves at the pigging bypass section. The separator was then charged with air from the works ring-main to a pressure of 0.6 barg and the oil and water passed through the coalescers to the storage tanks. It was not possible to perform the majority of the tests using the above batching procedure simply because the primary separator was filled with liquid so quickly requiring very frequent batching which was very time consuming. For such tests, the loop and separator were pressurised to 0.6 barg and the test pipe transducers set at this pressure. The procedure then followed was very similar to that above.

At the end of each testing period all liquid was removed from the test loop using air from the facility compressor. It was then possible to check the pressure transducers at static conditions to investigate any drift or problems with the instruments.

5.4 DATA PRESENTATION

5.4.1 Oil No. 1 Data

The oil which was first used for testing was a de-aromaticised kerosene, EXXSOL D80. The considerations and selection process undertaken in choosing this fluid are expanded upon in Chapter 4 and Appendix A. The density-temperature relationship of the oil is depicted in Fig 5.1. This was determined using hydrometers in a thermostatically-controlled water bath where the oil sample was heated to a maximum temperature of 40°C.

The viscosity-temperature curve is shown in Fig 5.2, where an Ubbelohde (U-tube) capillary viscometer was used in the same temperature-controlled apparatus as above. These measurements were made where the oil had been marked by the red tax powder and saturated with microbiocide-dosed water as described in Chapter 4. This was considered necessary to preserve consistency and to make sure any effects of the dosed chemicals were being taken into account. The oil-air surface tension was measured at 20°C as 28.5 mN/m (dynes/cm) using the drop-volume method. Oil-water interfacial tension was measured at 20°C as 32.5 mN/m using the same equipment. It was not possible to obtain accurate interfacial tensions at other temperatures using the available apparatus. It is expected that the variation in both oil-air and oil-water interfacial tensions over the anticipated running temperatures (ambient) would be very small - see Backes et al (1990) and Reid et al (1977) . The importance of liquid viscosity and surface and interfacial tensions to many 2-phase and 3-phase phenomena is discussed in Chapter 8.

5.4.1.1 Experimental Variables

The choice of experimental variable and their range and manner of investigation was a very important exercise. Since in this exercise facility constraints meant that only one pipe diameter and inclination and near-atmospheric conditions would be involved, the key variables were oil and water flow rates. It was doubly important at this stage to ensure an acceptable range in flow rates since the oil-water data points would form the basis for the oil/water/gas tests.

When we define:

$$V_{SO} = \text{oil superficial velocity} = \frac{Q_O}{A_p}$$

$$V_{SW} = \text{water superficial velocity} = \frac{Q_W}{A_p}$$

$$\text{then } V_{SL} = \text{liquid superficial velocity} = V_{SO} + V_{SW}$$

$$\lambda_W = \text{input water cut} = \frac{V_{SW}}{V_{SO} + V_{SW}}$$

$$(WC)_a = \text{in-situ water cut} = \frac{Q_W}{V_W \times A_p}$$

where Q_o = oil volumetric flow rate
Q_W = water volumetric flow rate
A_p = pipe cross-sectional area

The initial argument was put by the author that an investigation of V_{SL} - λ_W would provide most information as to the influence of water cut in the total liquid flow. Note that λ_W need not be equal to the in-situ water cut $(WC)_a$, even for a liquid-liquid system in a horizontal pipe. The research sponsors, however, decided that a matrix of V_{SO} - λ_W experiments should be performed. As shown in Table 5-2, V_{SO} varied from 0.1-1.0 m/s, λ_W over 0-70% (vol), and V_{SW} up to 1.2 m/s. The upper limit for V_{SL} was 2m/s which was the anticipated maximum V_{SL} for the oil/water/gas tests (based on pumping capability). Therefore, running the tests meant keeping the value of V_{SO} fixed, adding water to the flow to a maximum amount of 70% (vol), and then setting the new V_{SO} value with single-phase oil before recommencing water addition.
The test coding employed in Table 5-2 was as follows:

O W K 0 1 A

Code	A	B	C	D	E	F
λ_W(-)	0	0.2	0.4	0.5	0.6	0.7

Code	1	2	3	4	5	6
V_{SO} (m/s)	0.1	0.15	0.3	0.5	0.8	1.0

Pipe Inclination 0°
Oil No1 (Kerosene)

5.4.1.2 Flow Regime Data

The nature of the flow regimes are displayed in Fig 5.3 . The flow pattern classifications were made as follows:

A : Oil/water stratified flow where the oil and water are well separated and no mixing was observed at the interface and in either phase.

B : Similar to flow A but where some mixing was apparent at the interface in the form of occasional large droplets.

C : Stratified flow where a thick continuous layer of large droplets exists at the interface but where the bulk oil and water phases are clean.

D : Similar to flow pattern C where the oil phase is more turbulent in nature with smaller, chaotic droplets and some mixing of small oil droplets in the water phase is evident.

E : Semi-homogeneous flow where a layer of dilute oil-in-water emulsion at the bottom of the pipe resulted in a concentration gradient.

F : Fully homogeneous oil/water flow.

The flow regimes are displayed on maps in terms of V_{SO} - V_{SW} coordinates, Fig 5.4, and V_{SL} - λ_W coordinates on Fig 5.5. This enables convenient comparison with existing oil-water flow pattern maps. All information is also contained in Table 5-2.

5.4.1.3 Oil/water Phase Fraction Data

Measured in-situ water cuts $(WC)_a$ are plotted against the no-slip values, λ_w, in Figs 5.6 to 5.10. The numerical data is given in Table 5-2.

5.4.1.4 Oil/water Pressure Drop Data

As noted in 5.3, pipeline pressure loss was measured in all tests. However, review of the data suggested that losses through the glass/instrumentation spacers were slightly above those expected from the single-phase Moody (1944) smooth pipe relation and it was decided that a new measurement approach was required; this is discussed later. Therefore the pressure drop data from these tests is considered unreliable and will not be presented.

5.4.2 Oil No 2 Data

Oil/water tests were also conducted using a different test oil. Following the testing with the kerosene, another oil was sought mainly to investigate if any effect of oil viscosity was apparent, particularly in terms of pressure drop. Although it was the intention to increase the oil viscosity, an attempt was made to keep the oil density and surface/interfacial tensions as close as possible to the first test oil. It must also be mentioned that, whilst the use of an oil which had significatly different mixing properties with the water would have produced a useful fluid-property departure, practical considerations meant that the use of an oil which formed relatively stable dispersions with the water would not be possible. This resulted in a significant selection period involving the testing of many oil samples for fluid physical properties and foaming/dispersion characteristics. The background to these tests, the construction of the small scale test rig and the test results are contained in detail in Appendix A. The oil finally selected for testing was a light lubricating oil, BP Oil 7269, where the density-temperature and viscosity-temperature relations were as depicted in Figs 5.11 and 5.12 respectively, where the curves for Oil No 1 are included for comparison. The measurements were made using the same apparatus as for Oil No 1. The surface tension at 21°C was 27mN/m and the interfacial tension over water was 40.9mN/m at the same temperature. A full comparison of oil physical properties is given on Table 5-1.

5.4.2.1 Experimental Variables

The variables examined using the new oil were altered from Oil No 1 tests principally due to a proposed change of philosophy for the Oil No 2 three-phase tests. In these tests it was proposed to have set values of total liquid superficial velocity V_{SL} and vary the no-slip water fraction λ_W over the range 0-70% (vol) as previously. This suggestion had already been made to the sponsors regarding the Oil No 1 tests but rejected in favour of the approach outlined in 5.4.1.1. However, for the new tests the sponsors agreed that structuring the tests in this manner would provide useful information.

Superficial velocity V_{SL} was varied over the same range, 0.1 -2 m/s, as previously and the λ_W values were similar to the earlier values. Therefore a change results in the discrete values of V_{SO} and V_{SW} which were investigated. It is worth mentioning that the original (Oil No 1) values had been chosen to give some, if only a limited, comparison between total V_{SL} and the Oil No 2 V_{SL} values.

The test coding on Table 5-3 is as follows:

O W L 0 1 A

Code	A	B	C	D	E
λ_W(-)	0	0.2	0.4	0.5	0.6

Code	1	2	3	4	5
V_{SL} (m/s)	0.1	0.3	0.5	1.0	2.0

Pipe Inclination 0°

Oil No 2 (Lube oil)

It should also be pointed out, as in Chapter 2, that this approach is more representative of what occurs in a producing oilfield; over time oil production declines and is usually associated with a corresponding increase in the quantity of water being produced from the reservoir. Also, much of the previous research of oil/water pipe flow, reviewed in Chapter 3, has adopted this approach.

5.4.2.2 Flow Regime Results

Observed oil/water flow patterns were classified according to the sketch Fig 5.3 and the descriptions of 5.4.1.1 for the Oil No 1 tests. The data is shown is Fig 5.13 in terms of V_{SO}-V_{SW} coordinates and in Fig 5.14 in the form of V_{SL}- λ_w. Photographs of several of the flows are shown in Fig 5.15. All of the Oil No 2 data are tabulated in Table 5-3.

5.4.2.3 Oil/water Phase Fractions Results

In-situ fraction of water, $(WC)_a$, are compared to the no-slip values for each V_{SL} in Figs 5.16 to 5.19. The tests involving the highest V_{SL} produced problems where the test loop bursting disc ruptured on the second test. For this reason, although the disc was easily replaced, no further holdup trappings were made at this flow rate.

5.4.2.4 Oil/water Pressure Drop Data

For all Oil No. 2 tests pressure loss was measured through a 6m steel section on the return leg of the test loop. The pressure drops are plotted against the λ_w values in Fig 5.20. A sensitivity test was also run at V_{SL} = 2m/s where the value was varied in smaller increments up to 70% (vol). This information is given in Fig 5.21. Fig 5.20 does not include the data for V_{SL} = 0.1m/s as it was considered that the pressure drops involved at this flow rate were small enough to be considered outwith the reasonable expected accuracy of the measurement.

5.5 DISCUSSION

5.5.1 Flow Regime

As mentioned earlier, a wide range of oil/water flow regimes were observed within the experimental range of this study. However, it should be borne in mind that the experimental range was determined by requirements for the oil/water/gas tests, and as such the span of flow rates was somewhat smaller than was the case for oil-water regime studies by other investigators. It is instructive to discuss each flow pattern in turn and then consider the results in the context of data collected by previous workers.

Focussing attention initially on the Oil No. 1 tests, the lowest values of V_{SO} and V_{SW}

produced a separated flow where no interfacial mixing was observed at the end of the pipe. This flow represents the case which is envisaged as one of the easier regimes to tackle theoretically and has been the subject of much research attention in the past. However, this flow does not necessarily imply that turbulent conditions are completely absent, although at high oil viscosity and low water fractions laminar flow will certainly predominate. If we calculate a Reynolds number based on the total liquid superficial velocity, $(Re)_{SL}$, which uses 'averaged' quantities for viscosity and density, then even at the lowest liquid velocity the total liquid $(Re)_{SL}$ range is 2.6-3.6 x 10^3. Several analyses require calculation of the phase superficial Reynolds numbers $(Re)_{so}$ and $(Re)_{sw}$. At the lowest value of V_{SL} a $(Re)_{so}$ of less than 1500 is obtained, but in all other cases (including all the water flow rates) superficial Reynolds numbers in excess of 2000 are obtained. Therefore the present tests indicate that the vast majority involved turbulent-turbulent (oil-water) behaviour with a small number displaying laminar-turbulent flow. This point markedly illustrates the effect of the oil-phase viscosity; much of the early work involved significantly higher oil viscosities and consequently laminar flow is favoured.

As λ_w (and so V_{SW}) was increased, a small amount of liquid/liquid mixing was observed to produce occasional droplets at the interface of large diameter (up to approximately 10mm). Above and below these interface droplets the bulk oil and water phases were observed to be perfectly clear and quiescent with no indication of any turbulent-dispersion phenomena. A small increase in the oil flow rate meant that this flow regime became far more regular with the interface consisting of large droplets in an unbroken, thick droplet train . However, again the bulk phases above and below the droplet interface were observed to be very clean with insufficient mixing energy being available to induce oil/water emulsification. The droplets were of a similar diameter to those of flow B, therefore the increase in flow rate has served to produce more droplets rather than break up the existing droplets into smaller droplets.

This situation changes when, following further increase of flow rates, the mixing energy is sufficient to produce a highly turbulent interface consisting chiefly of much smaller droplets than in flow B and C. For this flow regime, type D, the oil layer above the thick, turbulent interface was observed to remain relatively clean. The layer below the interface was seen to consist of a very dilute oil-in-water dispersion, with the oil droplets being very small. This flow pattern was thus considerably more complex and chaotic than any of the flows observed to this point. Increase of the oil and water flow rates was seen to lead to progressively increasing homogeneity, as would be expected, where the oil and water were well-mixed in a pink-coloured dispersion where it was impossible, by visual means, to determine which phase was continuous and which was dispersed. In flow pattern E a colour gradient in the mixture was apparent suggesting incomplete mixing of the oil and water near the bottom of the pipe. Finally, the highest value of V_{SL} produced a flow pattern which appeared to be fully homogenous i.e. there was no observed colour gradient in the flow at the end of the pipe. It was noted in some cases that where a fully homogenous mixture was observed at pipe outlet, that observed near the pipe inlet was in fact semi-homogeneous. This demonstrates the effect of residence time, which is a function of flow rate and pipe length, on the homogeneity and tightness of the liquid/liquid dispersion.

Examination of Figs 5.13 to 5.15 reveals that very similar flow patterns were observed in the Oil No. 2 tests. This is expected to be due to the relative similarities of the viscosities of the two oils; the ratio between the oils' viscosities is about 3 at the upper limit and this difference was not expected to produce any noticeable effect on flow patterns. The flow rates and hence shear rates required to homogenise the oil/water mixture are very similar to the Oil No. 1 case: had an oil of significantly higher viscosity been selected this observation may not have resulted. It must also be admitted that the flow pattern maps, whilst covering a considerable range in flow rates do not contain sufficient data points to pin-point the flow regime transition bands with the accuracy of previous oil/water studies. However, comments on the flow pattern maps of several previous workers are considered relevant and are now discussed.

Of the large number of flow regime investigations undertaken in oil-water flow, the vast majority have involved small-bore pipes. Here, attention will first be centred on systems having pipe diameters and fluid properties which are not radically different to the present system. Thereafter, comments will be made as to the applicability to systems somewhat removed from that involved in the current study.

The map of Guzhov et al (1973) as shown Fig 3.3 contains detailed classification of the flow structures, and for convenient comparison only the transitions representing departure from stratified flow and the onset of fully homogeneous flow are shown on Figs 5.5 and 5.14. Guzhov's data was collected from a 39mm i.d., 18m long steel horizontal pipe, where the oil was a transformer oil of 22cP viscosity at operating conditions. The Guzhov map is seen to suggest non-stratified flow at higher values of V_{SL} than was observed here. The transition to fully homogeneous flow is similar to that observed in this study, although the earlier worker had a more discrete test matrix on which to base his judgement. The difference in the oil viscosity and interfacial tension as compared to the current test oils may also contribute to some variation between the data.

The map of Malinowsky (1975), which is in V_{SO}-V_{SW} form is depicted in Figs 5.4 and 5.13. In this case, comparison is convenient only for the transition to fully homogeneous flow. Malinowsky's study involved flow in a 1.5-inch i.d., 97ft long horizontal acrylic test loop, with a diesel oil of approximately 6 cP viscosity. This transition is shown to agree well with the current data.

The final comparison in this instance is made with the data of Oglesby (1979) as shown in Figs 5.5 and 5.14. This worker used a range of oils in an 84ft long, 1.5-inch i.d. acrylic horizontal test pipe. The transitions given in Figs 5.5 and 5.14 relate to the data from a 32 cP oil system. The stratified-non-stratified limit is seen to lie in the same region as obtained in the current tests, as does the transition to a fully mixed flow. Perhaps a more interesting feature here is the difference exhibited between the Guzhov et al and Oglesby results: these systems have similar pipe diameters and oil viscosities. The main differences in the experimental systems are the pipe length and material of construction. These points will be taken up in a later section.

Turning now to how the current and earlier work can relate to flows in systems of practical industrial interest, one is left with very little information with which to identify any clear trends. The main points of interest here are the oil properties, particularly viscosity and the pipe diameter and length/diameter ratio. Unfortunately, the only work available on oil-water flows in large-diameter pipes involves vertical or near-vertical flows where gravitational effects are much more dominant than in the horizontal or near-horizontal situation. Vigneaux et al (1988) and Zavereh et al (1989) have looked at flows in 7-8 inch i.d. pipes inclined at up to 65° and 15° respectively from the vertical. Both used oils of less than 3 cP viscosity, and in the former study no emulsification of the phases was obtained. Therefore little can be learned from these studies where a useful pipe-diameter scaleup is present.

The work of Oglesby (1979) does however allow limited insight into the role of oil viscosity in identical test equipment. Oil viscosity was varied over the range 32 to 167 cP. The study revealed that oil viscosity had little effect on the stratified-mixed regime transition. However, it was found that the homogeneous and semi-homogeneous patterns were obtained at lower V_{SL} as oil viscosity was decreased. This observation is to be expected due to the easier mixing of the lower viscosity oils. This character was only obtained with oil as the dominant phase: to the right of the inversion point, where water is the dominant phase, oil viscosity had little effect on the flow transitions.

5.5.2 Oil/Water Phase Fraction Results

Examination of the Oil No. 1 results, Figs 5.6 to 5.10 indicates that, within the accuracy of the measurement for the range of flow rates involved the oil and water exhibited no slippage between each other. This trend was observed irrespective of the prevailing flow regime. The Oil No. 2 data, which involved the same range in liquid superficial velocity V_{SL}, displayed identical trends as shown in Figs 5.16 to 5.19. Therefore, for a horizontal 2-inch pipe with oil viscosity up to 5 cP, oil-water flow is of a no-slip nature. The aforementioned study of Malinowsky (1975) also produced several measurements of oil and water fractions in oil-water flow. For flows in the segregated regime, the average ratio of the input oil fraction to the in-situ fraction was 1.06; for tests in the dispersed flow regime this value was 1.11. So this study also suggests little slippage between oil and water in horizontal flow. This is perhaps not surprising, given that the phenomenon of fluids slippage is usually associated with two-phase gas/liquid flow, where large differences in fluid physical properties are present. In such cases, density ratios of 800:1 (low pressure) and 10:1 (high pressure systems) are encountered.

Bouyancy helps the gas to slip over the liquid phase such that the liquid travels at a velocity lower than the no-slip case, and so is held-up to occupy a larger proportion of the pipe cross-section. In typical oil-water situations, the density ratio of the oil to water is in the range 0.7 to 0.9 and so the momentum of each layer is comparable. This, coupled to the relative similarity of each of the liquid's viscosities (in most cases), would suggest fluid slippage to be relatively insignificant in a horizontal pipe. Very little information is available on this aspect concerning larger-scale systems. Vigneaux et al (1988) observed oil-water flow in a large diameter pipe but only in inclined flow to a minimum of 35° to the horizontal. At certain flow rates, significant slippage between the oil and water was obtained. Although there is no quantitative data to support the observation, the flows in the inclined system of the current study (1 degree from the horizontal) in Chapter 7 did produce oil-water slippage at low liquid superficial velocities.

It appears that, for the range of oil viscosity involved in the current study, the effect of pipe inclination is more important to oil-water slippage than is the present fluid physical properties range.

In summary, these tests have shown that in the horizontal flow no slippage occurred in oil-water flow. Therefore any oil-water slippage identified in the horizontal oil/water/gas tests should be attributable to the presence of the gas phase.

5.5.3 Oil-Water Pressure Drop

The collection of pressure drop data in oil-water flow was an important aspect of this experimental phase. The data collected from the Oil No. 2 horizontal tests will now be discussed.

Inspection of Fig 5.20 reveals that at the lower values of total V_{SL} a relatively flat dp/dx -λ_w relationship was obtained suggesting that no significant effects regarding the formation of oil-water dispersions were present. At the lower V_{SL} the measurement uncertainty is higher than that for the higher V_{SL} tests, but the data at all values of V_{SL} were found to be repeatable. The major interest in the pressure drop is revealed to be at the highest V_{SL} where a fully homogeneous (by visual observation) flow was obtained. Whilst no droplet-size measurements were undertaken, photographs revealed droplets of a size easily observable to the human eye, although of course the photographs centre on one plane and were, unfortunately, open to distortion effects.

This homogeneous flow produced a slight dip in the pressure loss as water was added to the flow, the minimum being at approximately 40% input water content. At the highest $\lambda_w = 0.7$, the pressure drop was seen to increase to a pressure loss similar to the oil-only value. Additional experiments were performed to specifically examine this behaviour. The results are shown in Fig 5.21 which confirm the existance of a pressure-loss minimum. The pressure drop reduction is of some 15-20%, based on oil-only flow as the datum (although it should be remembered that different oil volumetric throughputs are involved for each water cut).

The possibility of using water to help reduce oil-line pressure losses produced many studies from Canada in the 1960s. However, important differences are present between such work and the current study.

In the early situations, the oil was very much more viscous than the water in most cases, with viscosities up to tens of thousands of centipoise being present, and laminar-laminar flow was predominant. In the current study, much lower viscosities are present, and the vast majority of the runs involve turbulent flow in each layer (on a superficial basis for each of the oil and water flows). This may help explain why, for a fixed V_{SO}, no situations arose where the introduction of water resulted in a decrease in the measured friction losses. Attention will now focus on further data examination and results viewed in the light of more recent data involving conditions more closely related to the current experiments. The V_{SL} = 2 m/s data only is examined due to the existance of more data points and the possibility that homogeneous type flows will be important to the three-phase data.

It is of interest to compare the results of the present study to the predictions from the empirical method of Charles and Lilleheht (1966); the data is given on Table 5-4 with Fig. 5.22 showing the results at the 4 pricipal values of input water cut. This correlation is basically a liquid-liquid adaptation of the gas/liquid correlation of Lockhart and Martinelli (1949). The X^2 parameter increasing can be viewed as input water fraction decreasing for the current data. It is seen that for most cases the correlation gives consistent underpredictions of about 10-20% when compared to the present data. This is encouraging from the viewpoint of obtaining quick estimates for oil-water pressure drops. What is also interesting, is that the data for V_{SL} = 2 m/s are all of turbulent-turbulent character. Previous comparisons to the Charles-Lilleheht correlation have involved chiefly laminar-laminar and laminar-turbulent behaviour (Wang and Charles (1981)). Therefore, based on the limited data of the present study, the correlation would seem to be of value when turbulent-turbulent behaviour is involved.

Closer examination of Fig 5.22 shows that at the lowest values of X^2, the pressure loss ratio does start to diverge somewhat from the empirical curve. This is consistent with the increase in losses measured at water cuts of greater than 50%, and would be expected to be related to the phenomena mentioned previously.

A similar comparison between the present data and the model of Brauner and Maron (1989) is expected to yield similar results to that depicted in Fig 5.22. These authors applied a separated-flow physical modelling approach, and the pressure drop ratio - X^2 relationship produces similar results to the Charles-Lilleheht empirical curve, but is much more rigorous and less convenient to apply. This result is intriguing in that the model was developed specifically for stratified flow conditions. Although no information was presented by the authors regarding the model applicability to non-stratified flows, they do comment that it is not necessary for a stratified flow to exist for the range of variables considered in their model. It should be added that turbulent-turbulent flow is considered which, to this author's knowledge, has not received attention until this time. The model does not include considerations concerning the interfacial behaviour or system chemistry: a non-stratified case would be expected to exacerbate this practical deficiency. Attention will now turn to other oil-water experimental efforts.

The study of Guzhov et al (1973), also mentioned in 5.5.1, produced pressure drop data worthy of discussion. Referring to Fig 5.23, a peak in the dp/dx - λ_w relationship was obtained in some cases, with the peak generally occurring as V_{SL} was increased from 0.5 m/s to 1.1 m/s at λ_w = 0.1 to 0.2. At V_{SL} = 1.7m/s no peak was obtained. The authors attribute this to the nature of the flow regime, but it should be remembered the overall behaviour is one of decreasing pressure drop as the input water fraction is increased.

Work by Oglesby (1979) suggested the oil-water pressure drop to be dominated by dispersion-type phenomena. The least viscous oil used by this author was approximately 32 cP, and attention here is focussed on this data. Inspection of Fig 5.24 reveals that the pressure loss behaviour is far from simple. At mixture velocity (i.e. V_{SL}) 0.9m/s, a semi-homogenous flow regime was noted and this is seen to produce a pressure loss which is fairly insensitive to the quantity of water input. However, as the flow rate is increased interesting features in the dp/dx - λ_w relationship emerge. At V_{SL} = 2.1m/s, a slight peak in pressure drop is noted at λ_w = 0.3, falling to a pressure loss slightly below the single-phase oil value. The existance of this peak can be attributed to increased shearing of the flow to produce smaller droplets; this tighter dispersion would generally be favoured to produce a higher apparent viscosity than one containing larger dispersed droplets, although caution should be urged here in that it is both the droplet size and the size distribution which are of influence. Commencing from single-phase oil, dispersed water will initially form a water-in-oil dispersion with water droplets dispersed in the oil-phase continuum. The limit to which the water can be dispersed (in volume fraction terms) until the dispersion inverts to an oil-in-water type is expected to be a function of shear rate and oil phase viscosity. An oil-in-water dispersion viscosity is expected to be close to that of water-alone (at least when compared with the dry oil viscosity), and so a decrease in pressure loss is favoured. In Fig 5.24, an increase in V_{SL} to 3.7m/s should result in further increased shearing of the dispersion, however only a fall in pressure drop was recorded as water input was introduced and increased. Therefore, a curious situation has arisen in that the case where the maximum mixing intensity is present did not produce any peak in the pressure loss. This further complicates any possible explanation of the oil-water pressure drop, as will be encountered later.

Recent work by Stapelberg et al (1991) included measurement of oil-water pressure loss in 59mm i.d. piping, using a test oil of 31 cP viscosity. These authors also chose to plot their data in the manner of the Charles and Lilleheht (1966) modifications of the Lockhart-Martinelli (1949) method. Generally satisfactory agreement was obtained, although at high X^2 the pressure loss ratio was less than 1. Included in Fig 5.22 is a sample of their data for their oil drops flow regime, where it can be seen that below X^2 values of about 2 the measured pressure drop ratios were considerably higher than those expected from the Charles-Lilleheht method. This is a trend supported by the present data, although the departures from the correlation for the Stapelberg et al data are somewhat higher. Given that the oil viscosity in the current study was about 15% of that involved in the Stapelberg et al study, it may be that the higher oil viscosity has promoted the discrepancy between correlation and data.

The present pressure-drop data cannot be explained precisely by existing theory or conveniently in terms of existing data. The reduction in pressure drop as water cut is increased is consistent with the notion of the formation of a lower-viscosity substrate which would favour lower friction losses. The increased pressure drop as the water fraction is increased above 50% may be suggested to be due to the formation of a more viscous water-in-oil emulsion. However, both these suggestions encounter difficulty when the system is examined in detail. The highest V_{SL} data which produced the small dip in the pressure loss characteristic was considered to be homogeneous with no water or emulsion film being observed at all. This statement cannot truly be said to be certain if one is considering the microscopic system as opposed to the macroscopic situation.

Unfortunately, within the scope of the current study it was not possible to examine such microscopic effects. It is clear, however, that trying to apply the emulsion viscosity method of, for example, Woelflin (1947), to the current dynamic oil-water pipe flow is a severe oversimplification of the physical effects which are simultaneously at work during the flow. The current data suggests that the following aspects are of critical importance to the frictional pressure loss behaviour:

i Energy input

ii Degree of homogeneity

iii Droplet coalescence and breakup phenomena

The three factors above are of course strongly related. In the pipeflow situation the energy input can be viewed as the velocity attained by the flow. The interesting feature here is that only at the highest value of V_{SL} was the pressure drop found to give a dip at mid-range λ_w i.e. the velocity at which the system would be expected to be at it's most chaotic. The degree of homogeneity is related to the system energy input; again, where semi-homogeneous flows were observed no pressure drop dip was measured. The degree of homogeneity (or extent of dispersion) may be an important consideration in the current system.

In a semi-homogenous flow, it is conceivable that the upper fluid is a water-in-oil type dispersion, whereas the lower fluid is largely a dilute oil-in-water dispersion.This will then complicate the pressure drop so far as the overall apparent viscosity is concerned: if such a mixture of dispersions does exist, the pressure loss will depend on the extent to which the water has dispersed in the oil phase and vice versa. The pressure drop behaviour of the V_{SL} = 2 m/s data would seem to be best explained by the third aspect above. If an oil-in-water dispersion is flowing through the pipe, it would seem that around the λ_w = 0.4-0.5 range there is some effect on the droplet behaviour in the dispersion.

The increase of λ_w from 0.4 to 0.7 may have altered the coalescence/breakup behaviour to an extent where the droplet size and droplet size distribution have been modified. The physical properties of the system, their chemistry and their interaction with the pipe wall (in terms of wettability) are factors which may be crucial to describing the system behaviour properly. The manner in which the droplets interfere with each other is critical to the droplet size distribution; it is easy to focus only on the element of droplet breakup when considering this aspect. Much depends on the droplet size distribution which is generated at the liquid/liquid mixer; it is conceivable that coalescence plays a dominant role in the pipeflow where no obstructions are present, as noted by Ward and Knudsen (1967). It is suggested here that the key to explaining the pressure drop behaviour lies in the extent of liquids mixing and the relative influences of coalescence and break up on the droplet size distribution for the homogenous and semi-homogeneous flows.

For the largely stratified flow with the continuous bubbly interface, the relatively flat pressure loss relationship may be due to the degree of mixing at the oil-water interface. It should be noted that although the Oil No.2 viscosity was about 4 times that of the water, the predicted pressure drops at the extreme ends of the water cut scale are not entirely dissimilar, and therefore a large difference between the oil-only and water-only friction losses was not expected.

Much of the above discussion will be returned to in the discussion of the three-phase data in later chapters.

5.6 CONCLUSIONS

The collection of oil-water data has provided valuable information as to the hydraulics of a liquid-liquid mixture, which will be useful when used as a comparison for the oil/water/gas tests. The principal conclusions arising from the experiments are:

i Flow regimes were generally similar to those reported for earlier experimental systems.

ii Complete oil-water mixing was only visually observed at the highest V_{SL} of 2 m/s.

iii The holdup data indicated the oil-water flow to be of a no-slip nature.

iv No existing theories are completely adequate in representing the trends observed in the pressure loss data.

v For homogeneous flow at a single value of V_{SL}, input water fractions up to 0.5 produced pressure drops which were slightly above that predicted by the Charles and Lilleheht (1966) correlation.

vi No significant peak in pressure loss due to dispersion phenomena was measured.

CHAPTER 6

OIL/WATER/GAS FLOW IN A HORIZONTAL FLOW

6.1 INTRODUCTION

This chapter presents the data collected involving oil/water/gas flow in a horizontal pipe, the procedure adopted in obtaining the data and a discussion of the results. These tests involved using alternative test oils, both of low viscosity, and are the same oils discussed in Chapter 5.

Whilst it was not proposed to enter into any physical modelling of the data, arguments and thoughts as to the relative importance of many aspects as regards the scaling to different systems will be based largely on the experimental work of this chapter.

6.2 TEST OBJECTIVES

The primary objectives of the experimental exercise were:

i To collect flow regime, liquid holdup, in-situ water fraction and pressure drop data in a three-phase system.

ii To examine the effects of using 2 different viscosity test oils.

A secondary objective of the study was to collect slug characteristics data from an oil/water/gas flow system.

6.3 EXPERIMENTAL PROCEDURE

Following the period spent during facility commissioning and experience gained from the oil-water experiments a stringent running procedure was adopted when performing the oil/water/gas tests which required the presence of at least 2 workers.

Instrumentation was switched on early each morning, and the works compressor was used to charge the control valves and fast-acting pneumatic valves with instrument air. At this point the test loop was open to atmosphere as a general practice of facility shutdown from the previous operational day, and the loop pressure transducers could be zeroed using the signal conditioner and computer outputs. The test loop was then isolated from the separation system using a manual valve at the pigging bypass and, using the charged-up facility compressor, air was bled into the loop to a pressure of 0.7-1 barg as monitored on an analogue pressure gauge on the loop. At this point gas input was halted, and the pressure transducers were checked to ensure conformity at the shut-in static gas pressure. Occasionally at this point the loop was blown down, and re-charged to the same pressure to check any instrumentation drift effects; however, no problems were encountered. Also, occasionally the test loop was pressurised to up to 2 barg to check for leaks and to check the performance of the pressure transducers.

With the test loop isolated at 0.7-1 barg, the gas outlet valve on the primary separator was shut and, using air from the works ring-main, the separator was pressurised to 0.6 barg as indicated by the Bourdon gauge on the vessel. Once the required pressure had been reached, air input was halted, and the pneumatic pressure controller set to maintain the pressure in the separator.

A final check was then made to ensure the test loop gas input valves were closed, and that nitrogen back-pressure was on the test loop pressure transducers. The latter was accomplished via a 2-minute pressure transducer sampling which was written to file STATEST: once the operator was satisfied with the instrumentation, oil was passed to the test loop. As oil passed to the loop the outlet valve was gradually opened against the separator back-pressure, and in a short period the flow stabilised as did the loop and separator pressures. The oil was then set to the desired flow rate using the appropriate turbine flowmeter and manual ball valve. Water was then added to the flow and, when the water had passed through the test loop, gas was introduced at the gas/liquids mixer. The gas was then set to the lowest anticipated value of V_{sg} = 1m/s at the metering section. In most cases, following the addition of gas, the oil and water inputs required adjustment and sufficient time was allowed to enable the three flow rates to stabilise against the constant separator pressure.

The separator level control valves and coalescer valves were adjusted to maintain the required liquid level in the separator, as indicated on the sight glasses, and pass fluids back to the storage tanks.

When the flow rate had stabilised sufficiently, the test loop data acquisition program was run for a period of 3 minutes. Following this, three measurements of liquid holdup were taken using the fast-acting valves, where it was ensured that between each trapping the fluid flow rates were stable at the required values. Once the holdup measurements were complete, the next gas rate was set which also involved re-adjustment of the oil and water flow rates to maintain the original values.

This procedure was repeated up to the required maximum gas flow rate - usually at least V_{sg} = 7m/s at the gas metering section. The water input was then stopped, the gas adjusted to a low flow rate and oil and air were used to sweep water from the loop in preparation for the next set of oil/water ratios. Finally, oil input was halted and air at high velocity was used to purge the test loop of any oil. When the pipe was dry, air input was stopped and the test loop was shut in at the original before-test pressure of 0.7-1 barg. The data acquisition program was run for 2 minutes writing to file STATEST1, which was inspected to check any drift of the pressure transducers over the testing period. The results for these static tests were noted on the test data record sheets.

If another testing session was planned later in the day, the loop and separator were left pressurised to expedite start-up and in addition the former meant that checking of the pressure transducer performances was possible. The pressure in the loop and separator was lowered to atmospheric at the end of the working day.

6.4 DATA PRESENTATION

6.4.1 Oil No. 1 Data

The first set of tests used the kerosene test oil, designated Oil No. 1, the selection and properties of which are given in detail in Chapters 4 and 5. The following section outlines the information collected and it's relation to the previous oil-water tests and further oil/water/gas investigations.

6.4.1.1 Experimental Variables

As mentioned in 5.4.1.1, it was intended that the oil/water/gas tests should be an extension to the oil-water tests in the sense that similar oil-water flow rates would be used, where the gas flow would be superimposed on the total liquids throughput. Therefore the oil flow rate would be maintained constant and increasing volumes of water were added to change the system input water-cut λ_w. This means that the tests were conducted at set values of gas-oil ratio (GOR), and the oil superficial velocity V_{so}, and water superficial velocity V_{sw}, ranges were the same as for the Oil No. 1 - water experiments. It was decided that, to keep the number of experiments required to a manageable level whilst still capturing the required information, 4 values of gas superficial velocity V_{sg} would be involved. As mentioned in previous tests, the gas is metered at near - pipeline inlet pressure, although this is corrected at the location of the pressure drop measurement sections. The target values decided upon for the V_{sg} at the gas metering section were 1, 3, 5 and 7 m/s; this meant that the maximum in-situ V_{sg} was approximately 10 m/s.

This upper value of V_{sg} was found by early experience to be the maximum value obtainable at the highest liquid flowrates. At such flowrates, the pressure loss through the test loop was so high that, when combined with the separator back-pressure, the capacity limit of the oil pump was being met.

In experimental terms, the range of V_{sg} chosen is within reasonable practical limits, and is shown on both the Mandhane et al (1974) map, and the Beggs and Brill (1973) flow regime map on Figs 6.1 and 6.2 respectively, where V_{SL} refers to the total liquid superficial velocity. These maps show the range of experimental data available from the literature on three-phase flow: it is clear that the proposed experiments largely encompass and significantly add to the sparse information. Although not shown on Fig 6.2, the vast majority of the data of Stapelberg et al (1991) also fall within the current flow rate matrix. In field-relevant terms, the following approximate ranges are involved:

GOR: min = 1.2 m^3/m^3 = 12 scf/bbl in oil-field units

max = 87 m^3/m^3 = 830 scf/bbl

GLR: min = 0.6 m^3/m^3 = 9 scf/bbl

max = 87 m^3/m^3 = 830 scf/bbl

In typical oil/gas field systems, GORs of less than 1000scf/bbl are very common, and so the experiments here are intentionally more tailored to an oil/gas system than a gas/condensate fluid, where in the latter very high GOR/GLR are evident with only a small amount of liquid being carried with the gas. It is also worth pointing out that the three-phase flow problem is identified as being one associated to a greater extent with crude oil/water systems than gas/condensate hydrocarbon fluids.

The following test coding has been employed for the Oil No. 1 data presentation:

3 P 0 1 A A

Code	A	B	C	D
Target V_{sg} (m/s)	1.0	3.0	5.0	7.0

Code	A	B	C	D	E
λ_w (-)	0	0.2	0.4	0.5	0.7

Code	1	2	3	4	5
V_{so} (m/s)	0.1	0.15	0.3	0.5	1.0

Pipe Inclination 0°

6.4.1.2 Flow Regime Data

For all tests during stable conditions, flow regime was noted at 4 locations on the loop: loop inlet, at points approximately 4m upstream and downstream of the return bend, and approximately 3m from the pipe outlet. This was accomplished using 2 observers, and observation alone (coupled with videos where appropriate) was used to denote the prevailing flow pattern. Both the way in which to present the data and how to describe the flow regimes suggested a number of possibilities. Examination of the map produced by Sobocinski (1955), Fig 3.9, shows how one could use a large number of patterns to describe the flow. It was decided in this study that to provide such an analysis was outside the capability of the identification method, as it was impractical to take videos/fast-frame photographs on every test. Instead, it was decided to use classifications which are an extension of the two-phase flow regimes and to highlight the behaviour of the oil and water in the mixture. The flow regime descriptions used are as follows:

A: Predominantly stratified flow where the liquid film consists of a separated oil-water mixture or an oil-dilute emulsion mixture.

B: Predominantly stratified flow where the film was observed to be a homogeneous mixture of oil and water.

C: Infrequent slug flow where the slug and film are of oil/water or oil/emulsion separated flow.

D: Regular slug flow with features as for flow C.

E: Infrequent slug flow where the slug and film are a homogeneous oil/water mixture.

F: Regular slug flow with features as for flow E.

G: Infrequent pseudo-slug flow where the pseudo-slug and film are of an oil-water or oil-emulsion separated nature.

H: Frequent pseudo-slug flow where the oil and water display characteristics as flow G.

J: Infrequent pseudo-slug flow where the pseudo-slug and film consist of a homogeneous oil/water mixture.

K: Frequent pseudo-slug flow with oil/water behaviour as for flow regime J.

Fig 6.3 depicts the descriptions above. In the two-phase literature, much discussion exists as to the limits of stable slug flow. Many workers have broken up the intermittent flow regime into a number of sub-regimes, where overall an intermittent type of flow is observed to exist. In this study it was felt appropriate to indicate intermittent flow as two forms; slug flow and pseudo-slug flow. Pseudo slugs were defined as liquid units of less than 10 pipe diameters in length, and which were observed to only briefly touch the top of the pipe before decaying. More discussion on this subject is given in a later section.

The manner in which the flow regime data is presented also gave several options. As it was intended that any map should be useable to workers familiar with existing two-phase terminology, it was felt necessary to avoid any complications which would somehow overshadow such information. The idea of presenting the data in V_{so} - V_{sw} - V_{sg} format had been put forward but rejected by the research sponsors. It was decided that the best method was to present the data in V_{so} -V_{sg} form, with a different map for each λ_w, which is consistent with the vast majority of flow regime maps which are in V_{SL} - V_{sg} format. The only possible confusion is that the V_{so} axis will under most cases not be the V_{SL} (due to the added water), but the sponsors indicated that this format was the most acceptable. The data is given in Figs 6.4 to 6.8 for the location at the test loop half-way point, and in Figs 6.9 to 6.13 for the observation point close to the loop outlet.

This was felt necessary to show up any effects of the (L/d) ratio on the flow structures, particularly on the oil-water behaviour. All the information in these figures is summarised in Tables 6-1 to 6-5 . Fig 6.14 shows photographs of the flow regime observed near the pipe outlet for several conditions.

6.4.1.3 Liquids Holdup Data

Liquid holdup (i.e oil plus water) was measured upstream of the return bend, and for each value of V_{so}, the H_L - V_{sg} relationships are displayed on Figs 6.15 to 6.19. The values given, also included in Tables 6-1 to 6-5, are the average of the 3 trappings taken in each test. Further tests were undertaken to look at the sensitivity of taking a higher number of samples, which are described later. In each test, it was also possible to determine the in-situ water cuts, $(WC)_a$, and these are plotted against V_{sg}, and compared to the no-slip values λ_w, in Figs 6.20 to 6.24.

6.4.1.4 Pressure Loss Data

Pipeline pressure drop measured in the steel section of pipe downstream of the return bend is given in Figs 6.25 to 6.29 where each graph concentrates on a single V_{so} and the relationship with λ_w and V_{sg} as in 6.4.1.3. In order to display the effect of water cut where V_{SL} and V_{sg} are fixed, Figs 6.30 and 6.31 show the measured pressure drops at V_{SL} = 0.5 and 1.0m/s, respectively. All of the above data is given in Tables 6-1 to 6-5.

6.4.1.5 Slug Characteristics Data

Intermittent-type flows were expected to be present for much of the experimental range, and so it was considered that slug characteristics data would be of value, although it was a secondary objective of the experiment. In order for the existing facility hardware and software to be used, experience proved that the use of the LED liquid level probes was impractical in the small-bore pipe. Using a video-based technique, slug lengths and slug front velocities have been obtained for a small number of liquid flow rates at the lowest V_{sg}, and taken at a location approximately 800d from the pipe inlet.

Figs 6.32 to 6.34 show this data with the associated water cuts and liquid mixture velocities. The slug length given is the average of 100 values, as are the front velocity and slug frequency data. The slug length distributions are displayed in Figs 6.35 to 6.40. Raw data is presented on Table 6-6.

6.4.1.6 Slug Holdup Sensitivity Tests

These tests were set up not to obtain data to be used for comparison against correlation predictions, but more as a method of which to test the performance of the holdup measurement technique where slug flow is prevalent. Slug flow is by its nature a process where regular slug units are not always produced, and is a complex phenomena involving mass-balance, hydrodynamic and terrain-induced effects. The most suitable method to measure slug body holdup i.e. the actual liquid fraction in the liquid slug unit is to use a continuously-sampling method where a large number of slugs can be counted and an average obtained which is statistically acceptable. The measurement of slug body holdup, for a small-diameter pipe, requires an instrument of fast response (typically 50Hz or better). The measurement of average slug holdup i.e. that involving both the slug body and liquid film, is subject to the same constraint in terms of statistical acceptance, but not perhaps in the case of required response time. In this study, although the pneumatic valve bypass was designed to incorporate the measurement of slug body holdup, it was found impractical for several reasons. Average slug holdup has been measured using the apparatus, where as mentioned 3 trappings of the flow are involved, each being a 10m sample of the flow. There is an element of operators judgement involved in when to switch the valves and record the sample: this subjectivity is clearly most important when an infrequent slug flow is present. Therefore for 3 values of V_{SL}, 10 samples of the flow for average holdup determination were taken, with 2 values of λ_w in each case.

The particular flow rate combinations were taken to satisfy the requirement that infrequent and regular slug flows should be involved, and the tests should be consistent with previous tests to enable a direct comparison between the 10 sample and 3 sample methods. The values obtained for each test and sample are depicted in Figs 6.41 to 6.46, where the oil and water contents and the comparison between the average values for 10 and 3 samples are also included.

6.4.2 Oil No. 2 Data

Consistent with the discussion of Chapter 5, three-phase tests involving the same number of data points as collected using Oil No. 1 were run where the earlier kerosene test oil had been replaced by an oil of higher viscosity. The characteristics of this oil in terms of basic fluid properties are given in Figs 5.11, 5.12 and Table 5-2. Of additional interest are the tests performed to investigate oil-water dispersion aspects and oil-gas foaming; see Appendix A.

6.4.2.1 Experimental Variables

As mentioned in 5.4, the use of the different test oil was accompanied by the decision to run the tests in a different manner which meant that tests were conducted at constant gas/liquids ratio (GLR) for values of water cut λ_w, rather than gas/oil ratio (GOR) remaining constant as in 6.4.1. However, the values of λ_w, the maximum V_{SL} and the values of V_{sg} involved were kept consistent with the Oil No. 1 tests. The approximate ranges of GOR and GLR involved are:

GOR: min = 0.6 m^3/m^3 = 9 scf/bbl

max = 280 m^3/m^3 = 2700 scf/bbl

GLR: min = 0.6 m^3/m^3 = 9 scf/bbl

max = 84 m^3/m^3 = 800 scf/bbl

The range of GOR is thus seen to incorporate the values involved in the Oil No. 1 tests, and the GLR range is identical to that present in these earlier tests. The experimental test coding employed was as follows:

3 D 0 1 A A

Code	A	B	C	D
Target V_{sg} (m/s)	1.0	3.0	5.0	7.0

Code	A	B	C	D	E
λ_W (-)	0	0.2	0.4	0.5	0.7

Code	1	2	3	4	5
V_{SL} (m/s)	0.1	0.3	0.5	1.0	2.0

Pipe Inclination 0°

6.4.2.2 Flow Regime Data

The operational and data collection procedures devised in the commissioning Oil No. 1 three-phase tests was followed strictly in these tests also. The flow regime was noted at the same locations, and the descriptions used to characterise the data is as used in 6.4.1.2.

This was considered necessary to provide consistency and ease of comparison. It is noted that the use of a different oil, though perhaps not expected to radically affect the bulk flow regime, could lead to characteristics which are apparent through close examination and such information is given in the discussion of the flow pattern data. Note also that the presentation in terms of V_{sg} and V_{SL} coordinates is strictly consistent with two-phase flow regime maps, unlike the Oil No. 1 tests, where the V_{SL} value should be obtained by corrections using the V_{so} and λ_w values. For each of the Figs 6.47 to 6.56 then, V_{SL} is the total liquid superficial velocity, where a map has been produced for each input water cut. Fig 6.57 depicts the flow regimes at several conditions observed near the pipe outlet.

6.4.2.3 Liquid Holdup

Total liquid holdup (oil plus water) is plotted against V_{sg} for each V_{SL} in Figs 6.58 to 6.62, where all water cuts are included. The in-situ water cut $(WC)_a$ was also recovered in the vast majority of tests, despite fears that unacceptably long settling times may be required to properly separate the oil and water. Those tests with no $(WC)_a$ value are those where insufficient oil-water separation occurred. The data is plotted consistent with Oil No. 1 nomenclature on Figs 6.63 to 6.67, and all data is produced in Tables 6-7 to 6-11.

6.4.2.4 Pressure Loss Data

Pressure drops measured in the steel section downstream of the return bend are plotted in dp/dx - λ_w format for each V_{SL} in Figs 6.68 to 6.72. Each graph gives plots for the V_{sg} involved in each run; the V_{sg} value is the average value of the V_{sg} encountered in the each of the λ_w tests for the appropriate V_{SL}. A small number of tests were also run where the V_{SL} and V_{sg} were held at a constant value, and the water cut was changed by much smaller increments than the standard λ_w values: these data are shown in Figs 6.73 and 6.74.

6.5 DISCUSSION OF RESULTS

In this section, discussion of the results of the 3-phase tests will concentrate on each technical area at a time, but at this point the data will be treated together, with particular aspects of each of the oil's behaviour being highlighted.

6.5.1 Flow Regime Data

6.5.1.1 General Overview of Oil No. 1 and Oil No. 2 Data

For the Oil No. 1 data, at the lowest V_{so} and $\lambda_w = 0.2$, stratified-type flows A and B were observed. For flow A, the settled layer was very bubbly at the downstream end, as depicted in Fig 6.14, but at the 500d position a much calmer water film was observed. This point is then the first of many demonstrating the value of looking at flows at L/d = 500 and 1000; an initial idea of the effect of pipe length can be obtained. This flow is as close to the idealised three-layer concept as was observed in any of the tests.

This case presents a theoretical basis by which mathematical modelling of the flow structure could be approached: this is discussed in a later chapter. At V_{SL} of 0.1m/s in the Oil No. 2 data (Fig 6.54), stratified-type flows were also in evidence, but this time the production of infrequent slugs/roll waves was also apparent.

For V_{so} = 0.1m/s and λ_w = 0.4, Fig 6.11, the increased liquid throughput (equivalent to V_{SL} = 0.167m/s), was sufficient to induce infrequent slugs and pseudo-slugs, approximately every 15 to 20 seconds or more. The flows labelled "infrequent" slug or pseudo-slug involve the situation where there was typically no more than one slug/pseudo-slug in each leg of the test loop simultaneously, and often only one slug/pseudo-slug would exist in the loop. These flows, types C and G, involved the slug/pseudo slug consisting of a bulk oil layer over an oil-water emulsion layer. Often the slug or pseudo slug was sufficiently unstable and would disintegrate within a few metres of forming, suggesting that the flow was within the transition range between stratified and intermittent-type flow patterns. As the input gas rate is increased at V_{so} =0.1m/s, the oil-gas flow approached an annular type character at the highest value of V_{sg}, where a continuous wall-wetting structure was obtained and the inside of the pipe was coated with a film which was thin at the top and thick at the bottom, as would be expected. However, the introduction of water in small amounts resulted in a flow where droplets were torn from the film and deposited on the wall in the form of streaks; often the streaks coalesced with other droplets into larger aggregations which look like soapy 'suds', and eventually drained back to the liquid film.

However, no classic annular flow as observed in oil-gas was obtained. The introduction of water has clearly upset the liquid entrainment behaviour existent in the oil-gas flow situation. With Oil No. 1, the film was observed to be fully homogeneous, however Oil No. 2 data suggested that the oil-water mixture remained semi-homogeneous at the highest V_{sg}. This is consistent with the suggestion that a higher viscosity liquid should be slightly more difficult to mix fully with the water, but it is admitted that the viscosity difference of the oils is relatively small.

Remaining with the Oil No. 2 data, V_{SL} = 0.1 m/s, it is pertinent to mention the infrequent slug/pseudo-slug flows. At a water cut of 50%, the infrequent slugs were observed to consist of 3-layers - a clean oil layer, dilute oil-water mixture and a clear water film. No flows were observed where the entire slug consisted of a clear oil-clean water composition, indicating that by the nature of slug flow itself, there is sufficient energy to disrupt the oil-water interface, even at the lowest value of V_{sg}. Video analysis revealed that, at the slug front, the oil and water were quite well mixed. However, a certain distance behind the slug front the main body was observed to be of a well-separated oil/water nature. Therefore the turbulent mixing associated with the slug front helped homogenise the oil and water, but it was insufficient to homogenise the bulk of the slug. This is an important point regarding corrosion of pipelines involving slug flow at low velocities and hilly terrain: if any settled water can be homogenised with the slug, corrosion attack will be mitigated. However, as mentioned later, it is an extremely difficult subject in which to derive useable theoretical relationships.

At V_{so} = 0.1 m/s, λ_w = 0.7, sufficient liquid had been added to the flow which resulted in more stable intermittent flows, types D, F and K; note that this is equivalent to V_{SL} = 0.33m/s. As seen in Fig 6.14, the slug front formation shows a vertical colour gradient which is typical of many of the slugs observed at the lowest V_{sg} where a concentration gradient was a more accurate description than a separated oil-water mixture. Also, it should be mentioned that even for λ_w of up to 0.5, the slug front was, judging from the colour, of a higher oil concentration.

At V_{so} = 0.3 m/s and for all water cuts, Oil No. 1 produced intermittent-type flows in all cases, where the frequency of slug production was mostly dependent on the liquid input. At the higher values of V_{so} and λ_w, homogeneous flows were observed at the end of the pipe in all cases. The Oil No.1-water tests had indicated complete oil/water mixing at V_{SL} approximately 1.5 m/s : it is clear that in the 3-phase tests a lower V_{SL} was needed to obtain oil-water mixing . This is not surprising, if one considers the energy imparted to the flow by the gas phase. Complete mixing was not always achieved, however, at the 500d location. It was evident that the degree of oil-water mixing between the start of the loop, the midpoint and the 1000d position, was in many cases variable. The increased length of pipe, subsequent friction loss and gas expansion contribute largely to the improved oil-water emulsification. As regards the length and frequency of the slugs produced in the Oil No. 1 tests, this is reserved to a later discussion.

For Oil No. 2, V_{SL} = 0.3 m/s resulted in intermittent flows at all conditions, as for the previous test oil. Fairly frequent slugs of up to 3m long were observed at the lowest V_{sg}: increase of V_{sg} resulted in progressively shorter slugs and eventually the production of regular pseudo-slugs. At the highest V_{sg}, in between the pseudo-slugs and roll waves the wall-wetting was of an annular nature, for the zero water cut flow only. At λ_W = 0.2, the oil-water mixture was observed to be homogenous only at the highest V_{sg}. It was evident that at this V_{SL} a higher degree of foaming existed in the flow than was apparent at the previous V_{SL}: this observation was also made of the Oil No. 1 tests. It is perhaps more accurate to describe the flows at highest V_{sg} to be pseudo-slug/foam flow, and Fig 6.57 shows the nature of the flow in between the pseudo-slugs, which is very chaotic.

The behaviour of a 3-phase flow is complicated particularly by oil-water dispersion aspects, which are of principal interest here, but it must be mentioned that gas/liquid foaming may have an effect not previously considered; this is considered later.

At V_{SL} = 0.5 m/s, visual observations suggested the oil/gas and oil/water/gas slug lengths to be similar at the low V_{sg} but slugs were produced slightly more frequently in the water-cut case. At λ_w = 0.2, oil/water mixing was observed to be complete at the end of the pipe. However, increased λ_w meant that higher V_{sg} was required to fully mix the oil and water. At the highest V_{sg} an almost continuous film of small droplets/foam were seen to coat the pipe wall. Especially at the higher water cuts, large liquid droplets and foam layers were observed to exist on the wall at h_L/d > 0.5, between the produced slugs and pseudo-slugs.

A superficial liquid velocity of 1m/s resulted in slugs which, at the lowest V_{sg}, were produced at 1-2 second intervals. For all conditions, oil/water mixing was complete at the end of the pipe; however, this was not necessarily true at the 500d position. Large droplets were seen to exist above the liquid film between pseudo-slugs, and, especially at the higher V_{sg}, the pseudo-slugs consisted of chaotic foamy packets of fluid.

Where the V_{SL} was at its highest value of 2 m/s, in all cases the flow regime was of an intermittent and oil-water homogeneous form. At the lowest V_{sg}, a very regular train of slugs was produced, with some froth remaining on the pipe wall in the gas bubbles between the liquid slug units. The coalescence of small droplets on the wall was not so apparent at these conditions mainly due to the fact that the very frequent slugs/pseudo-slugs gave little time for the coalescence to take place. It was observed that at these conditions the slug lengths were much more regular than was observed at the lower V_{SL}, and although no data can be offered to back up this observation, observation of one video test did suggest the slug length spectrum to be narrow. At the lowest V_{sg} it was observed that as water cut was increased, the liquid slug unit appeared to take on a progressively squarer profile, where the slug front and tail are sharper than was noted for the oil-only flow. Videos of the Oil No. 1 tests suggested the same character was obtained with the kerosene. At the higher V_{sg} values, a continuous foam/droplet coating was observed on the wall between the slugs at the end of the pipe. At this V_{SL} it was not possible to obtain the required V_{sg} to break up the slugs into a pseudo-slug flow for practical operational reasons mentioned previously.

To conclude, the main items of this investigation being the influence of water on bulk oil/gas flow regimes, oil-water mixing effects and the effect of increased oil viscosity, the following comments are made:

i Maintaining a constant V_{SL} suggested that the introduction of water had no observable effect on bulk gas/liquid flow regime.

ii At low liquid superficial velocity, gas shear has a dominant effect on the extent of oil-water mixing.

iii The formation of slugs does not necessarily mean the oil and water will be well mixed.

iv Pseudo-slug flow was seen to persist at the vast majority of high gas superficial velocity conditions.

v Increasing gas superficial velocity promotes foaming.

vi The increased oil viscosity has had a small effect on the flow rates required to homogenise the oil and water.

6.5.1.2 Oil No. 1 Slug Characteristics Data

Collection of slug characteristics data was made only at the lowest gas superficial velocity. This was necessary due to the fact that the video technique was not considered sufficiently accurate at higher V_{sg}. The analysis was performed on a video editing suite, at 25 frames/second, where the target area on the pipe was 1m long. For V_{sg} = 1m/s, this gave a residence time of typically 1 second or more, which was considered adequate. For slug length determination, where the residence time is converted to a length, the slug body has been assumed to move at the mean superficial mixture velocity.

Examination of Figs 6.35 to 6.38 reveals that at moderate V_{SL}, the 100 slugs counted gave a fairly broad range of slug length from less than 1m to over 3m. This is indicative of the case that at this superficial liquid velocity range regular slug flow has not yet been established, and occasional large slugs are the result of a sweepout action where the stratified liquid level builds up and, at a particular location, the slug is "triggered". The figures also suggest that the λ_w = 0.5 case gives a slightly broader slug length spectrum than obtained for the dry oil case. Fig 6.36 shows the average slug length, where for the lower V_{SL} values the water-cut cases produced slightly higher mean slug lengths. At V_{SL} = 1.67 m/s, a regular train of slugs is produced, and the average lengths for different λ_w appear to be very similar. The maximum slug lengths ranged from just over 70 pipe diameters in the low V_m tests to around 40 diameters at the highest superficial mixture velocity.

Fig 6.33 shows the average slug-front velocities, $\overline{V}_s$, in relation to V_m. The slug front velocity is commonly expressed as a multiple of the superficial mixture velocity, and the figure illustrates that at low V_m, $\overline{V}_s$ approaches $1.5V_m$ for the λ_w = 0.5 case. The difference between the λ_w = 0.5 and dry oil cases is not large, however. It should be mentioned that, since the oil-only slug tends to have a front which is less well defined than the water-cut case, there is present an unavoidable element of subjectivity in the video analysis. As V_m is increased to the value associated with very regular slugging, $\overline{V}_s$ approaches $1.2V_m$, which is consistent with many experimental and theoretical findings present in the literature.

Fig 6.34 depicts the slug frequency which is easily obtained from the video material. At the lower values of V_m, slugs are produced slightly more often in the dry oil case than with the water-cut situation. This observation is consistent with the slightly shorter average slug lengths of the dry-oil tests. From mass-balance considerations, if the holdups in the slug and film are similar in each case, the production of more frequent slugs requires that the slugs are on the average slightly shorter for a constant film height. At V_{SL} = 1.67m/s, the very regular slug production results in the λ_w = 0.4 and 0.7 tests giving very similar data, consistent both with the length and front velocity data of previously.

In conclusion, these slug tests have enabled a preliminary investigation of the effect of water cut on oil/gas slug flow phenomena. The data suggests that water cut does have a small effect, albeit a consistent effect in these tests. However, an extended examination of slug flow was not possible within the study framework, although the work did allow several recommendations to be made which are contained in the closing chapter.

6.5.2 Liquid Holdup Results

6.5.2.1 Oil No. 1 Data

Fig 6.15 reveals that, for V_{so} = 0.1m/s, there is no huge differences between the measured holdups throughout the water cut range. Unfortunately, the scatter in the data can point to no definite trend. Note that the V_{sg} on these graphs is the gas superficial velocity corrected to a point midway along the 10m holdup measurement section. The data suggests that, despite the increased liquid input as water cut is increased, the in-situ fractions are fairly similar across the V_{sg} and λ_w range. This can partly be explained by the water causing a lower liquid slippage, as has obviously occurred at this low V_{so}. A V_{so} of 0.15 m/s, Fig 6.16, shows significantly higher holdups for the λ_w = 0.7 situation. This arises due to the formation of more regular slugs at this condition, and the presence of liquid slugs in the pipeline (and so in the measurement section) contributes to a higher liquid fraction in-situ. The other water cut holdups are increased, but the behaviour is somewhat erratic. It is in this region that most uncertainty exists as to the reliability of the trap-and-bypass method of discrete samples used here, and this point is taken up later in this section.

As V_{so} is increased to 0.3m/s and higher, Figs 6.17 to 6.19, similar behaviour is observed where there is a rapid decrease in holdup up to V_{sg} = 4m/s, and more gradual decreases at higher gas superficial velocities. The general trend is for a range of holdups to be produced at each V_{sg}, where the λ_w = 0.7 produces the highest holdups.

This latter point is expected in slug flow, but as mentioned did not carry in the lowest liquid velocity tests. The general scatter in the data as regards to trends with particular water cuts, remains in many of the tests.

An advantage of using the adopted technique was that it was possible to recover in-situ water fractions. After switching the valves and draining to the perspex suction pots, after at least 5 minutes most tests meant that adequate oil-water separation had taken place to determine the water cut of the trapped liquid. Figs 6.20 to 6.24 show this data against the no-slip water cuts λ_w. Where inadequate separation occurred the data is not presented. The data reveals that in most cases the in-situ water cut was similar to the no-slip value. This behaviour agrees with the results of the oil-water tests, where no oil-water slippage was obtained in the horizontal pipe across a range of flow patterns. It is accepted here that an oil-water flow has been established before gas is added to the mixture, but nonetheless, the presence of gas in varying quantities has not had the effect of causing significant slip of the oil over the water.

At this point it is pertinent to discuss the slug holdup sensitivity tests previously mentioned. These tests were run to obtain an estimate for the acceptability of taking 3 samples for slug flow average holdup determination. Where a continuous sampling method, such as nuclear densitometer, is not feasible the only way to achieve a fully acceptable result is to take a large number of samples or to trap the entire pipeline volume. The former approach was used where 10 samples were involved.

Figs 6.41 to 6.46 show that, as expected, a range of average holdups are recorded in each case. At the lower V_{SL} of 0.3m/s, the fluctuations are greater which is indicative of a more irregular slug flow where the slugs are fairly infrequent. In one of these cases, Fig 6.42 the average obtained from 10 samples differs markedly from that obtained from the earlier test involving only 3 trappings. However, in the other cases the agreement is acceptable, including the dry-oil test at V_{SL} = 0.3m/s. As regular slug flow ensues at progressively higher liquid superficial velocity, the variation in average holdup becomes less significant, and so the result from 3 samples would be expected to be satisfactory.

In short, this brief series of tests has indicated that there is some uncertainty in the holdup measurement at low V_{SL}, but the data is considered to be basically sound. At higher V_{SL}, these tests suggest that an acceptable value of average holdup in slug flow will be recorded. It must also be mentioned that, with an element of subjectivity always present, care and attention by the operator to achieving a representative sample was a key role.

6.5.2.2 Oil No. 2 Data

The way in which the lubricating oil tests were run enabled direct comparisons between tests where the liquid throughput was the same, gas input equal, and the only variable being the input water cut. Figs 6.58 to 6.62 demonstrate that, in general, whilst there is some scatter in the data, similar holdups are obtained for the range of λ_w involved. Examination of the data reveals that no particular water cut was associated with holdups significantly removed from the dry-oil run . This lack of any clear trends suggests the span of holdups is due in part to the uncertainty of the measurement method, discussed in the previous section. The result of a significantly higher or lower holdup in a horizontal pipe would arise principally from:

i Alteration of fluid physical properties

ii Change of flow regime.

Dealing with the first point, the most important change in the liquid property would be to either increase or decrease the viscosity. Addition of water could theoretically, as a worst case, produce oil-water dispersions of high viscosity if enough energy is imparted to the mixture. Increased liquid viscosity is associated with increased liquid holdup due to higher slippage of the gas phase over the liquid.

Alternatively, the effect of introducing water can lower the effective liquid viscosity and so reduce the gas-liquid slippage. However, the data does not support either argument. It would appear that to substantially change the holdup a significant departure in liquid viscosity is required, which would also have repercussions to the pressure drop data, discussed later.

So far as the second point is concerned, the bulk flow regimes did not appear to be affected by the water fraction in the liquid stream, and so this too suggests little effect on the liquid holdups.

Examination of Figs 6.63 to 6.67 shows that there was little slippage between the oil and water for all flow conditions, regardless of water cut, gas rate or prevailing flow regime. This observation is in agreement with the oil-water data, as well as with the Oil No. 1 oil-water and oil/water/gas data. Therefore, in no tests involving flow in the horizontal pipe was significant slippage obtained between the oil and water. Despite the fact that it is unknown how far this trend would carry in terms of higher oil viscosities, it is an important result regarding any modelling of the flow, since it means that the oil-water interaction terms can be written in no-slip form.

6.5.3 Three Phase Pressure Loss

6.5.3.1 <u>Oil No. 1 Tests</u>

Inspection of Figs 6.25 to 6.29 shows that, generally, pipe pressure drop was increased as more water was added to the flow at constant gas/oil ratio. The pressure losses for the dry oil and λ_w = 0.2 are in most cases very similar, and this is indicative of only slightly greater liquid throughput. The addition of water results in the oil phase occupying a smaller volume of the pipe, and continuity dictates that the oil phase will then travel faster through the reduced cross-sectional area. In the stratified case, the water in contact with the pipe wall will have an associated friction loss which is smaller than that due to the oil. There is then competing forms in terms of pressure drop increase/decrease. It appears that the 20% water addition has had negligible effect on the pressure drop, though this is probably due to the similarity in viscosity of the oil and water. Had the oil been very viscous, where laminar flow would be expected to dominate, the effect of water may have been greater.

It is also important to note any effect of flow regime. Low velocities give a three-layer stratified-type flow. The production of liquid slugs suggests that the character of the slug is important in determining the pressure drop: in some cases the slug/pseudo-slug is oil-water homogeneous. It is interesting to note that, for the $\lambda_w = 0.2$ tests, even at the highest V_{SL}-V_{Sg} combination, the pressure drops were similar to the oil/gas result.

As expected, the pressure drops at higher λ_w were much removed from the $\lambda_w = 0$ data. At $\lambda_w = 0.5$, there is then twice as much liquid travelling through the pipe as was the case for $\lambda_w = 0$. In many cases, the $\lambda_w = 0.4$ data is closer to the $\lambda_w = 0.2$ results than the $\lambda_w = 0.5$ data. This result is interesting in that it indicates, for a pipe carrying a certain flow of oil, in several cases the pressure loss is not greatly changed by the addition of up to 40% total flow water. However, at the highest V_{so}, this observation does not hold. At sufficiently high flow rates, particularly of the gas phase, the oil and water form a dispersion where the stability and state of the dispersion is of importance. This will be discussed later.

Finally, at $\lambda_w = 0.7$ in all cases the pressure drop was significantly higher than for all the other water cuts. Again, this result is not unexpected due to the fact that in this case the pipe is carrying over 3 times as much liquid compared to the respective $\lambda_w = 0$ case. It is considered here that this large increase in liquid throughput will mask any features due to dispersion/rheology/foam-flow effects for this oil viscosity.

It is difficult to draw any firm conclusions from the data discussed above. Better insight is available if one considers the total liquid throughput fixed, and the input water cut the variable. It was possible in some of the Oil No. 1 cases to extract this information from the tests. Figs 6.30 and 6.31 show the pressure loss at 3 water cuts for $V_{SL} = 0.5$ and 1.0 m/s respectively.

Examination of these graphs immediately provokes more thought as to the oil/water/gas pressure loss behaviour. For the $V_{SL} = 0.5$ m/s case, the 40% water cut, at the higher values of V_{sg}, results in a significant decrease in pressure drop. This behaviour is especially important due to the fact that it ties in very closely with what was observed on the Oil No 2 oil-water pressure loss data of Chapter 5. In Fig 6.30, the principal effect of gas flow seems to be from approximately V_{sg} = 3m/s, and even at the highest V_{sg} the λ_w = 0.7 pressure drop is only slightly above that for zero water cut. At V_{SL} = 1m/s, Fig 6.31 shows that, except for the highest V_{sg} case, the 0, 50 and 70% water cuts gave similar pressure drops. However, at the highest V_{sg}, there is both a marked decrease in pressure drop at λ_w = 0.5, and a pressure loss at λ_w = 0.7 which is somewhat above that of the oil/gas result.

In summary, the Oil No 1 tests have shown that increased pressure drop is generally not attained until λ_w exceeds 0.2 (for a constant oil flow), and that this is insensitive to gas superficial velocity. Results have been re-cast in V_{SL} - V_{sg} format in a small number of cases. These results suggest that there is a major effect of water cut in several instances, and that future tests carried out in this manner would yield more useful information.

6.5.3.2 Oil No 2 Data

As mentioned previously, the Oil No 2 tests afforded the opportunity of conducting the tests at constant gas/liquids ratio. Figs 6.68 to 6.72 display the pressure drop data for each set V_{SL}.

Turning to the lowest V_{SL} = 0.1m/s, Fig 6.68, firstly it must be conceded that at such very low pressure drops, the measurement uncertainty is considerably higher than for the other V_{SL} cases. Nonetheless, there does seem to be a trend of slightly lower losses as the water cut is increased to a value of about 40%, with a slight increase in pressure drop at higher water cuts. This trend agrees with that noted in the Oil No 2 oil-water tests (V_{SL} = 2m/s only) and the Oil No 1 constant-GLR data. At V_{SL} = 0.1 m/s, the flow regime in most cases was well-separated or semi-homogeneous. This suggests the pipe is conveying a separated oil-water mixture or oil- (oil-in-water) emulsion mixture at λ_W up to 0.4. Higher λ_W, however, suggests a mixture of oil - (water-in-oil) emulsion is being transported which should produce slightly higher friction losses, although admittedly this is also a function of the tightness of the dispersion.

Similar trends were obtained for the V_{SL} = 0.3m/s data. At the lowest V_{sg}, where regular slug flow was observed, a fairly flat dp/dx - λ_W relation was observed.

Increase of V_{sg} again results in a minimum in the pressure loss at the λ_W =0.4 - 0.5 region, with the depth of the minima apparently increasing at higher gas superficial velocities. In all cases, the pressure losses at λ_W = 0.7 were comparable to the λ_w = 0 tests, giving no apparent peak in the pressure drop characteristic.

At V_{SL} = 0.5m/s, similar trends were observed to prevail. At this point, a detailed analysis of the data was made to determine any operational factors which could be contributing systematically to the pressure drop trends. As outlined in Chapter 4, input gas flow rate measurement is at or near pipeline conditions so a correction is made for the expansion effect downstream to give V_{sg} values for the end of the steel pressure drop measurement section. It was found that the variation in V_{sg} across the different water-cut series for a fixed V_{SL} was no more than 5% and in most cases was well below 3%. Also, there was no trend in the V_{sg} for λ_W = 0.4 being lower than that for the other water cuts. Therefore, the possibility that variable V_{sg} is responsible for the variation in pressure drop is considered remote. The other factor that was examined was any effect of variability of the input oil viscosity. As Fig 4.1 shows, there is no temperature control on the facility except for heaters in the oil and water tanks used only during cold winter periods. There is then no control over the oil input temperature due to variability of temperature in the facility building, and hence no close control over oil viscosity. However, on examination of the oil viscosities logged during each test series, it is considered highly unlikely that any effect on the pressure drop data has been realised.

The V_{SL} = 1m/s data continued to display the same trends as previously, with minima in the pressure drop occurring at λ_W = 0.4. Whilst not undermining the consistency of this trend, it should be noted that in the vast majority of cases the pressure loss was around 20% less than the respective dry-oil pressure loss. Also of note from this data set is that the λ_W = 0.7 pressure loss, in relation to the λ_W = 0 value, is sensitive to the gas superficial velocity. Examination of Fig 6.72, where V_{SL} = 2 m/s, also shows similar trends although the minima appear to be closer to the λ_w = 0.2-0.3 band.
With several trends apparent, it was decided to spend more time to look closer at the pressure-drop characteristics. A number of tests were run, at V_{SL} = 1m/s, to establish the pressure drops at water cuts different from those examined previously to give a more discrete dp/dx - λ_w relationship. Two values of V_{sg} were involved, and no holdup data was collected in these tests. The results are depicted in Figs 6.73 and 6.74. At the lower V_{sg}, increasing λ_W results in a decreasing dp/dx to around λ_W = 0.4, as obtained previously. At λ_W = 0.4 - 0.5 an increase in pressure drop is obtained. This, coupled with Fig 6.30, displays repeatable trends showing that a marked pressure loss characteristic is obtained with this three-phase system.

Consideration of all the oil/water/gas pressure drop data must also be linked to the oil-water pressure loss results. The oil-water data at V_{SL} = 2m/s displays identical trends to that observed for the oil/water/gas data in terms of water cut effect. Oil-water flow at V_{SL} = 2m/s was observed to be fully homogeneous. Most of the three-phase tests at the higher values of V_{sg} also resulted in full oil-water mixing, with an increase in V_{sg} i.e. mixing energy, promoting higher friction losses when compared to the oil/gas data. Similar trends were obtained where only a semi-homogeneous mixture was involved. No method of measuring droplet sizes was considered practical, and so there was no way of determining the tightness of the oil-water dispersion.

Following the success of using a conductivity probe in the oil-selection tests, Appendix A, a similar probe was mounted in the test loop a short distance downstream of the pressure drop measurement section. Unfortunately, experience proved the instrument was not sufficiently robust for the task and no data was collected.

The Oil No.1 data presented in constant GLR form supports the Oil No.2 observations, although it is admitted that given the amount of kerosene data available for correct comparison means that this conclusion is somewhat speculative.

6.6 CONCLUSIONS

The results of the horizontal three-phase tests allow the following preliminary conclusions to be made:

i) Bulk oil/water/gas flow regimes were observed to be similar to oil/gas flow regimes where liquid superficial velocity is held constant.

ii) Oil-water mixing was observed to be considerably influenced by gas superficial velocity.

iii) Average liquid holdup shows no systematic sensitivity to water cut in a constant superficial liquid velocity flow for the velocities studied.

iv) Oil-water behaviour within the three-phase flow is of a no-slip nature.

v) Pressure drops were found to give a slight minimum at approximately 40% water cut.

vi) No peak in pressure drop due to dispersion phenomena was obtained.

vii) The observations above are consistent for two oils of different viscosities.

CHAPTER 7

OIL/WATER/GAS FLOW IN A PIPE WITH SMALL INCLINATION

7.1 INTRODUCTION

The two previous chapters have been concerned with multi-phase flow in a pipe set at the horizontal configuration. In the general field of multi-phase flow it is true to say that the majority of both experimental and theoretical studies have been involved with flows either in horizontal or vertical pipes, with very little studies being made of inclined flows, especially at low inclination. This is not because such flows are of no practical interest, but chiefly due to the complexity of the subject necessitating that work first concentrates on the possible extremes. The natural starting point for a flowline study is to examine a horizontal flow, and for a well flow the examination of vertical flow is a logical first approach. Most pipelines are not truly horizontal but slightly inclined, even if only to a fraction of a degree in some instances. Also, progress of drilling technology has enabled engineers to complete highly-deviated production wells and, recently, horizontal wells. So far as the pipeline problem is concerned, many multi-phase research studies are now aimed at predicting the effect inclination will have on the fluid mechanics and operability of a system. In this study, the nature of the problem has once again dictated that most effort has been centred on investigation of horizontal flows. However, the opportunity was taken to conduct a series of tests where the pipeline was inclined at an inclination of 1-degree. This allows the effect of pipe inclination to be assessed, even if only at a preliminary level, by comparison with the horizontal data for a similar system. The author is not aware of any study of the effect of pipeline inclination on oil/water/gas flow available in the literature.

7.2 OBJECTIVES

The objectives of this series of tests were as follows:

i The collection of flow regime information in an uphill-downhill system.

ii The collection of liquids holdup data in an uphill flow.

iii The collection of pressure drop data in uphill and downhill three-phase flow.

iv A preliminary investigation of slug flow in uphill and downhill oil/water/gas flow.

v Comparison of the data to previously obtained horizontal data where appropriate.

7.3 FACILITY CONFIGURATION AND DATA ACQUISITION

7.3.1 Test Loop Configuration

As was mentioned in Chapter 4, the design of the facility was undertaken with flexibility in mind regarding being able to study non-horizontal flows. Although it would have been possible to set the loop at an inclination of 2 degrees, discussions with the research sponsors indicated that an inclination of 1-degree would be of interest. The far end of the structure was raised using a hydraulic jack, and the nuts locked in position and mid-point supports set to give an angle of 1-degree, which was set using a laser and clinometer device concentrating on the pipe loop slope. The resulting configuration, depicted in Fig 7.1, means that the flow effectively encounters a 'hill-dip' configuration where the outward 550 pipe diameters are uphill and the return 550d is downhill.

7.3.2 Test Loop Data Acquisition

The data acquisition system was based on that for the horizontal experiments with a few modifications. First of all, it must be admitted that the configuration only allowed liquid holdup to be measured in the uphill leg. Whilst it would have been desirable to install a similar pneumatic valve and bypass system at the end of the return leg the expense and delay which would have resulted rendered this possibility impractical. It was, however, required to measure pressure loss in the uphill and downhill legs. A 6m glass section of the uphill pipe upstream of the holdup measurement section was replaced with a smooth plastic section which also gave a visual capability. These instruments were the only additional inputs to the Tulip PC-based data acquisition system, outlined in Chapter 4. Slug flow information was obtained using the same video technique employed previously.

7.4 EXPERIMENTAL PROCEDURE AND TEST VARIABLES

The experimental procedure devised in Chapter 5 and continued in the Chapter 6 experiments was adopted for these experiments also.

The experiments were run with the kerosene, test Oil No. 1. The tests were set up in the same manner as the Oil No. 1 horizontal tests i.e. gas/oil ratio (GOR) was held constant as water was added to the flow to give set values of no-slip water cutλ_w. The same ranges in oil superficial velocity, no-slip water cut and gas superficial velocity were used as were employed for the horizontal experiments. Data was not collected for the V_{so} = 1m/s, λ_w = 0.7 case due to existing facility constraints, detailed previously. Oil viscosity over the period of the test programme was in the range 1.8 - 2.1 cP. Oil/air surface tension was 28.5 mN/m and oil/water interfacial tension was 32mN/m; these measurements were made at 20°C using the drop-volume method.

Test coding was similar to that used for the horizontal Oil No. 1 tests thus:

3 P 1 1 A A

Code	A	B	C	D
Target V_{sg} (m/s)	1.0	3.0	5.0	7.0

Code	A	B	C	D	E
λ_w (-)	0	0.2	0.4	0.5	0.7

Code	A	B	C	D	E
V_{so} (m/s)	0.1	0.15	0.3	0.5	1.0

Pipe Inclination 1°

7.5 DATA PRESENTATION

7.5.1 Flow Regime Results

Flow regimes were determined by visual observation and classified as for the horizontal studies thus:

A: Predominantly stratified flow where the liquid film consists of a separated oil-water mixture or an oil-dilute emulsion mixture.

B: Predominantly stratified flow where the film was observed to be a homogeneous mixture of oil and water.

C: Infrequent slug flow where the slug and film are of oil/water or oil/emulsion separated flow.

D: Regular slug flow with features as flow C.

E: Infrequent slug flow where the slug and film are a homogeneous oil-water mixture.

F: Regular slug flow with features as for flow E.

G: Infrequent pseudo-slug flow where the pseudo-slug and film are of an oil-water or oil-emulsion separated nature.

H: Frequent pseudo-slug flow where the oil and water display characteristics as for flow G.

J: Infrequent pseudo-slug flow where the pseudo-slug and film consist of a homogeneous oil-water mixture.

K: Frequent pseudo-slug flow with oil-water behaviour as for flow regime J.

The flow patterns as classified above are given on V_{so} - V_{sg} coordinates for each water cut λ_w, at observation locations at the end of the uphill and downhill legs, and are displayed on Figs 7.2 to 7.11. This information is also contained in Tables 7-1 to 7-5 for uphill flow, and Tables 7-6 to 7-10 for downhill tests.

7.5.2 Liquid Holdup Data

Total liquids holdup in uphill flow for each oil superficial velocity is plotted against gas superficial velocity as shown in Figs 7.12 to 7.16. These figures also depict several of the curves obtained for the same fluid system but in horizontal flow, although for the sake of clarity only the boundary water cuts are shown. Figs 7.17 to 7.21 display the in-situ water cuts, $(WC)_a$, in relation to gas superficial velocity and the no-slip water cut values.

7.5.3 Pressure Loss Results

Pressure drops measured in the uphill portion of the test loop are given in Figs 7.22 to 7.26. Pressure losses measured in the downhill leg for V_{so} = 0.1 and 0.15m/s were found to be very small and outwith the reasonable accuracy of the measurement system, but pressure drops for the higher values of oil superficial velocity are displayed in a similar fashion to the uphill data, in Figs 7.27 to 7.29.

7.5.4 Slug Characteristics

Video analysis was performed of slug flow at V_{SL} = 0.3m/s and 1.0m/s and at V_{sg} = 1.0m/s: several no-slip water cuts were involved. The videos were taken at the end of each section and involved the analysis of 100 slugs in each case. Figs 7.30 to 7.32 show the average slug length, slug front velocity and slug frequency respectively, for these cases. The data for the appropriate horizontal experiments are also included for comparison. Table 7-11 contains both the uphill and downhill characteristics data.

7.6 DISCUSSION OF RESULTS

7.6.1 Flow Regime Data

7.6.1.1 Uphill Flow

Figs 7.2 to 7.6 show that for flow in the uphill section intermittent-type flows were obtained in all test conditions. At the lower liquid and gas throughputs, the effect of the pipe inclination is very noticeable and the process of slug generation occured in distinct steps. Initially, a stratified-type of flow appears to be underway but gravity aids drainage of liquid down the slope, particularly at the lowest gas superficial velocity. This drainage was observed to increase local liquid levels to such an extent that the liquid film holdup was increased sufficiently to generate slugs which Taitel and Dukler (1976) suggest to be due to a mechanism known as Kelvin-Helmholz instability.

However, at the low flow rates, the liquid units produced were rather short slugs with the occasional roll wave and pseudo slug. The result is a somewhat chaotic situation where local liquid levels are building up, triggering the formation of slugs which in many cases were unstable to the extent that they had completely dissipated before they reached the return bend and hence downward inclination. To complicate matters, the produced slugs occasionally consumed liquid from dissipating slugs which had been triggered earlier. At these flow rates, the flow could be termed a slug-churn type of flow, as is observed in vertical flows, where slugs are formed but in many cases lead to a chaotic situation of production/dissipation/drainback through the uphill leg. The corresponding tests on the horizontal pipe, discussed in Chapter 6, did not lead to such regular intermittency in the flow. It is therefore correct to refer to the slugs in the uphill leg at the lowest V_{so} as being terrain-induced. This is a term often applied to a system where a downhill pipe precedes a vertical riser, but the description is no less valid in this instance. It is also of note that even at the highest superficial gas velocity intermittency was still present in the flow in the form of pseudo-slugs, although the terrain effect was still evident.

As liquid flow rates are increased the terrain-induced slugging was replaced by a more regular slugging. This is to be expected since the increased liquid holdup will aid the production of slugs more frequently, and as the liquid flow rate is increased the degree of liquid drainback decreases progressively as a more classical type of slug formation begins. Although the effect of pipe inclination and hence gravity would be expected to be present at all flow conditions, it was clear that at the higher liquid and gas superficial velocities the presence of an uphill incline was much less of a controlling factor as to the bulk flow structures. At the two highest liquid superficial velocities a regular train of short slugs was present in the uphill flow, with a character very similar to that observed in the corresponding horizontal experiments.

It was observed that the degree of oil-water mixing in the flow was generally similar to that observed in the horizontal tests. At the lower liquid throughputs the slugs/pseudo-slugs were usually of an oil-water or oil-emulsion-water character until the higher V_{sg} where sufficient agitation was available to homogenise the oil and water. At low liquid flowrates, the chaotic churning of the flow noted earlier would be expected to provide a better environment for improved oil-water mixing, but observations suggest that the degree of mixing was not substantially different from that seen in the horizontal tests where infrequent slug/pseudo-slug flow was obtained. It appears that the gas superficial velocity is a much greater factor in determining oil-water mixing at low superficial liquid flow rates than is the churning effect induced by the pipe inclination. As water cut and oil superficial velocity is increased, the effect of the pipe incline is reduced further and the oil-water mixing observations were found to be very similar to that for the corresponding tests involving the horizontal test pipe. It should be stressed that the degree of oil-water pre-mixing which is available in the oil-water supply flow in the gas-liquids mixer was identical for both the horizontal and inclined tests for the same flow rates, and hence any observable differences would have been attributable to the different pipe inclinations. However, it would appear that as with the horizontal tests the degree of oil-water pre-mixing is an important factor at low-to-moderate liquid flowrates.

7.6.1.2 Downhill Flow

Downhill flow resulted in a wider variety of flow patterns. Firstly, it should be stated that the downhill leg is not strictly a downward pipe in the sense that the flow is not downhill from the gas/liquids mixer: flows which are observed in the downhill leg have already been subjected to an uphill flow condition. The extent of this uphill flow influence is difficult to quantify, but in the following comments on this aspect are raised where appropriate.

An oil superficial velocity of 0.1 m/s produced predominantly stratified flow for all values of water cut. This is to be expected as a downward incline will enlarge the stratified flow region by reducing liquid holdup and delaying the formation of slugs. As gas flow is increased at this liquid rate, liquid is torn from the film and deposited on the pipe wall as a series of droplets which form streaks along the wall before returning to the liquid film. Slug flow from the uphill leg was observed to dissipate rapidly when encountering the downhill section, with slugs collapsing and consequently increasing the local liquid level for a short time.

As V_{so} is increased, intermittency appears in the flow in the form of infrequent pseudo-slugs, particularly at the highest superficial gas velocity. This could be an effect from the uphill flow regime: at high V_{sg} the uphill intermittent flow will be expected to be less affected by the action of gravity in the downhill section. There will then be a decreased possibility of slug collapse in the downhill leg, and also pseudo-slugs rather than slugs are expected at these velocity conditions. However,

at V_{SL} higher than about 0.5 m/s, regular slugs were observed in the downhill leg, with regular pseudo-slugs at the higher V_{sg} values. Examination of the flow regimes observed in the horizontal tests reveals that in those tests intermittent flow was obtained at lower V_{SL}: the enlargement of the stratified envelope in the current tests is a direct result of the 1-degree downward incline. It can be argued that the stratified region would be slightly larger if the uphill leg was not providing the input to the downhill section.
Oil-water mixing in stratified flow was found to be primarily dependant on gas superficial velocity. Progressively smaller values of V_{sg} were required to mix the oil and water as V_{SL} is increased and the flow approaches intermittent. This is to be expected since, at these conditions, slug flow in the uphill section tends to increase oil-water mixing, resulting in less gas being required to homogenise the mixture. The nature of the oil-water mixing in downhill slug flow was the same as that for uphill slug flow, where increased liquid rates aid oil-water mixing and gas input further mixes the oil-water combination.

7.6.2 Holdup Data

Liquid holdup, measured in uphill flow only, was found to exhibit a similar degree of data scatter as was obtained in the horizontal tests. However, there is a clearly higher holdup for the λ_w = 0.7 and 0.5 data than for the λ_w = 0 and 0.2 results, as would be expected due to the increased input (no-slip) total liquid holdup. It is also apparent that at V_{so}=0.1m/s and 0.15m/s, increased holdup was recorded for λ_w=0 when compared to the respective horizontal results: all the V_{so}=0.1m/s data followed this trend but the other curves have not been shown for the sake of clarity. The same is true for the comparisons with the V_{so}=0.15 m/s and 0.3 m/s data regardless of input water cut. This difference in holdup is to be expected given that, even with a relatively small slope of the pipe being present, action of gravity on the liquid phase and a slight bouyancy effect on the gas will contribute to more slippage between the liquid and gas. This is also responsible for the production of liquid slugs at lower liquid superficial velocities than required for horizontal flow, as discussed previously. However, as Figs 7.15 and 7.16 show, increasing V_{SL} tends to lessen the effect of inclination, and at the highest V_{so}- λ_w combinations the holdups recovered are similar to the results from the horizontal experiments, at least to within the accuracy of the measurement method.

Figs 7.17 to 7.21 show interesting trends have been obtained for the in-situ water cuts in uphill flow. At water cuts of 20% and 40%, the in-situ water cuts were seen to differ markedly from the no-slip values, and the effect becomes more signficant as the superficial gas velocity is increased. Unfortunately, insufficient time was available in this series to perform oil-water tests before proceeding to the three-phase tests, as was the approach with the horizontal tests. However, observations were made of the oil-water flows at all conditions before the gas was passed into the flow. At the V_{so} = 0.1m/s test conditions, it was clear that at low λ_w the in-situ water fraction in the oil-water flow was higher than the no-slip input.

Although this is based solely upon visual observation, it does indicate that considerable oil-water slippage can be obtained at low liquid flow rates and pipe inclination: significant oil-water slippage was not obtained in any of the horizontal experiments.

At input water cuts excepting $\lambda_w = 0.7$, the increase of oil-water slippage as V_{sg} is increased is consistent with the idea of an oil film being dragged over the separated water film by the action of gas shear, whilst the effect of inclination introduces a gravity factor which will also tend to increase the in-situ water volume. Again, this suggests an application for an idealised three-layer concept. In horizontal flow, the gas shear effect should be similar, therefore it is the introduction of the gravity factor which has resulted in higher in-situ water fractions. The relative absence of this effect at $\lambda_w = 0.7$ may be related to the much-decreased depth of the oil layer on top of the bulk water film, where it can be envisaged that, in the formation of slugs for example, a higher proportion of the water layer will be subject to the suction/shearing of the gas phase, and hence less water will slip from the oil layer.

The case of $V_{so}=0.15$ m/s gives results for $\lambda_w=0.2$ which are very similar to the $V_{so}=0.1$ m/s data. However, the data at higher λ_w, whilst consistently giving $(WC)_a$ greater than λ_w, fail to show clearly the previous effect of gas superficial velocity. At $V_{so} = 0.3$ m/s however, the V_{sg} effect is evident at $\lambda_w=0.2$ and 0.4, suggesting there is a small amount of scatter in the data. However, again the $\lambda_w=0.7$ data exhibits no significant oil-water slippage.

As oil superficial velocity increases to 0.5 m/s, the only oil-water slippage of note is at $\lambda_w=0.2$, the other in-situ water cuts being very close the no-slip values. The continuous reduction in oil-water slippage as liquid superficial velocity is increased may also be a function of the degree of oil-water mixing, where one would expect the three-layer concept to reduce ideally to a two-layer, gas-liquid problem. Finally, at $V_{so} = 1.0$ m/s results indicate that there was no slippage between the oil and water for any of the input water cuts. Therefore, only at the highest oil superficial velocity tests the situation was comparable, in terms of oil-water slippage, to that in a horizontal pipe.

It was possible to make a small number of comparisons where the V_{sL} is fixed and various λ_w are involved. Fig 7.33 shows the $V_{SL}=1.0$m/s situation. This data indicates that at low V_{sg} similar results were obtained for $\lambda_w=0,0.5$ and 0.7. However, as V_{sg} is increased the difference in the results also becomes more marked with the $\lambda_w=0.5$ results being consistently higher than data for the other λ_w, even if only by a small amount in terms of magnitude. Fig 7.34 shows that at $V_{SL}=1.67$m/s, the second-highest V_{SL} involved, the holdups for $\lambda_w=0.4$ and 0.7 are very similar. At this value of V_{SL} little oil-water slippage is expected, from the previous discussion.

Therefore this brief comparison suggests that oil-water slippage may contribue slightly to higher liquid holdup in the inclined pipe. However, it is admitted that the detailed features of the flow regime structure may also play a large part, although this was outside the scope of the current study.

7.6.3 Pressure Drop Results

7.6.3.1 Uphill Flow

Pressure loss for all uphill conditions is reported in Figs 7.22 to 7.26. At the lowest V_{so}=0.1 m/s, pressure loss generally increases as λ_w is increased, as before. However, for all the λ_w except for λ_w=0.7, a peak in the pressure loss is obtained and a subsequent flattening of the pressure loss results as superficial gas velocity is increased above 6m/s. Although at these very low flow rates the relative measurement uncertainty will be increased, further tests suggested that this characteristic was repeatable. A possible explanation for this lies in the fact that as V_{sg} is increased, whilst the frictional pressure drop may increase the hydrostatic pressure loss component reduces somewhat due to the accompanying reduction in pipeline liquid holdup. At V_{so}=0.15 m/s, Fig 7.23, this effect was not as significant, but a slight flattening is again obtained at V_{sg} greater than 6m/s: consistent with V_{so} = 0.1m/s data, no such result was obtained at λ_w = 0.7. At higher values of oil superficial velocity, the above effect had disappeared completely, suggesting that the much-increased friction-loss effects have rendered any elevation-component effects negligible. At V_{so}=0.3, 0.5 and 1.0m/s, regular curves are obtained at the different λ_w with increasing water cut giving progressively higher pressure drops as was obtained in the horizontal tests.

As with the horizontal data, it was possible to extract results to indirectly give comparisons for different λ_w and constant superficial liquid velocity. Fig 7.35 which is for V_{SL}=1.0m/s, shows that the pressure losses for λ_w=0.5 were consistently higher than the λ_w=0 and 0.7 values. This is supported by the holdup result of Fig 7.33 which suggested slightly higher holdups at λ_w=0.5, and this may be more important than any oil-water dispersion aspects. At V_{SL}=1.67m/s, Fig 7.36, the λ_w=0.4 pressure losses are consistently higher than the λ_w=0.7 values. Now Fig 7.34 suggests that the total liquids holdup for these flow conditions are very similar. At this flow condition, the oil and water form a mixture which was observed to be fully homogenous. It is suggested here that the difference in pressure losses lies in the exact state of the oil-water mixture in terms of droplet size and droplet size distribution which in turn is dependent upon the mixing intensity. The inclined pipe may be expected to complicate matters further as regards oil-water mixing/settlement/coalescence, but this is impossible to quantify within the scope of this work.

7.6.3.2 Downhill Flow

For the tests involving $V_{so}=0.1$ and 0.15m/s the pressure losses measured were very small and insufficient confidence existed in the pressure drop measurement therefore this data has not been presented. However, at higher V_{so} pressure drops were of a magnitude considered acceptable as regards measurement confidence. Fig 7.27 for $V_{so}=0.3$m/s, shows that $\lambda_w=0$, 0.2, 0.4 and 0.5 gave very low pressure losses at the lowest value of V_{sg}. This data possibly indicates that at this low gas superficial velocity, a recovery in the hydrostatic head component of pressure drop is being obtained in the downhill flow. As V_{sg} is increased the data then tend to separate into curves for each λ_w, with the $\lambda_w=0$ and 0.2 results being very similar, as was obtained in the horizontal tests. Although not shown on this graph, comparison with the $V_{so}=0.3$ m/s, $\theta=0$ results showed little difference in pressure loss, especially at the higher gas superficial velocities. This indicates that, at sufficiently high flow rates, the frictional pressure drop is the dominant component. However, this is only true where the flow regimes in the pipe are comparable, and this was indeed the case.

Fig 7.28, for V_{so} = 0.5 m/s, shows the same type of trends as obtained with the horizontal tests, and the values obtained were in general very similar to those measured in these earlier experiments. Finally, at $V_{so}=1.0$ m/s, Fig 7.29, similar trends to those discussed above are again obtained.

Comparisons are given for the downhill pressure losses at $V_{SL}=1.0$m/s and 1.67m/s in Figs 7.37 and 7.38 respectively. Fig 7.37 shows similarity with the curves obtained from horizontal flow, both for the Oil No. 1 and Oil No. 2 cases. Pressure losses at $\lambda_w=0.5$ are considerably below those at λ_w = 0 and 0.7 at the highest superficial gas velocities.

Therefore the slight pressure drop reduction in three-phase flow may not be confined only to horizontal pipes, but is possible in downward-inclines also. Fig 7.38 reinforces this statement: pressure drops at λ_w = 0.4 were found to be consistently below those where λ_w = 0.7. Although there is very limited data on which to form any strict conclusions, it is interesting to note here the reversal of the situation present in the uphill tests, Fig 7.35, where the $\lambda_w=0.4$ pressure drops were consistently above the $\lambda_w=0.7$ data. This suggests uphill flow is somewhat more complicated than either horizontal or downhill flow for a wide range of flowrates, whereas at this low inclination, at a certain magnitude of V_{SL} the downhill pressure loss characteristic becomes very similar to that produced in a horizontal pipe. However, the extent to which these trends carry to higher pipe inclinations is unknown.

7.6.4 Slug Characteristics Data

7.6.4.1 Uphill Flow

Fig 7.30 shows that the average slug length at V_{SL} =0.3m/s in the uphill flow was considerably lower than observed for the horizontal tests. As mentioned in 7.6.1.1, at these conditions slug generation was observed to be influenced by a liquid drainback effect which raised local liquid levels and hence induced slugging. It appeared that whilst slugs were being produced more frequently, their average lengths were slightly shorter than those observed in the horizontal pipe where slugs were generated les frequently. At this superficial liquid velocity, there was no observable difference in the λ_w=0 and λ_w=0.5 data. It seems that oil-water slippage effects are of secondary influence as regards average slug lengths compared to the effect of gravity on the bulk liquid phase. Average slug length at V_{SL} = 1.0 m/s were also lower than for the horizontal tests, for both the λ_w=0 and λ_w = 0.5 cases. There were slight differences in the λ_w=0 and λ_w=0.5 data, but the differences are fairly minor with respect to the expected measurement accuracy at this superficial mixture velocity.

Examination of Fig 7.31 indicates that in the 1-degree uphill flow, the slug front velocities are only slightly lower than those measured in the horizontal system. This is in general agreement with previous two-phase studies (see Mattar and Gregory (1974), for example) which suggest that for inclinations of up to about 10-degrees from the horizontal slug translational velocity does not change by an appreciable amount. Finally, Fig 7.32 backs up the experimental observations that slugs tended to be produced more frequently in the uphill system than in the horizontal system with, as mentioned, the resulting slugs being slightly shorter on average.

7.6.4.2 Downhill Flow

At V_{SL}=0.3m/s, slug generation in the downhill leg was very infrequent and thus no slug characteristics data was collected. At V_{SL} = 1.0m/s, as shown on Fig 7.30, average slug lengths were lower than in the respective horizontal experiments, but were very similar to the data recorded from the uphill pipe. This is indicative of a stable situation where no noticeable slug growth is occurring between the mid-point and end of the pipe, but the effect of the uphill inclination on the resulting downhill slug characteristics should be borne in mind. The uphill flow on the outward leg will, being 550 diameters long, set up slug flows of certain characteristics which, depending upon the relative extent of slug stability and the downhill inclination, may not vary through the length of the downhill pipe. At low superficial liquid velocities, in many cases the downhill section provided sufficient scope for gravity to overcome the slug stability and hence slugs in general would disintegrate in their downhill passage. At higher V_{SL}, the slug stability is expected to be higher, given a fixed gas superficial velocity, and hence gravity will have less effect on the overall slug hydrodynamics. However, it should not be adjudged from this data that at a $V_{SL}=1.0$m/s will produce similar slug lengths in all cases: the effect of a longer downhill pipe may well be to vary slug lengths considerably. The required pipe length to attain any degree of flow stability is of course a much-discussed topic, especially in multi-phase flow. The question is considered to be even more important as regards slug flow parameters.The length of the test loop involved here is similar to, if not considerably greater than in some cases, other test installations where similar data has been collected.

Figs 7.31 and 7.32 reinforce the slug length observations in that slug front velocity and slug frequency were similar in the uphill and downhill situations. It should be noted that very little slug characteristics data has been reported for inclined multi-phase flows, especially in the case where the flow is downhill.

7.7 CONCLUSIONS

The experimental data from these inclined-flow tests allow the following comments to be made:

i The 1-degree incline has resulted in an expansion of the slug flow envelope uphill and an expansion of the stratified region in downhill flow, which is consistent with gas-liquid data.

ii The V_{sg} required to homogenise the oil-water mixture was observed to be little affected by the inclination involved.

iii Liquid holdup was increased slightly when compared to the horizontal data, but as liquid flow rates increase the difference is diminished.

iv Oil-water slippage was obtained for V_{SL} below 0.5 m/s in uphill flow.

v Limited comparisons for constant V_{SL} but different λ_w yielded different trends for the uphill and downhill pressure drop data.

CHAPTER 8

DATA TREATMENT AND DISCUSSION

8.1 INTRODUCTION

The preceding chapters have covered in detail the study activities from the point where an experimental facility was required to the stage where data was collected and presented for discussion. This exercise enabled convenient comparison between numerous data sets and helped to establish several fundamental trends in terms of the behaviour of oil-water, oil/gas and oil/water/gas flows. Additionally, a common purpose of the study was to establish to what extent a modified two-phase (liquid/gas) flow approach could be employed to give predictions for the three-phase data. It was also considered important to look closely at the small amount of data collected in earlier experimental and field systems. With this base of experimental knowledge, another exercise may be performed to examine the fundamental fluid mechanics involved and the interaction with other fluids-related aspects. It was not an objective to develop new design methods, but more to show the applicability or otherwise of the current approaches and provide suggestions for which future work could be fruitfully undertaken, and this is summarised in the closing chapter.

8.2. OBJECTIVES

The objectives for this part of the study were:

(i) Comparison of flow regime, liquid holdup, pressure drop and slug characteristics data with predictions from several modified two-phase flow models and correlations.

(ii) Discussion of previous data and their description by a modified two-phase flow approach.

(iii) Examination of possible three-phase flow regimes, their liquid holdup and pressure drop characteristics and the role of the oil-water modelling.

8.3 DATA TREATMENT BY A MODIFIED TWO-PHASE FLOW APPROACH - CURRENT DATA

8.3.1. The Linear Mixing Rule

The use of a two-phase gas/liquid method for predicting three-phase flow effects immediately forces one to adopt a simplification which, at first glance, does not appear strictly justified. The two-phase flow correlations which are widely used by the industry are chiefly empirical and have been tested initially on data sets of relatively similar fluid physical properties. To produce, for example, a two-phase friction factor, requires as a part of the input data the fluid properties of the single phases. This, then, introduces bulk liquid and bulk gas property terms into equations. This cannot be expected to represent the microscopic interactions between the phases which may ensue upon mixing in a multi-phase flow. For representation of the liquid physical properties for the present three-phase data, the approach of using a "linear mixing rule" has been adopted. Whilst this is undoubtedly a simplified approach it has also been used since it is a useful first approximation in terms of data comparison: there would seem little value in expending effort on developing a more elaborate method unless comparison with the data suggests that such an exercise is warranted. The linear mixing rule is applied to the oil-water mixture to give 'averaged' liquid physical properties as follows:

$$\begin{aligned} \rho_L &= \rho_o(1-\lambda_w) + \rho_w \lambda_w \\ \mu_L &= \mu_o(1-\lambda_w) + \mu_w \lambda_w \\ \sigma_L &= \sigma_{o/g}(1-\lambda_w) + \sigma_{o/w}\lambda_w \end{aligned} \tag{8.1}$$

Therefore the average is made on the basis of the respective no-slip volume throughputs of the oil and the water. One would expect that the adoption of this approach for the density term has physical appreciation in terms of the bulk mass effects, whereas the latter properties can be closely tied to the microscopic chemistry of the oil-water mixture. Only at either ends of the water cut range - ie - clean oil or clean water, does the viscosity term receive strict representation. In the course of this chapter much reference will be made to the adoption of this simplified approach.

8.3.2. Correlations Used For Comparison

8.3.2.1. Flow Regime

In common with the multi-phase flow literature in general, many design methods are available which attempt to predict the prevailing pipeline flow pattern. As would be expected, the earlier approaches tend to be more reliant on empiricism than methods developed more recently, and many are relatively untested against reliable data from field-scale systems. A departure from this basically empirical approach was made by Taitel and Dukler (1976) who developed a method which embodies a significantly higher degree of physical and mathematical modelling. The original method was intended for horizontal and near-horizontal pipes, and is probably the most widely used flow regime prediction method openly available to the industry. For this number of reasons it was decided to test the method against the present experimental data. It should however be stressed that the model, in common with all other two-phase methods, does not attempt to describe the oil-water interactions or fluid chemistry aspects such as foaming and emulsion formation.

8.3.2.2. Liquid Holdup

Again, a large number of correlations are available to give predictions of average hold-up in multi-phase pipes. As it was the intention to take widely-used correlations for data comparison it was felt inappropriate to test a large number of holdup correlations and their relative performance. The widely used correlation of Eaton et al (1967) has been used for comparison purposes. This correlation was developed for flow in horizontal pipelines, but it was also tested against the inclined-pipe data of Chapter 7. However, it was considered worthwhile to compare the inclined-pipe data with a holdup method which could incorporate pipe inclination effects and for this reason the correlation of Mukherjee and Brill (1983) has also been used.

8.3.2.3. Pressure Drop

Of all the phenomena involved in two-phase flow, perhaps most attention has been focussed on the development of design methods which accurately predict the pipeline pressure loss, whether in vertical, horizontal or inclined flow. After over 20 years, methods still appear in the literature which prove to describe particular data-sets more accurately than previous correlations. Testing against data from field-scale systems is encountered much less frequently in reported studies, and the work of Gregory and Frogarasi (1985) provides the most comprehensive review published to date. It was decided in this work to use two pressure drop correlations for predictions. The first is that of Dukler et al (1964) which is an example of the early correlations. The second was that of Oliemans (1976) which field data suggests to give less conservative predictions than those from the earlier correlations. A holdup prediction must be used in conjunction with the pressure loss correlation; in this case the Eaton et al (1967) correlation was the logical choice following from the previous section, and to be consistent it was used with both the pressure loss prediction methods.

8.3.2.4. Slug Flow Characteristics

The collection of slug characteristics data was a secondary objective of the experimental work, but it was nonetheless considered valid to give predictions for slug frequency and average slug front velocity for comparison with the data. The commonly-adopted method of Gregory and Scott (1969) has been used for comparison with the experimental data.

8.3.3. Comparison With The Present Horizontal Data

8.3.3.1. Generation Of The Predictions

A FORTRAN computer program was written to enable convenient data input and the output of flow regime, holdup and pressure specific to the test loop geometry. The program consists of the two-phase flow methods of the previous section which were written around a flexible framework which enabled different methods of fluid property modelling to be input if desired. The following information was entered:

Inlet temperature, pressure
Air Metering Section temperatures, pressure
Oil Viscosity - Temperature curve
Oil Density - Temperature curve
Oil-Air Surface Tension
Oil-Water Interfacial Tension
Phase Input flow rates

This information represents the basic data required by the prediction program 3PHANL. Choices were then available to the user as to how the liquid physical properties should be represented in the correlations as follows:

Single-phase (oil or water) values
No-slip linear mixing rule
Woelflin (1947) correlation (for oil-water viscosity only)

In the current comparisons, only the no-slip linear mixing law was used. Output data was then generated for the specific correlations as required. The final comparison against the test data was made using hand-calculations.

8.3.3.2. Flow Regime

Examining first the Oil No. 1 data, Figs 8.1 to 8.5 show the flow regime data of Chapter 6 replotted with the transition curves from the Taitel-Dukler model included for comparison: only the curves for L = 1000d are given due to the observation that for the horizontal flow the pipe length was seen to modify only the oil-water mixing behaviour for some conditions, the bulk gas/liquid flow pattern prevailing at both L = 500d and L = 1000d. The transition stratified smooth - stratified wavy flow is predicted at the lowest V_{SG} and V_{SO} - λ_w combinations: it was obvious that in these tests, whilst the flow was of a more quiescent nature in terms of turbulence, wavy crests on the liquid film were observed at most test conditions.

The generation and character of waves on both a stationary and moving liquid film is a complex subject, and the pipe diameter is thought to be of some influence (Lin and Hanratty (1977)); this subject will be discussed later. The transition from stratified-wavy to intermittent flow is perhaps the transition which has received most attention from researchers, chiefly to help predict whether a pipeline will produce large liquid slugs which necessitate terminal processing capacity. As with all flow regime transitions, the transitions are not sharply-defined lines but rather a broad range where the flow characteristics contain aspects of each of the boundary flow patterns. At λ_w = 0 and 0.2, the lowest values of V_{SO} were associated with stratified or infrequent-slug flow patterns, where the latter can reasonably be said to be within the flow pattern transition range. When the input water cut has reached 40 to 50%, the frequency of slug production was continuously increasing but the Taitel-Dukler model predicted stratified wavy flow to persist to a higher V_{SO} than was actually observed. Only when λ_w = 0.7 does the model predict intermittent flow at all V_{SO} for the lowest superficial gas velocities, which is as was experimentally obtained. Given that this particular transition is a somewhat uncertain area, the model predictions are not unsatisfactory except in that they will give less conservative results in the sense that the required V_{SL} for slug flow will be over predicted rather than under predicted. As was mentioned in Chapter 6, intermittency in the flow was noted at virtually all test conditions, even if only in the relatively minor form of infrequent pseudo-slug production. No classical annular flow was observed in the experiments, and the intermittent-annular dispersed transition from the Taitel-Dukler model appears to be inaccurate for the current set of data. Annular-type flows were predicted to appear at approximately V_{SG} 4m/s or above at the low end of the V_{SL} scale, but this was not backed up by visual observation (and video/photographic examination). The modification proposed by Barnea et al (1982) to the original Taitel-Dukler intermittent-annular transition criteria is expected to produce more accurate results for the current experimental system. This modification takes the hold-up of the liquid slug into consideration, with the consequence being that the transition line is moved to the right of the flow regime map - ie - to higher gas superficial velocities. It is also worth pointing out that the original Taitel-Dukler intermittent-annular transition, for the present data set, describes quite well the change from slug flow to pseudo-slug flow at low-to-moderate V_{SL}. Therefore in this instance it can be of use in determining where the production of relatively large liquid slug units will tend to decrease.

Examination of the Oil N°2 horizontal data, Figs 8.6 to 8.10, suggests that success with the Taitel-Dukler predictions has been very similar to that obtained with the lower-viscosity oil. Regular slug flow was observed to occur at superficial liquid velocities above the Taitel-Dukler wavy-intermittent line, with either infrequent intermittent-type flows or stratified flows at the lowest value of V_{SL}. A similar extent of discrepancies between data and model for the intermittent-annular transition is obtained to that observed in the Oil N°1 tests. Again, the latter transition by Taitel-Dukler gives a good description of the slug - pseudo slug transition range.

Overall the comparison between flow regime data and the Taitel-Dukler predictions is quite encouraging: the extent of agreement for three-phase flow being similar to that for oil/gas. It should be noted that to effect substantial changes in bulk two-phase flow patterns, large changes in fluid physical properties are expected to be required. Many laboratory studies have been undertaken to determine the required departures in liquid viscosity which are necessary to radically shift the flow regime transition boundaries. Weisman et al (1979) is a good example of such work involving several of the flow pattern transitions.These workers found that for flows in a 51mm i.d. pipe, two-phase flow regime transitions were altered only slightly for the liquid viscosity range 1 - 150cP. Whilst this study was concerned with global changes, other studies have been made to look at particular transitions only. The effect of liquid viscosity on the stratified-intermittent transition has received attention, most recently by Andritsos et al (1989). The transition from stratified to intermittent flow was investigated experimentally for several liquid viscosities up to a maximum of 100cP. For viscosities up to 20cP, the transition is chiefly dependant upon the relative magnitudes of two factors; first, the extra liquid height required to produce Kelvin-Helmholz waves (stabilising effect) and, secondly, the increase in liquid height produced by the higher viscosity liquid for a given V_{SL} which is a de-stabilising effect. A small overall effect on the transition for viscosities 1 - 20cP is obtained, but at higher viscosities the second factor above outweighs the stabilising effect and so slugs are produced at a considerably lower V_{SL}.Data is presented for 1 and 4cP viscosity liquids, approximately corresponding to the viscosity range of the current study, and changes on all flow regime transitions were found to be negligible. This is supported by the present data, although it should be remembered that the purpose of introducing Oil No 2 was not to attempt to radically alter the bulk two-phase flow regimes, but more to show up any oil-water mixing phenomena. The fact that no noticeable change on bulk flow regime was observed on moving from oil-gas to oil/water/gas for both test oils would suggest that problems arising from the formation of the viscous dispersions were minimal, a point which will be taken up frequently in the course of this chapter.

8.3.3.3. Liquid Holdup

Both the Oil No1 and Oil No2 horizontal data were compared to the Eaton et al (1967) correlation. The results from the Oil No 1 comparison are depicted in Figs 8.11 to 8.15. Data points lying below the 1:1 correlation line represent the situation where the in-situ liquid fraction was underpredicted. As can be seen from Fig 8.11, at the lowest V_{SO} satisfactory agreement was only obtained at the lowest superficial gas velocity (so corresponding to the highest liquid holdup). As V_{SG} was increased, the liquid was found to be held-up to a higher degree than was predicted by the correlation. Similar observations were made by Minami and Brill (1987), who tested the same correlation against data from a 3-inch i.d. test line conveying air/water and air/kerosene mixtures.

Returning to the current data, as the oil superficial velocity, and so consequently the total liquid superficial velocity, is progressively increased the agreement between correlation and data improves markedly. This suggests that for this system the correlation is most appropriate at moderate-to-high liquid superficial velocities. Examination of the data did not reveal any trends with respect to particular water cuts giving a markedly better comparison with the predicted values, which may in itself be an important result.

When the correlation was compared to the Oil No 2 horizontal data, the results are as given in Figs 8.16 to 8.20. At V_{SL} = 0.1m/s, significant underpredictions were obtained at the higher values of V_{SG}, as was obtained with the kerosene-based data. Increase of V_{SL} to 1.0m/s and higher resulted in satisfactory prediction of the holdup for the majority of flow rates, although again no particular values of λ_w resulted in significantly better or poorer predictions from the correlation.

An estimation of the sensitivity to liquid viscosity of the holdup correlation can be made by comparing spot-values for the Oil No. 1 and Oil No. 2 data, oil-gas only, where the superficial velocities of each phase are sufficiently close for each oil. Although based on a small number of cases, the holdups predicted for the Oil No. 2 tests were typically less than 5% higher than for the corresponding Oil No. 1 inputs. Therefore, over the range of base-oil viscosity encountered in these tests, the Eaton et al correlation sensitivity to liquid viscosity is within the expected accuracy of the holdup measurement. Significant changes in liquid viscosity are required to significantly alter the liquid holdup; the data agreement with the correlation prediction indicates that sufficient increases in effective liquid viscosity did not arise in any of the tests.

Although every effort was made to preserve the interfacial tensions when changing from Oil No. 1 to Oil No. 2, it is worth briefly pointing out any sensitivities to the surface/interfacial tensions. The differences between the test oil interfacial tensions were 5% (surface) and 18% (oil-water interfacial). Little information, either in experimental or theoretical form, is available to assess the effect of gas-liquid surface tension on multi-phase phenomena in general. The aforementioned study of Minami and Brill (1987) included some investigation of the effect of gas-liquid tension on the liquid holdup, and the results suggested the influence of surface tension to be very small. A recent study by Hart et al (1989) also concluded that liquid holdup was relatively unaffected by surface tension over the range 38 to 72 mN/m.

Therefore these studies suggest that the tensions have been adequately controlled in the present experiments to minimise any influence on the bulk liquid holdup when comparing the Oil No 1 and Oil No 2 test data.

In summary, prediction of total liquids holdup in a horizontal three-phase flow was seen to be satisfactory at V_{SL} of 0.5m/s or higher, when an existing two-phase correlation was used incorporating the oil-water linear mixing rule modification, and for each of the test oils a no-slip assumption for the oil-water mixture has been found to be acceptable.

8.3.3.4. Pressure Loss

Prediction of Oil No 1 oil-gas and oil/water/gas pressure drops using the Eaton-Dukler method resulted in over predictions as shown in Figs 8.21 to 8.25. At the lowest values of V_{SO}, V_{SG} and λ_w, the measurement uncertainty is increased when compared to the higher flow rate tests which have significantly higher pressure drops, but the trends in terms of prediction success are nonetheless similar. Close examination of the figures reveals that for λ_w = 0.5 and lower, the measured values tend to be farther from the 1:1 correlation line than the corresponding λ_w = 0 and 0.7 values.

One important aspect is that the correlation is consistent in that overpredictions are obtained in all cases, and this is essential if it is required to validate the correlation against a set of high-confidence field data. Predictions using the Eaton-Oliemans method resulted in smaller differences between the predicted and measure values as can be seen in Figs 8.26 to 8.30. At V_{SO} of 0.1 - 0.3 m/s, the λ_w = 0.7 data resulted in under-predictions by the correlation at the highest V_{SG} values. At the highest V_{so} excellent accuracy is obtained for the low V_{SG} cases, but again as V_{SG} is increased the measured pressure drops exceeded those predicted by a considerable margin at the higher liquid superficial velocities. However, it can be said that for V_{SL} up to 1m/s and water cuts of 0.5 and lower, the predictive accuracy of the Eaton-Oliemans method was very satisfactory. Therefore within a particular flow rate and water cut range, the correlation can be of use for three-phase predictions, but outwith these constraints the existence of underpredictions does not encourage one to recommend the method as readily as when consistent over-predictions are the comparison result.

The pressure drop comparisons for the Oil No 2 data were deliberately configured in a different format to provide quantitative differences between predicted and measured values. The average error $\bar{e}$ has been used to provide this information and is defined thus:

$$\bar{e} = \left\{ \frac{\Delta P_{calc} - \Delta P_{meas}}{\Delta P_{meas}} \right\} x\,100$$

The results using the Eaton-Dukler method are depicted in Figs 8.31 to 8.35. As with the Oil No1 horizontal data. consistent overprediction of the pressure loss is obtained and the error magnitudes are in some cases dependant upon the no-slip water cut. This is a result partly due to the characteristic dip in the (dp/dx - λ_w) relation which was noted in Chapter 6, since the linear-mixing rule is not able to represent this phenomenon adequately. Comparisons using the Eaton-Oliemans method for the Oil No 2 data are given in Figs 8.36 to 8.40. The two lower values of V_{SL} result in slight overpredictions by the correlation, but at higher V_{SL} the effect of increasing V_{SG} on the predictive accuracy is clearly visible. At V_{SL} = 0.5 m/s a small number of under predictions are obtained, whereas at V_{SL} = 2.0 m/s all but 2 of the data points indicate underprediction by the correlation with the poorest result being an under prediction of over 30% based on the measured value. The Oil No 2 data then clearly supports the Oil No1 data in terms of the trends observed and the success of using the different pressure drop methods. Since the same holdup value - ie - that predicted by Eaton et al, is used for each method, it is the nature of the pressure drop correlations themselves which has produced the differences in the predictive characteristics. But it is of importance that for a large part of the flow rate envelope involved each of the correlations resulted in satisfactory performance when compared to the experimental measurement.

It is pertinent to comment here on the effect of the acceleration (kinetic-energy) component of the total pressure drop for the current system. This part of the total pressure drop is often neglected in field systems where the inlet and outlet pressures are relatively high and so the change in mixture momentum is not expected to be great. However, in low-pressure, especially near-atmospheric systems such as flare headers, this component is influential. The fact that a small back-pressure exists in the current system slightly reduces the effect, but the most important factor is that the maximum mixture velocities are moderate. To obtain an estimate of the acceleration losses, the E_k factor of the Dukler et al (1964) correlation was calculated for the highest V_m . The calculation predicts that the contribution to total pressure drop by the acceleration component is less than 2%. Using the Beggs and Brill (1973) correlation suggested that at the low end of mixture velocities, the acceleration component was within 1% of the total loss, and at the highest V_m this figure was about 5%. The recent study of Barua et al (1992) suggested that the estimates from the latter correlation may be very conservative. Therefore it is considered that in this study the acceleration pressure drop was relatively insignificant.

8.3.3.5. <u>Slug Characteristics</u>

Horizontal slug flow characteristics data was collected only for the Oil No 1 system and involved a small number of flow rates. Fig 8.41 shows the slug-front velocity data (average of 100 values) and suggests that this velocity lies between 1.2 and 1.5 times the superficial mixture velocity V_m. This is in agreement with a number of experimental and theoretical studies, an example being that of Gregory and Scott (1969) who gave the factor as 1.35.

Slug frequency is an important parameter as far as several hydraulic phenomena are concerned, the most important being pressure loss within the range of the current work. Measured slug frequencies are compared against the predictions of Gregory and Scott (1969) in Fig 8.42, where it is apparent that only at the highest V_m (corresponding to the highest V_{SL} since V_{SG} has been held constant) is there a good agreement between prediction and measurement. Note that the Gregory-Scott method is a function of V_{SL}, V_m and pipe diameter only, and therefore will predict the same value of slug frequency for various λ_w provided the above conditions do not vary. The current data clearly suggest that slug frequency is dependent upon λ_w to some extent, therefore ideally the prediction of slug frequency should incorporate water-cut effects; this is taken up later. The frequency of slug production is a transient effect and so also is the resulting size of the liquid slug unit which will be produced. The length of the slug depends on many parameters, including phase flow rates, pipeline inclination, pipeline length and local liquid levels; therefore both fluid dynamic and geometric characteristics are important. The description of slug length for a particular systems is usually in terms of an average slug length, a maximum slug length and, where possible, parameters which describe the shape of the slug length distribution. Obviously, the higher the number of slugs involved in the calculation the more statistically acceptable is the output result. For small-diameter pipes of typically 2-inch i.d. or less, a rule-of-thumb which emerged from early slug flow studies was that an average slug length of 10 - 30 pipe diameters was to be expected. For the current data, Fig. 8.43 gives a plot of the average slug lengths obtained, where it appears that the trends obtained are in agreement with the early experimental studies. Worthy of comment is also the minimum stable slug lengths observed. A physical model by Dukler et al (1985) can give predictions of the minimum stable slug length, which for the current systems gives values of approximately 8 - 10 pipe diameters. Visual observation and examination of video material suggested that any liquid units of shorter length were usually in the form of pseudo-slugs or roll- waves which did not maintain their form over even relatively short distances of 5 - 10m (100 - 200 pipe diameters), and in many cases the formation of a liquid bridge immediately resulted in subsequent collapse of the roll-wave/pseudo slug.

In conclusion, the slug flow characteristics in two-phase and three-phase flow did show up small differences between the frequency and lengths with or without water cut, but a great deal more data is required to support the observations and help explain the phenomena by means of physical modelling.

Comparison with historical data-based work suggested the average and minimum slug lengths to be similar to previous small-scale experimental data.

8.3.4. Comparison With The Present Inclined-Pipe Data

8.3.4.1. Flow Regime Data

As with the horizontal data, it was considered useful to test the prediction from the Taitel and Dukler (1976) model against the inclined-pipe data. This model was developed for near-horizontal pipes as well as true horizontal pipes, and an inclination of 1-degree from the horizontal is well within the suggested applicability range of the model.

Turning attention first to the uphill data, for the range of superficial phase velocities involved, intermittent flow was predicted in all cases, which is as was experimentally observed.

The inclined pipe data of Barnea et al (1985) taken from air/water flow in a 5.1cm i.d. pipe included flows at 2-degrees uphill incline, and all data points in the current V_{SG} - V_{SL} matrix resulted in slug flow. Recent data from Kokal and Stanislav (1989), using air and light oil in a 5.1cm i.d pipe, also produces slug flows for an uphill incline of 1-degree for the respective superficial velocity range. For the latter study, stratified flows at even very small uphill inclines are encountered in only a very small area of the flow pattern map, but this range was not within the flow rate matrix of the present tests. One area of possible distinction lies in the capability to report where the slugs genuinely arise from a terrain-induced drainback effect to give the expansion of the slug flow envelope from that encountered in the horizontal flow tests. As mentioned in Chapter 7, the excellent visual capability of the current system allowed the drainback effects to be observed which are largely responsible for the early production of slugs.

Whilst in most practical applications slug flow is the dominant regime in a slightly uphill-configured pipe, downhill flow in the current experimental system is possibly more complex than would be the case if the pipe was inclined downhill from the gas/liquids mixer. Unfortunately it was impossible to completely de-couple the effect of the uphill flow structure from that resulting in the downhill leg, but it was still felt valid to compare the Taitel-Dukler predictions with the data. Figs 8.44 to 8.48 compare the observed regimes at the end of the downhill leg with those from the above flow regime model. The transition of particular interest is that from stratified to intermittent-type flows, and it appears that in all cases slug/pseudo-slug flows were observed at lower liquid superficial velocities than is predicted by the model.

A similar trend was observed for the horizontal flow comparison; however, in this case the effect in some instances may be exacerbated by the fact that the feed stream is itself intermittent. The disintegration of slugs encountering the downhill section is dependant upon both the stability of the slug and the length of the declined pipe. Low V_{SL} produces slugs in the uphill pipe which will be expected to be less stable than those produced at increased V_{SL}, and thus will be more easily destroyed in the downhill leg. It is reasonable to assume that there should be a limiting stability to the uphill slug which will affect it's course in the downhill flow, and would be a function of both the length and severity of the decline. The downhill section is 550 diameters long, and this is considered sufficient to provide a good indication of the extent of the slug envelope in the downhill flow. Therefore the expansion of the stratified envelope observed was less than that predicted by the model, but had the uphill incline not preceded the downward section, slightly better agreement may have been obtained.

Barnea et al (1982) made an extensive investigation of downhill two-phase flow regime transitions over the full range 0 - 90 degrees inclination from the horizontal. The authors first note the dramatic expansion of the stratified flow envelope over the range of 0 - 10 degrees decline, with the stratified region being almost unchanged over 10 - 70 degrees inclination. Comparison with the Taitel-Dukler model suggested that it was adequate for downhill inclines of up to about 10 degrees.

Andreussi and Persen (1987) conducted an experimental and theoretical study involving flows in a 5cm i.d, 26m long downwardly inclined pipe at 0.65 and 2.1 degrees declination. Whilst most attention was focused on modelling the stratified flow, an important observation was that, at high liquid loading, the transition to slug flow occurred at lower V_{SL} than predicted by the Taitel-Dukler model, in agreement with the present experimental trend. The authors suggest that the Taitel-Dukler transition assumption of $h_L/d = 0.5$ should be modified to a value of approximately 0.25 to better represent their data, and admit that pipe length has some influence to the result. These authors obtained no annular flow in their experiments.

In summary, the present oil/gas and oil/water/gas data suggest that in a weakly inclined downhill flow the prediction of the stratified-intermittent transition by Taitel and Dukler leaves scope for improvement, although intermittent data in uphill flow was correctly predicted by the same physical model. In all cases, any intermittency in the flow was observed to extend to a significantly higher V_{SG} than is predicted by the model, which agrees with the horizontal-flow data trends.

8.3.4.2. <u>Liquid Holdup Data</u>

Due to the facility constraints, comparison with uphill holdup data is only possible in the present study.

Comparison with the predictions from the Eaton et al (1967) correlation is given in Figs 8.49 to 8.53. At $V_{SO} = 0.1$m/s it is clear that considerable under-prediction of holdup is obtained. As V_{SO} (and so V_{SL}) is increased for each λ_w, the discrepancies between measured and predicted holdups decrease gradually until at $V_{SO} = 1.0$ m/s acceptable accuracy is achieved, although the trend is a fairly consistent small under-prediction of the holdup. Similar trends were noted for the horizontal data, but in the inclined case the 2 lowest V_{SO} sets give an underprediction which is more severe in the case of the uphill data. This is a direct result of the higher holdups which are obtained in uphill flow, even at very low deviation from the horizontal, due to the action of gravity, buoyancy and hence increased gas/liquid slippage.

The Eaton et al correlation was developed for horizontal pipelines, thus it was not surprising that under-predictions in holdup are obtained in many cases. However, these data do indicate that at sufficiently high V_{SL} the correlation can give acceptable predictions for this particular geometry, oil/gas and oil/water/gas fluid systems.

The uphill holdup data was also used to test the holdup correlation of Mukherjee and Brill (1983) which includes consideration of pipe-inclination effects. As can be seen from Figs 8.54 to 8.58, at the lowest value of V_{SO} the prediction is an improvement over that of the Eaton et al correlation, but at the higher V_{SG} considerable under-prediction of the holdup remains. As for the previous correlation, as V_{SO} is increased the Mukherjee-Brill method becomes considerably more accurate and produces predictions which in almost all cases are perfectly adequate. One aspect of note with this correlation is that both under and overpredictions are obtained, and so is less consistent than the Eaton et al method which fairly consistently predicts lower-than-measured liquid holdups. However, again there was no trend observed in that any particular λ_w data sets resulted in particularly better or poorer predictions by the correlation.

Therefore, the holdup correlations, which have been modified to incorporate oil-water weighted fluid physical properties, on the whole give satisfactory prediction of total liquid (oil plus water) holdup at moderate-to-high liquid superficial velocity. At low V_{SL}, and especially when coincident with high V_{SG}, severe underpredictions of liquid holdup are obtained, but in several cases the extent of this under-prediction can be reduced somewhat by the use of the Mukherjee-Brill correlation which incorporates pipeline-inclination considerations.

8.3.4.3. Pressure Drop Data

Figs 8.59 to 8.63 display the measured and pressure losses against the Eaton-Dukler correlation predictions for the uphill data. The first point of note is that at V_{so} = 0.1 and 0.15m/s, in several cases under-predictions are obtained, whereas in the horizontal tests the correlation gave over-predictions in virtually every case. It is suggested here that the principal reason for this observation lies in the use of the Eaton correlation for holdup prediction; as discussed previously, at low V_{SL} in both a horizontal and uphill incline the correlation underpredicts the liquid holdup considerably. A lower liquid holdup, when input to the Dukler et al pressure loss correlation, will then produce a lower value of predicted pressure drop. As V_{SL} is increased, the uphill incline has less influence over the average holdup and the Eaton et al prediction becomes more accurate, resulting in consistent over-prediction of the pressure loss by Eaton-Dukler at higher liquid superficial velocities. However, it is evident that the mean error is lower than that generally observed for the horizontal data comparisons, the use of the Eaton et al correlation tending to lessen the over-predictions which occur when using the Dukler pressure loss correlation. Figs 8.64 to 8.68 depict the uphill data against the Eaton-Oliemans predictions. For the V_{so} = 0.1m/s and 0.15m/s cases similar trends are observed as for the Eaton-Dukler predictions in that the use of the Eaton et al correlation has resulted in a large number of pressure losses being under-predicted by the correlation. However, as V_{SO} is increased the trends are similar to those obtained in the horizontal tests. Again, the use of the Eaton-Oliemans methods results in less conservative pressure drop predictions when compared to the performance of the Eaton-Dukler method, and in many cases a high degree of accuracy was obtained.

The pressure loss data for the 3 highest V_{SO} sets for downhill flow are compared to the Eaton-Dukler predictions in Figs 8.69 to 8.71. In contrast to the trends for the uphill pressure drop, in all cases the correlation suggested higher pressure losses than were actually measured. This can again be explained in part, due to the use of the Eaton et al holdup correlation: in the downhill flow the holdup may be less than that obtained for the same conditions in the horizontal flow, though to what extent depends largely on the magnitude of inclination and the superficial liquid velocity. The assumption of a higher holdup will increase the predicted pressure drop, resulting in a trend very similar to that obtained for the horizontal flow, when the V_{SL} has reached a critical value.

A similar result was obtained for predictions for the Eaton-Oliemans method, Figs 8.72 to 8.74, in that the trends were very similar to those obtained with the horizontal data. This method again proved to be less conservative when compared with the Eaton-Dukler correlations, although it should be noted that at high V_{SL} the correlation does tend to under-predict the pressure loss.

In summary, the pressure drop data from the inclined-pipe study have shown that for the small inclination involved, in most cases adequate pressure loss predictions can be obtained when the fluid properties for oil/water/gas flow are modified using a two-phase correlation and a simple oil-water linear mixing rule. The data have also indicated that at moderate to high V_{SL} the inclination is sufficiently weak to enable the use of a horizontally-based liquid holdup correlation to be coupled to the pressure loss method.

8.3.4.4 Slug Characteristics

Fig 8.75 depicts the average slug front velocity $\overline{V_S}$, for both the inclined and horizontal data compared to the bounding lines of $1.2V_m$ and $1.5V_m$. The data shows that in all cases the measured $\overline{V_S}$ lie between the limits which represent the range resulting from a number of correlations. Note that these correlations, of which Gregory and Scott (1969) is one of the best known, were developed for horizontal flow conditions. This indicates that for the present system the pipe inclination is sufficiently small to enable these horizontally-derived methods to give reasonable predictions, although the data represents a narrow V_{SL} - V_{SG} range. This observation is supported by Mattar and Gregory (1974), who measured slug characteristics in a 25cm i.d system using air and light oil (viscosity 9cP) as the working fluids. These workers found little effect of pipeline inclination, over the range 0-10 degrees from the horizontal, on slug front velocity and develop a correlation which gives predictions for the current system which lie between the 2 limits of the lines on Fig 8.75.

Fig 8.76 shows the comparison between the measured slug frequencies and the prediction by Gregory and Scott (1969). Both the uphill and downhill data appear to be closer to the prediction than was the case for the horizontal data.

As the correlation was originally developed for horizontal flow, it seems that the more regular production of slugs in the uphill situation (and so, providing stability is satisfied, slightly more frequent slugs on the downhill section) gives better agreement, but does not then support the nature of the assumptions made in the correlation development, underlining the requirement for a more physically-based modelling approach as opposed to empirically-based activities.

8.3.5 Comparison with Existing Data

Chapter 3 suggested that very little quality data is available for oil/water/gas flows. This is certainly true, but it is still of value to present briefly this data in the light of the current experimental findings.

The early data of Sobocinski (1953) contains information which can be re-cast in a form consistent with the current work. Of chief interest is the pressure drop data, where only the highest V_{SL} data has been replotted in Fig 8.77 since this was the only data within the present superficial velocity range. Peaks in pressure loss are obtained at λ_w = 0.6-0.65, with only minor sensitivity to gas superficial velocity being evident. Sobocinski did not report the prevailing line pressure during his experiments: an average value of 1 barg has been assumed in compiling the above plots. The test pipe was of 75cm i.d but only of approximately 120 pipe diameters in length. This fact suggests that what was actually being measured could have been significantly affected by entrance-effects. Sobocinski did not obtain intermittent flow in his experiments, the flow regimes being predominantly stratified or distributed in nature. There then lies the possibility that the different flow regimes have been responsible for the different pressure drop - λ_w characteristic when compared to the intermittent-dominated present data, which is a point of note in any future studies.

The data of Malinowsky (1975) is mainly concerned with oil-water flow, but a number of oil/water/gas tests were also run. Here the data has been replotted for cases where sufficiently close V_{SG} and V_{SL} values can be obtained from the raw data where the λ_w is variable within these constant V_{SL} and V_{SG} values. The pressure loss data is plotted in Fig 8.78 and the first point of note is that a very narrow range of both liquid and gas superficial velocities are involved. In most cases, a peak in pressure loss is obtained at the 35-45% input water cut range, although the author gives no data whereby the base (dry) oil-gas pressure drops can be added for comparison. This peak in pressure loss was also obtained with the Malinowsky oil-water data, so in this system the three-phase pressure drop is seen to follow the oil-water trends.

Slug flow was the regime observed in all of Malinowsky's oil/ water/gas tests and this was successfully predicted by the Beggs and Brill (1973) flow regime map using no-slip oil-water average fluid properties. The author also used the Dukler et al (1964) and Beggs-Brill two-phase pressure loss correlations to back-calculate an effective liquid viscosity; this is only a sound approach if the correlations have been proven over the appropriate viscosity range for the particular system and this is not strictly true. It is worth noting that the small number of data points where the in-situ oil/water ratio was measured resulted in flowing oil/water ratios which were very close to the no-slip values, as has been obtained with the current horizontal data. The appearance of a pressure loss peak in the three-phase slug flow is indicative of a case where the oil and water are more intimately mixed than in the present experimental system; the fact that Malinowsky's test pipe was of 1.5-inch i.d. supports this suggestion, as a smaller pipe will result in increased shearing of the oil-water mixture for a given superficial velocity.

Using the same experimental system as Malinowsky, Laflin and Oglesby (1976) also collected three-phase data. Their pressure drop data has been replotted in Fig 8.79 where, as for the Malinowsky data, the data sets correspond to cases where V_{SL} and V_{SG} values are within 5% of each other for variable λ_w .

Intermittent flow was obtained in the experiments, which is as was predicted by both the Beggs and Brill (1973) and Mandhane et al (1974) flow regime maps, based on oil-water linear mixing for the liquid properties. Consistent with the Malinowsky data, peaks in the pressure drop are obtained at approximately λ_w = 0.4 in most cases. Again, the three-phase pressure loss was seen to display the same characteristic with water cut as was obtained with the respective oil-water only system, although the tests were not structured in a manner where the oil-water total superficial velocity could be used as a basis for the three-phase tests.

The most recent three-phase experimental data has been presented by Stapelberg et al (1991). Oil-water, water-gas, oil-gas and oil/water/gas data was collected in a horizontal 5.9cm i.d pipe of length 35m. Oil-water data was concerned with pressure loss, and the authors modified the Lockhart-Martinelli (1949) correlation to incorporate liquid-liquid flow. Despite the fact that the oil phase was of 31cP viscosity, pressure loss data revealed that even in oil-water homogeneous flow there was no increase in pressure drop at any particular range of water cut, which is of importance as a precursor to the three-phase data. The tests involving a gas phase resulted in intermittent-type flows in all cases, and the slug frequency, slug length and pressure loss were measured for each test. Only 2 values of total V_{SL} were involved, and the pressure losses have been replotted as functions of V_{SG} and λ_w in Figs 8.80 and 8.81: the data has been smoothed for illustration, and for clarity only the V_{SG} data of 1m/s and higher is shown, which also corresponds to the minimum V_{SG} of the present work. The trends display similar characteristics to those obtained in the present work, except that in all cases the oil-gas losses are higher than the water-gas and oil/water/gas results for a set V_{SL} and V_{SG} : this can be explained by the fact that for Stapelberg's work a clean oil-water viscosity ratio of 30 exists which is far higher than the maximum value of approximately 4 which exists in the current study.

It is interesting to note that the water-gas pressure drops often exceed the three-phase values; Stapelberg develops his discussion on this front by collecting data specifically involving slug characteristics (although not slug body holdup). The slug frequency data indicated that the oil-gas system produced slugs most frequently, and this can be attributed to increased liquid height in the stratified flow due to high oil viscosity. However, although the water-gas and oil/water/gas slug frequency data exhibited much scatter the trend was for the water-gas system to give slightly higher slug frequencies than the three-phase systems, except at the lowest values of V_{SG} where the results were very similar. At V_{SG} of 1.0m/s and greater, the slug lengths measured in the three-phase flow were observed to be greater than those for the water-gas and oil-gas tests; this latter observation is supported by the present data. Stapelberg et al use the Dukler-Hubbard (1975) model to predict pressure losses with the Tronconi (1990) slug frequency method as input, and fair success is achieved. However, it would appear that to fully describe the slug flow pressure loss the slug body holdup in the three-phase flow is an important parameter which deserves attention. Unfortunately, the range of V_{SL} and V_{SG} reported was not sufficiently broad to draw any firm conclusions in conjunction with the current data.

To summarise, the existing experimental three-phase data is somewhat fragmented in that studies have concentrated on different areas of the flow regime map. The three-phase pressure loss data presents the most puzzling information; it would appear that the characteristics obtained are closely related to the behaviour of the oil-water mixture alone, in terms of its tendancy to form dispersions or stratified layers. This point and many others form the remainder of this chapter.

8.4. THE REPRESENTATION OF GAS/LIQUID/LIQUID FLOW USING MODIFIED GAS/LIQUID METHODS

8.4.1. Introduction

The previous sections have shown that in many cases the three-phase experimental data was adequately predicted using modified two-phase approaches. It would be optimistic to assume that such methods will always produce similar success when applied to fluid systems of significantly different character, both in terms of fluid physical properties and the geometry of the system. It has indeed been suggested that although based on relatively sparse data, different characteristics can be obtained from different systems, although one could strictly argue that such findings apply also to gas/liquid flow. The purpose of this section is to discuss the merits or otherwise of existing two-phase approaches where much can be learned in the development of suitable three-phase methods. Attention will be focused on the flow regimes observed in the present study and the resulting liquids holdup and pressure drop characteristics which are produced. Although horizontal flow is the main configuration studied, discussions will also include the effects of small inclinations where appropriate.

8.4.2. Three-Phase Flow Regime

As indicated, such is the nature of gas/liquid/liquid flow that the number of flow regime classifications can be a function of the imagination of the investigator. In this section most attention will be paid to stratified and intermittent-type flows and their transitions to other flow patterns. The extent of liquid-liquid mixing in the flow is addressed, though it should be pointed out that it seems this parameter may be a very difficult one to obtain a good grasp of in practical terms.

8.4.2.1. Stratified Flows-Liquids Separated

Stratified flow in this context should be taken to mean gas-liquid stratified flow - the liquids themselves may distribute themselves in several ways as discussed previously.

Turning first to horizontal flow where the liquids are separated, the gas-liquid interface is smooth, we have an idealised three-layer flow, as shown in Fig 8.82. The basis of several transition models including Taitel and Dukler (1976) for gas-liquid flow, and Brauner and Maron (1989) for liquid-liquid flow, involves writing the 1-dimensional momentum equation for each layer. If treated as a gas-single liquid i.e. if the upper liquid phase in Fig 8.82 does not exist, these equations are:

$$\begin{aligned} -A_1(dp/dx) - \tau_{w1}S_1 - \tau_{i2}S_{i2} &= 0 \\ -A_3(dp/dx) - \tau_{w3}S_3 + \tau_{i2}S_{i2} &= 0 \end{aligned} \qquad (8.2)$$

The approach adopted for the 2-layer case is to eliminate the pressure loss which enables one to write a single equation in which the parameters can all be represented by functions of (h_L/d), the non-dimensional lower-fluid depth. The adjustable hydraulic diameter concept is used for each layer, with the wall shear stresses calculated using existing single phase friction factor methods. It is clear that the extension to 3 layers provides no simple extension of the above, and the solution is outside this study scope, but certain comments should be made regarding the use of 2-layer models where the liquid phase is actually a stratified oil-water combination.

The gas-liquid interfacial shear stress, τ_i, is often taken to be a simple function of the superficial gas Reynolds number, or, even more simply by Taitel-Dukler:

$$\tau_I = \frac{1}{2} f_I \rho_1 U_1^2 \qquad (8.3)$$
$$\text{with } f_I = f_1$$

This approach would seem reasonable if the interface is smooth, which is the basis of Taitel and Dukler's argument. Where there are waves present on the interface, the use of another interfacial friction factor correlation such as that by Cheremisinoff and Davis (1979) or Kokal and Stanislav (1989) may be more appropriate, where the former includes a term involving the Reynolds number for the liquid phase.

Turning back to the set of equations (8.2), one simplifying assumption which could prove useful is the fact that if the liquid velocities are close the interfacial friction factor of the lower interface can be neglected. Friction factors (single-phase) can be assigned to each layer (wall shear stress) as before, but an additional complication is the possible effect the oil phase has on the water transition to turbulent flow: it is expected that in most cases the water layer will be first to achieve turbulence, although it does of course depend on the oil properties and the relative depths of each layer. Stellmach and Lilleheht (1972) suggested that the transitional Reynolds number should be 1500 for the water, a point included in the oil-water model of Brauner-Maron (1989). It is also worth commenting that Wang and Charles (1981) found that the onset of turbulence was promoted by a decrease in oil-water interfacial tension: no separated flow models incorporate the effects of surface/interfacial tensions. Therefore the extension of equations (8.2) to the three-layer case could give useful results regarding three-phase stratified flow, but in many cases the smooth interfaces of above will not exist as will now be discussed.

The appearance of waves in the stratified flow is of importance in that it can bring a marked increase in interfacial friction factor (Andreussi and Persen (1985)) at the gas-liquid interface and also represents a mechanism whereby waves on the lower interface could be generated and hence commence oil-water mixing, if only to a small extent. The transition to waves at the gas-oil interface is expected to be a function of gas shear and the inertial/transport properties of the oil phase. The work of Andritsos and Hanratty (1987) involves a thorough analysis of the stability of perturbations over a liquid interface which has compared favourably to data involving a wide range of viscosities. The critical gas velocity for the onset of waves increases as the liquid viscosity is increased: the use of oil viscosity then appears suitable. If, however, there is present a liquid-liquid interface, the situation will be further complicated. Firstly, one would expect the viscous damping to be dependant upon the oil/water depth ratio to some extent. This in itself then requires in-situ knowledge of the oil and water fractions. Second, it is not improbable that the generation of even low-amplitude, long-wavelength waves at the oil-water interface may modify the theory appropriate to the gas-liquid interface, including again the thickness of the oil layer and possibly the phase relationship between the liquid-liquid waves and the growth of any small disturbance at the gas-liquid interface. Surface tension may also play a part; interestingly, Sherman (1968) suggests that the degree of viscous damping in such waves can be much enhanced if surface-active agents are present in the liquid. For basically separated oil-water mixtures, it seems reasonable that to a first approximation, the effect of waves at the oil-water interface should be considered a secondary effect when the roughened waves at the gas-liquid interface are considered. This then assumes that the contributions to shear stress and pressure loss at the gas-liquid interface only are influential; although again it is the relative influences of each wave phenomena which should be considered in a thorough analysis.

One of the most important results of interfacial wave action is the transition from wavy flow to an intermittent-type flow pattern, particularly slug flow. Predictions for the transition in gas/liquid flow have commonly been associated with two approaches thus:

Kelvin-Helmholz stability methods

and The use of linear wave stability theory

Taitel and Dukler (1976), amongst others, employ the first method and develop a simple criterion based on the stability of a single wave which is mainly dependant upon the liquid depth and the in-situ gas velocity as shown in Fig 8.83. The extension of this approach to a 2-layer liquid film has not been reported: at first glance one would assume that the amplitude of any waves at the liquid-liquid interface and the oil/water depth ratio γ would be of importance regarding the transition to intermittent flow. Liquid viscosity and interfacial tension effects are largely ignored, with only the former appearing in the calculation of the equilibrium liquid depth in the 2-dimensional, 2-layer separated flow equations. Therefore the success of using the above approach would appear to be highly dependant upon:

$$\textit{oil/water viscosity ratio } \beta = \frac{\mu_o}{\mu_w}$$

$$\textit{oil/water depth ratio } \gamma = \frac{h_{L1}}{h_{L2}}$$

Where the upper liquid has a viscosity greater than about 10cP Lin and Hanratty (1986) propose that the second slug transition analysis above is the more valid. Andritsos et al (1989) suggest that the sequence of wave generation and the type of waves produced in high-viscosity liquids differs from that found for low-viscosity liquids, although it is worth pointing out that these authors found similar behaviour for 20cP and 100cP liquids in terms of transition to slug flow. For the case of oil-water-gas flows with a separated oil layer, it is suggested that the above method is reasonable as a first guess where the oil viscosity is greater than 20cP, although again the parameter γ is important : one would expect that the above approach would only suffice if the oil layer thickness is sufficiently large. In favour of the former method used in the Taitel-Dukler model is the fact that the present oil/water/gas data was predicted similarly to the oil/gas data, although it is admitted that a finer grid of experimental points would help to investigate the transition more closely to enable a fairer comparison. The Taitel-Dukler approach is to assume the onset of intermittent flow where the equilibrium liquid depth h_L/d exceeds a dimensionless parameter in wavy flow. Again, this forces one back to developing a method to calculate the equilibrium liquid depth in a 3-layer flow. One should also remember that whether full blockage of the pipe can be sustained on the growth of a transition wave may in some cases be a complex function of the oil-water interfacial mechanisms.

Turning away briefly from horizontal flow, it must be said that one cannot ignore the effect which small pipeline inclination can have on a low-velocity, 3-layer flow in terms of bulk flow effects such as liquids holdup, based on the quantitative and qualitative observations of Chapter 7. Fig 8.84 presents a picture of the inclination contributions which can be obtained. In terms of the separated-flow modelling approach, gravitational terms are introduced to equations (8.2) to further complicate the situation. For low V_{SG} flows and particularly where the gas-liquid interface is smooth, one aspect which it would seem impossible to ignore is oil-water slippage. Whilst for many horizontal situations the assumption of oil-water no-slip may suffice, this is unlikely to carry to inclined situations, even where the inclination is very small. The situation then arises where the oil may slip over the water or the water slip from beneath the oil layer, and this is embodied in Brauner and Maron's (1989) oil-water model. Again, sources of primary interest are the equilibrium liquid level and the transition to intermittent flow.

An uphill flow naturally induces a higher liquid level due to the relative buoyancy of the gas phase and the effect of gravity on the liquid - it is unknown how much more influential to gas-liquid slippage a two-component liquid film will be as compared to a single liquid layer. However, what can be said from the present work is that the ratio γ can be heavily influenced by a 1-degree inclination at low V_{SG}, where in some cases a 20% input water fraction resulted in a water liquid height of over 50% of the total liquid height. The magnitude of the gravitational effect on the liquid as a whole may mask the particular oil-water slippage, but the effects on the in-situ liquid fraction should be remembered. This may be particularly important if one is considering the transition to the dominant uphill flow pattern, slug flow.

A minor downhill flow helps promote both liquid-liquid and gas-liquid slippage as would be expected. The current study involved no downhill holdup measurements, but visual observations suggested that in stratified flow (the dominant regime) oil-water slippage was occurring at low-to-moderate gas superficial velocity. It can be then considered that for the transition to slug flow, the use of the oil phase viscosity would prove to be the most satisfactory course to take if a gas-liquid method is to be employed.

It must be said that downhill flow, especially at low velocities, has created much interest for two-phase investigation. For such flows, it is highly likely that the effect of pipeline terrain will dominate the problem and that the solution will depend to a great extent on a thorough understanding of oil-water slippage phenomena.

8.4.2.2. Stratified Flows - Liquids Mixed

Having briefly discussed stratified flow where the liquids are separated, it is relevant to comment on the situation where the liquids are mixed. The most straightforward case is to assume the liquids form a homogeneous mixture, though where appropriate the effect of partial mixing or a concentration gradient will be mentioned.

At first glance a gas/liquid/liquid flow where the liquids form a pseudo-phase appears to offer a better chance of fundamental understanding than the 3-layer case, since here we now have, notionally, a 2-layer flow as depicted in Fig 8.85. Defined strictly, it does in fact represent a multi-interface problem in that every dispersed particle forms an interface with the continuous carrier liquid or the gas at the gas-liquid interface.

The case of a smooth stratified flow in the situation where the liquids are mixed would not be expected to occur very often in practice : low V_{SG} and generally low mixing intensity may be envisaged as having minor effects in the sense of creating a dispersion. Perhaps one of the few cases in practice would be the production of a stable dispersion where the gas phase has either been injected or has come from a very high GOR well. Even in a large-diameter pipe one would expect that a mixed oil-water phase would be co-existent with a gas flow providing significant mixing energy. Therefore the remainder of this section will assume that the stratified flow is of a wavy nature.

Before discussing the use of two-phase modelling and the associated flow pattern transitions, it is of value to consider briefly the source of this flow regime from the standpoint of oil and gas production. The primary interest for this study is full well stream transfer, where raw well fluids pass up the well bore through a subsea wellhead and into a multi-phase pipeline. The oil and water will in many cases mix chaotically as the liquids pass up the well tubing string; at the same time solution gas is coming out of the oil and promoting further fluids mixing. Therefore much liquid-liquid premixing may have been obtained before the flowing gas content is significant: the current facility also results in a situation where oil-water premixing takes place (at sufficient velocities) before the gas is added to the mixture. The extra mixing provided by the gas phase or resultant bulk gas-liquid flow regime may or may not influence the state of the oil-water mixture, although the current data does indicate some influence in terms of apparent degree of oil-water mixing through the test loop. These aspects should be borne in mind in relation to the following discussion.

The mechanistic modelling of this flow regime can theoretically be tackled using the same approaches as for two-phase flow if the properties of the oil-water mixture can be satisfactorily predicted. The current experiments indicated that, in terms of transition from wavy flow, there may be a small effect upon the transition to pseudo-slug flow if the liquid is an oil-water mixture, but this cannot be expanded upon due to the small amount of data particular to this facet. The current dispersions are envisaged as fairly loose in that the average droplet diameter is expected to be above 100 μm , and as such the linear mixing rule for oil-water properties has proved as accurate as that for oil/gas flow only. Also, the current data suggests oil-water slippage to be minimal; one would expect a mixed oil-water film to approach no-slip behaviour to a greater extent than if the oil-water combination consists of two layers. Therefore use of the no-slip oil-water assumption should prove useful, at least for horizontal flows.

Where the oil-water mixture is a water-in-oil emulsion where the water droplets are very small and form a size distribution which is very narrow, available information suggests that a simple linear-mixing type approach will prove unsuitable. This may be the case: but it should be remembered that much can depend on the base-oil viscosity and the proximity of the flowing water cut to that required for emulsion inversion. Arirachakaran et al (1989) have presented a simple relation for the inversion water cut $(WC)_{inv}$ for a number of experimental systems:

$$(WC)_{inv} = 0.5 - 0.1108 \log \mu_o \qquad (8.4)$$

The application of Eq (8.4) to three-phase flows is uncertain: the base data for the correlation involves viscous oils in laminar flow. However, it remains the only relation which has been compared to a number of experimental oil-water systems.

Where the liquid system behaves as an oil-in-water dispersion, typified by high water cuts, one may have a little more confidence in using an approach such as the linear mixing rule in conjunction with a gas-liquid prediction method. This is due to the fact that many investigators have observed the apparent viscosity of oil-in-water emulsions to be not greatly different to that of water alone, although it is by no means certain that this is due only to the chemistry of the fluids.

Conceptually the transition from wavy to intermittent flow for a homogeneous liquid-liquid mixture appears more tractable than for 2 liquid layers. This may well be so, but it would seem to be dependant upon the relative influences of the following:

i) Liquid viscosity effect on equilibrium liquid depth

ii) The behaviour of waves propagating over a dispersion film

iii) Foam propensity at the gas-liquid interface.

The first matter above is largely concerned with the effect of liquid viscosity on the gas-liquid slippage, and as such is encountered later, but a higher liquid phase viscosity will in most instances promote a higher liquid depth and hence be more susceptible to the formation of slugs/pseudo-slugs due to waves at the interface. The second factor above is perhaps even more tied to the system chemistry : the viscous damping in the dispersion may be quite different to that for the oil-only layer. No instances were found in the literature involving examination of this flow pattern transition where the liquid was an oil-water dispersion. The third factor is included to provide reminder that oil-water emulsions are not the only dispersions of much importance to raw-wellstream production; even less work has been published concerning foams produced in crude oil transportation. The effect of water on the foam volume produced, which is tied to the fluids physical chemistry, may produce significantly different characteristics to that for the oil-phase alone. This may have an effect on the transition to intermittent flow, and it may affect other phenomena detailed later.

The effect of inclination on stratified flows - liquids homogeneous, as for separated oil-water layers, may be expected to be more dominant at lower fluid velocities. For uphill flow, the dominance of slug flow will perhaps be heightened to an extent less than that for separated layers due to the oil-water slippage. For a homogeneous film, the droplet size will depend on the fluid properties and the complex mechanisms at the gas-liquid interface, where gas superficial velocity will be expected to be influential. The present uphill in-situ water fraction data indicated an effect of V_{SG} at low λ_w which may help to give pointers as to the behaviour, but much of this data was separated or semi-homogeneous (oil-water). A reversal of this slippage behaviour in downflow will be expected. However, for inclined flows it is not inconceivable that the bulk gas-liquid effects far outweigh any extra contribution due to oil-water slippage in a homogeneous flow.

8.4.2.3. Oil/Water/Gas Intermittent Flow - Liquids Separated

The previous sections have briefly discussed the way in which transition to slug flow may be affected by either a two-liquid layer film or a film of an oil-water dispersion. This section will concentrate on the former, and in particular on the characteristics of slugs and their breakdown towards the annular regime.

Many instances in the present experiments provided the situation where slugs of liquid were produced where the slug was a well-separated 2-liquid film in regions removed from the slug front and tail : this is illustrated in Fig 8.86 . The transition to slug flow has achieved a high level of research attention, but this section will be restricted to models concerned with the nature of the slug flow once firmly established.

A gas/liquid slug model which has gained some acceptance is that of Dukler and Hubbard (1975), and Fig 8.87 depicts their vision of the slug flow mechanics. Basically, the model considers the slug flow zone to include the pickup of the film in the mixing zone, full-pipe flow in the slug body, and the shedding of liquid at the rear to leave a film region.

The steady-state behaviour is modelled by considering the dynamics of the film, slug front, slug body, the liquid pickup and shedding and the overall mass balance. In common with other models, however, it requires *a priori* knowledge of 2 of the following parameters:

i) slug frequency

ii) slug holdup

iii) slug length

to close the problem properly. Bulk liquid behaviour within the modelling can be accommodated using a fluid-property correction approach, but it is perhaps the features above which if properly accounted for in an oil/water/gas flow, could hold most promise for satisfactory slug flow predictions and the associated effects of average holdups and slug flow pressure loss.

The slug frequency correlation of Gregory and Scott (1969), showed very variable success when compared to the present data, although the range was admittedly narrow. The slug frequency is given by:

$$\nu_s = 0.0226\left(\left(\frac{V_{SL}}{gd_p}\right)\left(\frac{19.75}{V_m} + V_m\right) \right)^{1.2} \qquad (8.5)$$

Therefore no dependence on fluid properties is demonstrated by the correlation. The correlation of Greskovich and Shrier (1972) is similar in form to Eq (8.5) where no effect of liquid viscosity can be accommodated, but has compared favourably with one set of large-diameter pipeline data.

A departure from the correlation-based approach was again made by Taitel and Dukler (1977) who developed a theoretical model of slug frequency. The method includes the development of complex sets of equations, where the equilibrium liquid level is computed by the earlier Taitel and Dukler (1976) model. Therefore the model incorporates some account of fluid-property and pipe-diameter effects. One is then forced again to make the assessment of whether a simple approach such as the linear-mixing law is appropriate. A separated oil-water situation offers the possibility of using only the oil layer physical properties if desired; for a low-viscosity oil this may prove a reasonable assumption, but for very viscous oils the effect which may be exerted on the calculation of liquid level may prove overly conservative. Also, the above model indicated better prediction capability at low-to-moderate V_{SL}, which should be remembered when applying to three-phase flows equally.

Recently Tronconi (1990) extended previous theory on wave growth to develop a method for slug frequency determination. The theory assumes that the slug frequency ν_s is inversely proportional to the period of the waves in the inlet pipe which are precursors of the slugs:

$$\nu_s = \frac{C_1}{\omega} \qquad (8.6)$$

where C_1 is a constant. The author assumes that half of the inlet waves will form slugs which will be stable in the pipeline and so $C_1 = 0.5$. With this, the calculation reduces to:

$$\nu_s = 0.61\left(\frac{\rho_g}{\rho_L}\right)\left(\frac{V_g}{1 - h_L/d_p}\right) \qquad (8.7)$$

Where V_g is the actual gas velocity. The Taitel-Dukler (1976) approach for equilibrium liquid depth is used, except that for turbulent gas Reynolds numbers interfacial friction factor f_i is twice that of the gas i.e. $f_i = 2f_g$, which is a modification to the original Taitel-Dukler assumption. The method has found good agreement with a range of data sets, including viscosities up to over 30cP, and is expected to be a marked improvement over earlier empirical approaches, despite being relatively simple to use. The extension of this method to a three-phase flow involves chiefly the prediction of equilibrium level (i.e. for a three-layer stratified flow) from previously, and the value of the constant used for the critical wave fraction in Eq (8.6).

The determination of C_1 would ideally take into account any effect of liquid-liquid interfacial waves on the gas-liquid wave characteristics and hence slug frequency. If it is assumed ω is calculated for oil-only properties, a simple approach may be to take:

$$C_1 = f(\gamma, \beta) \qquad (8.8)$$

That is to say, the relative depths of the oil and water layers, and the influence of oil viscosity, should play important roles. The three-phase slug flow study of Stapelberg et al (1991) gave C_1 = 0.25 to 0.33 when oil fraction was 50% or more, whereas at λ_w = 0.75 and air-water only a C_1 of 0.5 described their data most accurately. This data is of much value in that it shows the conflicting effects of increased oil viscosity on equilibrium liquid level and wave dissipation by viscous damping (the oil viscosity was 31cP).

The development of methods to predict the slug body holdup has followed a similar path to that taken for slug frequency. The first major study was that by Gregory et al (1978), which resulted in the following empirical correlation:

$$H_{LS} = \left[\frac{1}{1 + (\frac{V_m}{8.66})^{1.39}} \right] \qquad (8.9)$$

The system used to generate the data on which Eq (8.9) is based were air and light-oil in a 2-inch i.d. pipe, with poor comparisons when tested against an independent data set which was regarded as being subject to large scatter. Recognising the limitations to this empirical approach, Barnea and Brauner (1985) have produced an approach which relies on the slug-dispersed bubble transition prediction of Taitel and Dukler (1976). The authors suggest that the void fraction in the slug is at a maximum at the slug dispersed bubble boundary: with this knowledge of V_{sg} and V_{SL} at the flow rates involved the holdup can be calculated. This method compared satisfactorily with the above data of Gregory et al (1978) and gives more confidence in terms of application to greatly different systems. However, for both two and three-phase flows alike, the method does rely on accurate determination of the transition, and there is no escaping the fact that for rapid determination for a three-phase flow application of Eq (8.9) is very attractive on the grounds of simplicity.

Recently, Andreussi and Bendiksen (1989) developed a model of the slug holdup in horizontal and inclined flow, which essentially examines the mechanisms of the entrainment of small bubbles at the slug front and their subsequent production at the slug tail. Whilst being theoretical in nature, practical simplifications are presented in several cases where data is available, although these simplifications relate to the bubble behaviour and independent data is used for comparison against the final relation. The method displays good agreement over a wide range of fluid properties and pipeline diameter. The authors suggest that based on the data trends of air with light oil or water, there may be an effect of surface tension. The influence involves the fraction of the gas bubbles which are formed by vortex motion at the slug front which are lost back to the gas bubble in front of the slug. When the initially stratified film consists of an oil-water double layer, again to use a method such as the above may require a departure from using a simple linear-mixing approximation. Similar to slug frequency prediction, it is suggested that:

$$H_{LS} = f\left\{\gamma, \frac{\sigma_o}{\sigma_w}\right\} \qquad (8.10)$$

Where the λ involved is low-to-moderate, a good approximation may be to use the oil-gas surface tension. On the other hand, at high λ_w the use of the linear mixing rule or the water-gas surface tension may be more appropriate. With no experimental 3-phase slug body holdup data available, this suggestion cannot be examined further. However, it should be mentioned that if conservative estimates are required (i.e. usually high slug holdups), the use of the water-gas surface tension is the more appropriate. One should also remember that in the present experiments observations were made where the vorticity at the slug front could produce a high level of oil-water mixing, even if the oil-water character in the slug body was of a well separated nature.

A similar mechanistically-based approach has been employed less frequently on the aspect of slug length determination, although advances have been made following the identification of the slug growth phenomenon by Scott et al (1987). Observations in small-diameter pipes has led several investigators to assume an average slug length at approximately 30 pipe diameters (e.g. Nicholson et al (1978), Kokal and Stanislav (1989)), but slug lengths are variable in nature, and little published theory regarding slug length exists. Dukler et al (1985) presented a theoretical approach to calculate the minimum stable slug length, but this is not the parameter usually of key interest. The development of a two-phase slug length method has been concentrated on statistical approaches and these can be used for 3-phase data: see Brill et al (1981), Scott et al (1987). However, one should point out that the current data suggested an effect on slug length of water cut where water-cut slugs were longer on the average for the flow rates examined. The more extensive data of Stapelberg et al (1991) supports this: in their study 3-phase slug lengths were in many cases considerably longer than their gas-liquid equivalents.

The increase of V_{SG} then results in a higher average void fraction and higher voidage in the liquid slug units, and there then commences the end of the slug flow region and ultimately to the annular or annular-mist flow pattern. Although several mechanisms have been proposed for the intermittent-annular transition, one feature of common agreement is that this transition is a gradual one and open to much subjectivity. The existence of 'pseudo-regimes' between the annular and intermittent regions has been described as wavy-annular and proto-slug by a large number of investigators. Taitel and Dukler (1976) suggest that the transition depends critically on equilibrium stratified liquid level, reinforcing that for a 3-layer flow the calculation of this parameter would prove valuable.

The pseudo-slug region in the current tests (two-phase or three phase) was dominant at the higher V_{SG} tests. Therefore annular flow was strictly defined as existing only if no pseudo-slugs were observed in the test run. The data of Lin and Hanratty (1987) also showed the pseudo-slug envelope to be extensive for air/water flow, where a non-visual determination was made.

Weisman et al (1978) found that from a wide range of experiments liquid viscosity had no effect on the transition to annular flow, and density and surface tension exhibited weak influences. This is reflected in their correlation thus:

$$1.9\,(V_{SG}/V_{SL})^{0.125} = \left[\frac{V_{SG}\sqrt{\rho_G}}{g\,\sigma(\rho_L-\rho_G)^{0.25}}\right]^{0.2}\left[\frac{V_{SG}^{\,2}}{g\,d_p}\right]^{0.18} \tag{8.11}$$

Therefore if one wants to apply Eq (8.11) to a 3-phase system, several choices exist as to which surface tension value to use. When one examines the likely range of tensions in the system of interest, whichever value is used will not greatly influence the predicted annular transition point.

Recently, Kokal and Stanislav (1989) developed a simple correlation embodying a number of physical considerations. They suggest that where the average holdup falls below 0.25, the transition to wavy-annular flow takes place and further considerations leads to a relation for the transition to annular flow:

$$\begin{aligned} V_{SG} &= 10.36\,V_{SL} + C \\ \textit{where } C &= 2.98\left[\frac{g\,d_p(\rho_L-\rho_g)}{\rho_L}\right]^{0.5} \end{aligned} \tag{8.12}$$

So this method only considers the effect of liquid-phase density in terms of liquid properties, and calculations using a linear-mixing approach should prove satisfactory. Comparisons with the current data however, has shown success similar to that for the Taitel-Dukler physical model, although the subjectivity of the regime should be borne in mind.

Pipeline inclination will exert a variable influence on the aspects discussed in this section, and at low superficial mixture velocities, several of the slug characteristics can be greatly modified by even very small inclinations.

With separated liquid layers a small incline will encourage oil-water slippage as discussed previously. This may not overshadow the bulk gas-liquid slippage which causes earlier transition from wavy to slug flow in uphill flow, but it may affect the slug frequency. Unfortunately, very little data concerning slug frequency in inclined flow is available in the literature. The present data, at only 2 values of V_m, suggested uphill γ_s for 2-phase and 3-phase flow to be similar in magnitude with the 3-phase values slightly higher. The Taitel and Dukler (1977) model can be used to predict inclined slug frequencies in three-phase flow with the usual necessary modifications. However, much depends on V_m and it would seem that at higher V_m the effect of inclination to both the oil-water and gas-liquid slippage reduces considerably.

The effect of inclines of less than about 10-degrees on the slug void fraction has been shown to be very small by Mattar and Gregory (1974) who measured this parameter in an air/oil experimental system. More recently, Andreussi and Bendiksen (1989) have presented experimentally and theoretically that the effect is very weak for V_m of 5 m/s or greater, with 4-degrees the maximum inclination considered. The authors comment that a higher slug void fraction is expected in uphill flow and the reverse in downhill flow, than in the horizontal case. Again, the use of this method for 3-phase slug flow with separated liquid layers should consider the fluid properties to employ, particularly surface tension. The effect of inclination on slug void fraction is most important at low velocities: the same argument applies to oil-water slippage. Therefore in a downhill flow, the use of oil surface tension may be best, and in uphill flow using the water surface tension may be the most realistic, with a simple weighting rule available for both if desired. For moderate to high V_m, the assumption of oil-water no-slip may still prove adequate in inclined flows.

The present work suggests that both two-phase and three-phase uphill flow slug lengths are significantly shorter, on the average, than in horizontal flow at the same V_m. However, the differences between the two-phase and three-phase L_s show a less consistent trend than the horizontal data. The theoretical modelling of slug flow and slug characteristics measurement is a subject of intense research and field interest at the present time. The data from this study and that from Stapelberg et al (1991) strongly suggests that slug characteristics can be noticeably altered when one moves from a gas/liquid to a gas/liquid/liquid flow structure.

8.4.2.4 Oil/Water/Gas Intermittent Flow - Liquids Mixed

At sufficient flow rates of gas and liquid in the slug flow, a mixed liquid phase may be expected for many fluid systems and was indeed the main flow regime noted within the current experimental matrix. Although the true extent of oil-water homogeneity could not be determined precisely, this discussion assumes that the liquid phase is well mixed with little or no concentration gradient across the pipe vertical cross-section.

The use of existing models such as Dukler and Hubbard (1975) has already been discussed, and the extension to a situation where the liquids are mixed in a loose dispersion may be adequately described by a simple mixing rule between the oil and water for the bulk liquid properties. It is important to stress that in this situation, the roles which macroscopic and microscopic factors play may influence the outcome to very different degrees.

Consider first the prediction of slug frequency. Present theory suggests the slug frequency to be tied to the wave mechanics in the pipe inlet prior to slug generation. No data was found concerning wave flow in pipes or channels where the liquid is a liquid-liquid dispersion. One can imagine the wave properties to be dependent upon one factor in particular; the dispersion apparent viscosity. The present data for pressure loss suggests that no significant increase in apparent liquid viscosity was obtained. It has been suggested (e.g. Andritsos and Hanratty (1987)) that competing effects exist in the viscosity effect on the liquid level and the characteristics of the interfacial waves. Use of the Tronconi (1990) method for example, suggests modification to the C_1 factor is appropriate to three phase flows. Where the apparent viscosity is not greatly different to that of the dry oil, a significant change in slug frequency may not be obtained. If the oil/water fraction and mixing intensities are able to produce a water-in-oil emulsion however, the conservative approach would be to expect a significant increase in the frequency of slug generation: this presupposes that the increase in liquid depth associated with the higher apparent viscosity overpowers the wave energy dissipation due to increased viscous damping. Certainly, the gas/liquid horizontal slug frequency data of Kago et al (1987) and Stapelberg et al (1991) support this approach. This assumption also supposes that wave behaviour on the surface of a liquid-liquid dispersion is similar to that for a pure liquid, although Sherman (1968) commented that this may not always be the case. It is valid to mention that non-Newtonian, pseudoplastic behaviour of the liquid phase may also have little effect on the slug frequency, based on the experimental findings of Rosehart et al (1972,1975).

The applicability of two-phase slug holdup methods to a homogeneous three-phase flow is complicated not only by the possible departure of liquid properties from a linear-mixing law, but also by the lack of knowledge of small-bubble penetration through a liquid-liquid dispersion. Methods exist (Hinze (1955), Levich (1962)) to predict the maximum stable droplet/bubble diameter in a turbulent shear flow, but for the present application the methods have two major failings:

i) They are for bubbles/droplets dispersed in a pure liquid.

ii) They are appropriate only when the dispersed phase concentration is very low (typically less than 2% (vol)).

The prediction of bubble/droplet size distribution suffers from similar drawbacks. The use of modified two-phase methods to determine three-phase slug holdup is at present the only practical approach. Of course, taking this course neglects many phenomena, some of which have little practical impact in any case. However, one aspect which may prove important is the effect of dispersed bubbles in the slug of liquid-liquid dispersion: these bubbles may alter the droplet-droplet interactions and, at the limit, modify inversion behaviour and hence the achieved apparent liquid viscosities. As regards the rheology of the dispersion, the data of Rosehart et al (1975) indicate that even for extensive pseudoplastic behaviour (flow index $n' = 0.4$, where $n' = 1$ is Newtonian behaviour), the slug void fraction was only very slightly different to that obtained for Newtonian, air/water flow.

8.5 THREE-PHASE LIQUID HOLDUP PREDICTION

8.5.1 Using Generalised Two-Phase Holdup Methods

8.5.1.1 Horizontal Flow

The term generalised in this context is to highlight that the holdup methods have been developed with a wide range of flow conditions as input and are not aimed at particular flow regimes or flow rate ranges. Most of the earlier correlations are of this class.

Many correlations have been derived using laboratory data; for the purpose of this discussion two correlations are considered which are still relatively widely used in the industry.

The correlation of Eaton et al (1967) was developed for use in horizontal configurations. A large amount of data was collected from principally air/water experiments in 2-inch and 4-inch i.d test lines, but some field data was also included. The liquid holdup was correlated with:

$$\left[\frac{N_{LV}^{\,0.575}}{N_{gV}N_{d}^{\,0.0277}}\right]\left[\frac{p}{p_b}\right]^{0.05}\left[\frac{N_L}{N_{Lb}}\right]^{0.1} \tag{8.13}$$

The use of N_L, N_{LV} and N_{GV} ensures that the 3 liquid fluid properties are accommodated in the correlation to different degrees. N_{Lb} contains the base liquid viscosity which conveniently for the present study is that of water. As in the current study, a simple modification for the liquid fluid properties can be employed. The authors urge that application be limited to 20cP liquid viscosity or less: the Eq (8.13) displays a rather weak sensitivity to liquid viscosity and it may be that at much higher viscosities the function is in actual fact stronger. Nonetheless, if one was to use the Woelflin (1947) emulsion viscosity correlation for a 10cP crude at relatively low λ_w , the recommended limits will be exceeded. The present data indicated a linear-mixing role to suffice for most flow rates: the poor performance at low V_{SL} is expected to be unconnected with the use of this approach since oil/gas data was similarly predicted. The existence of oil-water slippage cannot be accounted for in such correlations: if behaviour was different to the oil-water no-slip here, this would favour under-prediction of the total holdup.

It is worth stressing that the present tests involved largely intermittent flows, therefore the performance of the holdup correlation in a number of bulk gas/liquid regimes could not be adequately assessed. Mukherjee and Brill (1983) produced a holdup correlation intended to be generally applicable to two-phase flows. A very large bank of data was used and the resulting correlation is not dissimilar to that discussed above. Since much of the data used to generate the Mukherjee-Brill correlation was experiments using 29cP viscosity lube oil, one would expect it to perform better at high μ_L than the previous correlation. This suggests that if the formation of a viscous water-in-oil dispersion is to be expected, use of this correlation with suitably-modified liquid properties may be the best approach if a generalised method is desired.

8.5.1.2 Inclined Flow

The large changes which can result when even small uphill or downhill inclinations are involved has meant that few investigators have tackled the prediction of inclined-flow holdup where the flow pattern need not be known *a priori*. The aforementioned correlation of Mukherjee-Brill can be used for two-phase predictions. The correlation form is an extension of the horizontal relation to which inclination angle coefficients have been added. The method is not general in that for downhill flow 2 sets of coefficients are provided; one for stratified and one for non-stratified flow. The present data indicates that adequate uphill three-phase holdup prediction was obtained using the above correlation, where an oil-water linear mixing rule was used. This is interesting in that it includes the points where oil-water slippage was obtained, although in all cases slug flow was involved. At moderate-to-high V_m, the oil-water behaviour reverts to no-slip and indeed the effect of the inclination involved decreases as the mixture velocity increases. It is reasonable to expect that in systems involving higher oil viscosities and more pronounced inclination, the influence of oil-water slippage would increase for low V_m flows. This then requires the development of methods which can build the oil-water slippage effects into an existing two-phase correlation or ideally the derivation of a new physically-based three-phase holdup method.

8.5.2 Using Flow Condition - Specific Two-Phase Liquid Holdup Methods

8.5.2.1 Horizontal Flow

The recent trend has been in the development of correlations and models specific to particular flow regimes and where low liquid fractions are involved. Several of these methods will now be mentioned regarding their applicability to three-phase flows.

The Beggs and Brill (1973) correlation is still commonly employed to obtain two-phase pipeline liquid holdup predictions. The horizontal holdup was correlated using functions of the no-slip liquid holdup λ_L and the mixture Froude number N_{FR}; therefore no account of liquid viscosity or surface tension is included. Different correlation constants are used depending upon which flow pattern exists (ie., a flow regime model is then required), which the authors classify as segregated (ie., stratified or annular), intermittent or dispersed. The correlation is based on small-diameter, air/water data, suggesting that even if the effect of different flow patterns can be accounted for, the correlation favours predictions where low liquid viscosities are present. The correlation cannot accommodate a change to three-phase flow via alteration of the liquid physical properties: for low viscosity (or loose dispersion) systems this may not necessarily lead to unreasonable predictions. Alternatively, where high effective liquid viscosities are to be expected, underprediction of the total liquids holdup may well result. Despite these possible three-phase drawbacks, the method continues to be applied to field systems with variable success - see Gregory and Fogarasi (1985) for example.

One area where much effort has been made to achieve acceptable holdup estimates is that of high gas/liquid ratio (GLR) pipelines which are encountered in wet-gas and gas-condensate systems. In such systems a very low input liquid fraction may be involved, and the resulting flow pattern is usually one of wavy, annular or mist flow. It has been recognised that early global-type correlations could fail badly when tested against such data, and new specific methods were sought.

Minami and Brill (1987) collected data and developed a correlation specifically for liquid holdups of 35% (vol) or less. Various test fluids were involved and some tests were run to examine any effects of surface tension on the holdup. The data indicated surface tension had no observable effect.The authors comment that their original correlation included a liquid viscosity term but this was removed since it was contributing to data scatter. Therefore the correlation is μ_L independent and the authors claim applicability since they envisage utilisation for gas-condensate systems only.

This point then demands some guidance again from the fluids physical chemistry, when one is considering the water-cut case. Whilst the behaviour of crude oil-water mixtures has yielded some limited laboratory data in the literature, no such studies could be found where the hydrocarbon is a condensate or natural gas liquid (NGL) - type fluid as opposed to a black oil. The base oil viscosity of condensate liquids in many cases will be less than that of a crude oil, but it is still unwise to dismiss the effective viscosity from this viewpoint alone. However, if one is forced to consider the practical occurrences of this type of flow in large pipelines, it is fair to say that in the majority of cases the water content will be low, largely arising from saturation from the gas stream. Therefore for a condensate/water/gas system the use of a linear-mixing role may be effective for many systems. The 35% liquid fraction limit of Minami-Brill does not always preclude the existence of a crude-water mixture and for high GOR production the usual choice of whether to consider the oil-water properties as averaged or to take dispersion effects into account, is forced upon the designer. If the latter effects are expected to occur, the Minami-Brill method may result in under-prediction of holdup.

Several more recent holdup methods have emerged to predict holdup in low liquid fraction pipelines, typically much lower than the limit set by Minami and Brill to their correlation. Methods of note are those due to Baker and Gravestock (1987) and Hart et al (1989). The use of these methods in three-phase flow with a simple oil-water mixing role appears a logical first step. Mention should be made here of another fluid-chemistry effect which may prove influential at low liquid fraction - foaming. The above study of Minami and Brill (1987) indicated that at high V_m increased liquid foaminess led to drastic reductions in the measured holdups. Where the hydrocarbon liquid contains a water fraction, no methods were found which can predict the effect on fluid foaminess. This is a very complicated fluids chemistry problem and one would expect that the effect of water would depend on the compounds particular to the produced water involved, particularly the type and concentration of the salt elements. However, if one expects in oil/gas flow this holdup reduction mentioned above to occur, the conservative approach would be to neglect this effect if water is present in the flow.

8.5.2.2 Inclined Flow

Most of the methods mentioned in the previous section can give inclined-pipe holdup predictions also. The Beggs and Brill (1973) correlation was developed from a databank containing holdup data from the full range of pipe inclination. The authors develop an empirical inclination correlation factor which is dependent upon the flow regime and the scale and sense of the inclination. A very weak liquid viscosity dependance is included, which may result in poor predictions if the oil-water combination is a tight dispersion.

Where low-velocity, stratified (liquid-liquid) conditions prevail, oil-water slippage may be significant and could impact upon the bulk liquid holdup: depending upon the sense of the inclination, this could result in either over-or-underprediction of holdup.

It can be argued that methods which have been developed for low liquid-fraction pipelines (e.g. Baker and Gravestock (1987) above) are less susceptible to the effect of weak pipeline inclination. This is based upon the fact that V_{SG} is likely to be relatively large with an annular-mist type flow pattern. Oil-water slippage may be expected to produce a smaller influence than on low-velocity flows, therefore the manner in which such methods can be used for three-phase flow reduces to the assumptions discussed in relation to their horizontal flow application.

8.6 OIL/WATER/GAS PRESSURE DROP PREDICTION

8.6.1 Three-Phase Horizontal Pressure Drop Prediction Using Modified Two-Phase Methods

A great many pressure drop correlations have been developed for two-phase flow, but few have been properly tested against quality three-phase experimental data. A number of these methods will now be reviewed with respect to their application to oil/water/gas flow, and for clarity generalised and specific correlations are detailed separately.

8.6.1.1 Flow-Pattern Independent Methods

One of the earliest methods developed for two-phase flow was that of Lockhart and Martinelli (1949), which produced a correlating curve of two-phase pressure drop with parameter X^2 thus:

$$X^2 = \frac{\left(\dfrac{dp}{dx}\right)_{SL}}{\left(\dfrac{dp}{dx}\right)_{SG}} \qquad (8.14)$$

For three-phase calculations a value of the oil-water only pressure loss is then required. There is no problem in obtaining the 'pseudo-liquid' single-phase pressure loss if one can adopt a simple oil-water mixing rule: modifications to the liquid property terms produces values of Reynolds number and so friction factor. This approach may not be appropriate in many cases. If it is to be properly employed, this correlation ideally requires knowledge of the liquid-liquid superficial pressure drop *a priori*. This point highlights the importance of obtaining accurate oil-water pressure drop prediction: it serves as a useful input to many oil/water/gas considerations. It is possible to use the liquid-liquid extension of the above approach, suggested by Charles and Lilleheht (1966), to produce oil-water only pressure drop. One drawback with such approaches is, however, that the state of the oil-water mixture during three-phase flow may be completely different to that existing in the superficial flow i.e. the mixing intensity provided to the liquid-liquid combination could radically affect the prediction outcome.

The methods developed following the pioneering Lockhart and Martinelli work involved an approach where workers focussed on obtaining an equivalent two-phase friction factor which could be used in a calculation similar to that of the single-phase, Moody (1944) graphical method. All pressure loss correlations, with or without flow regime dependance, require a value for the in-situ gas and liquid fractions. The approach many workers have adopted is to couple particular holdup methods to a pressure drop correlation to compare against a data set which is representative of the system of interest, if such information exists, and then selecting the combination which on the basis of experience is most appropriate. This obviously is not a pleasing approach, and in three-phase flow the expansion of this approach would be very cumbersome if in addition a large number of, for example, oil-water mixing assumptions are tested.

A historical correlation which is often considered for pressure drop is that of Dukler et al (1964). The method, whilst using similarity analysis in the approach, is basically empirical and cannot take flow pattern into consideration. Generally conservative results are obtained, as was also observed in the current study where a linear oil-water mixing rule was used for three phase flow. However, much depends on the characteristics of the oil-water mixture: a loose dispersion or separated layers will be better represented by simple mixing rules than will be a tight water-in-oil emulsion. For the Dukler et al correlation, no data could be found to give indication of performance as the μ_L increases appreciably above around 20 cP. Fayed and Otten (1983) compared a small amount of oil/water/gas field data with the Dukler pressure drop method: overpredictions were attributed to the formation of water films and consequent drag reduction.

It is probable that the degree of system influence in a three-phase flow may be somewhat greater than that for a gas/liquid flow. Methods such as that of Dukler et al compound this further by not being able to distinguish between different flow patterns. Whilst in some situations the bulk two-phase flow effects (slugging, for example) may mask any effect of oil-water interactions, some systems may well display sufficient mixing intensities to create high apparent liquid viscosities. This will have a knock-on effect through flow regime (possibly) and liquid holdup (probably). The extent to which μ_L alone effects two-phase pressure drop is largely unknown. Eaton et al (1967) claimed that the effect of μ_L on two-phase pressure loss is essentially negligible for liquid viscosities less than 12-15 cP; many tight emulsion fluids will significantly exceed these viscosities.

8.6.1.2 Using Two-Phase Specific Pressure Loss Prediction Methods for Gas/Liquid/Liquid Flow

Where a worker has set out to examine pressure loss in a specific two-phase flow regime one naturally assumes that better predictions will result than if one uses a more general method. This section comments on a number of methods available for such situations in a horizontal flow.

The method of Beggs and Brill (1973) involves the use of different empirical correlations for different flow patterns. Relatively few parameters are required to obtain a pressure drop prediction:

Liquid physical properties

λ_L

V_{SL}, V_{SG}

For a relatively smooth oil-gas interface and stratified liquids, a weighted average for the oil-water properties can be used, and if oil-water slippage could be quantified this should ideally be incorporated. Increase of V_{SG} increases the influence of the oil-gas interfacial friction factor, but this aspect cannot be accommodated in this type of method.

Again, increases of V_m and V_{SG} and so oil-water mixing and dispersion formation will mean the physical chemistry of the fluids may dominate the attained liquid viscosity, but, as discussed previously, this need not mean that a large increase in friction loss will result.

One of the most commonly used methods for pressure drop prediction at low-to-moderate liquid contents is the method of Oliemans (1976) which has been used successfully in the present study. This author employs a novel approach where he considers that there is a region of the pipe cross-section which effectively sees no flow; this fraction is dependant upon the no-slip and actual liquid holdups. From the value of in-situ holdup, provided by any holdup correlation as desired, an effective diameter is calculated and a modified two-phase friction factor is produced.

One can go on to mention a very large number of methods which have produced good comparisons with at least one set of either experimental or field pressure drop data. Apart from the situation of a three-layer flow which is more amenable to theoretical treatment (chiefly because we have 3 smooth,clean layers) than other cases, it could be said that the chief goal in pressure drop prediction is dominated by the need to know how the oil-water combination will distribute itself and the mixing intensities to which it will be subjected . A simple mixing rule for the oil-water properties can be used successfully, as the current data has demonstrated. However, this does not carry to all systems for which data is available.

8.6.2 An Examination of Water-Fraction Influence on Oil/Water/Gas Pressure Loss - Horizontal Flow

The previous section has sought to demonstrate how one can use existing two-phase methods to obtain predictions for three-phase pressure drop. This section briefly outlines where and to what extent water cut may be expected to be influential, where attention is restricted to stratified and intermittent flow.

In a 3-layer stratified flow, the success of applying a two-phase approach would seem to depend much on the oil viscosity. When the oil viscosity is not too different from that of the water, the use of the μ_0 only may provide sufficiently accurate predictions in many cases. Admittedly the shear stresses for the oil and water-wetted portion of the wall will be expected to differ more if oil-water slippage is taking place, but for horizontal lines and low-to-moderate μ_0 minimal slippage will be expected in many cases. The prospect of using the water film to lower hydraulic losses is an attractive one, however it has so far proved useful only in special circumstances. Firstly, no gas has been present, and secondly the water film has only proved influential where very high oil viscosities and laminar flow are involved (Charles 1961).

Mixed liquids in a gas-liquids stratified flow presents the dilemma of which simplifying assumption to take. The most important factors here are the V_{SG}, λ_w and the initial state of the mixture. If the liquids are initially poorly mixed, the assumption of a linear mixing rule coupled to the correlation of Oliemans (1976) should prove valid. High V_{SG} of course helps to mix the oil and water, but it is unclear just how microscopic the transported dispersion will be. Increase of V_{SG} also increases the gas-liquids interfacial friction factor and the foam generation in the system. These factors may prove far more influential than the properties of the developed oil-water mixture, and in the latter aspect the work of Assar et al (1988) provides much useful information. The effect of λ_w on the foaminess depends much on the chemical composition of the produced water, but one would initially assume that increased λ_w may reduce the mixture foaminess. Bearing in mind that in high V_{SG} flows Minami and Brill (1987) found a reduction in liquid holdup, there are obviously a number of competing mechanisms to be considered.

Where possible, if oil/water/gas flow is envisaged, guidance can be obtained from oil-water laboratory data. Mixing apparatus can be used to generate oil-water emulsions and apparent viscosity data, but even then this approach involves uncertainties as pointed out by Hartley and Bin Jaidid (1988). The correlations of Woelflin (1947) or the Richardson (1950) method can be used to give apparent liquid viscosity, although in each case engineering judgement is required. In the former, this means the choice of which curve to use, and in the latter one has to choose a value of k to use in Eq. 3.3. It is possible that for a low λ_w (less than 30%, say) and a fairly low μ_o, the effect on two-phase pressure loss of a tight emulsion will not be significant. Much depends on the proximity to the dispersion inversion region, and Laflin and Oglesby (1976) have suggested that V_m may have influence here.

Assuming that one is flowing gas with a tight water-in-oil emulsion (i.e. perceived worst case), an increase in holdup and wall shear stress at first sight seems inevitable. The following may, however, be of influence:

i) The dispersion-wall contact characteristics

ii) The effect of foaming.

The first point is very often completely ignored by pipe-flow researchers, but this is because a single component liquid is usually the study matter. Several of the large number of studies of agitated liquid-liquid flows in vessels (both batch and continuous) have shown that the vessel material and impeller material can affect where the liquid-liquid inversion occurs, in terms of dispersed phase fraction ((Selker and Sleicher (1965), Guilinger et al (1988)). In a field-pipeline the metallurgy of the clean material will not be expected to vary greatly between systems (in terms of liquid wettability), but surface contamination of the wall by corrosion or the deposition of films due to scaling, wax, or inhibiting chemicals, could conceivably affect the disperson behaviour. In this respect, the oil-water interfacial tension is an important parameter. On the microscopic level, the impingement of water or oil droplets as appropriate will lead to variable shear stress at the wall; the extent to which the droplets will adhere is a strong function of the liquid-metal contact chemistry. It is also of great influence to pipeline corrosion, as acknowledged by Lotz et al (1990).

Many of the aspects described above also exert a strong influence on the intermittent flow pressure drop. Liquids stratified or mixed will again be discussed separately.

In the separated liquids case, observations suggest that the film between the slug units can still display some oil-water mixing. Since the majority of the pressure drop is contributed by that of the slug units, this should not affect the accuracy of the calculation to a significant extent. This is especially true if one considers that in such conditions, the flow may be of relatively low mixing intensity and loose dispersions will be favoured: the time between the slugs in some cases may be sufficiently long to allow bulk re-settlement of the oil and water to occur also. The slug pressure drop is then strongly influenced by the affect, if any, of the water cut on the slug characteristics. This aspect has already been discussed but it is wise to stress its importance to the pressure loss problem. One could of course adopt the extreme approach assuming a slug holdup of unity, for example, slug frequency based upon oil-only properties and slug length increased due to 3-phase flow, where the latter trends were observed in the work of Stapelberg et al (1991) and the current work. However, these are the only current studies where three-phase slug characteristics have been measured, and in a field system a combination of all effects will exist. In all cases, however, prudence suggests that a simple mixing law for the slug front pressure drop is applicable; the mixing energy at this location was found to have a major influence on the oil-water mixing.

With an oil-water homogeneous slug flow, the influence on the slug characteristics mentioned previously are the primary factors in the pressure loss calculation. Again, one is forced to weigh up the likely behaviour of a tight emulsion as opposed to using a simple mixing law; information may not exist as to the oil-water behaviour. If the designer assumes a tight emulsion is flowing, the following should also be considered:

i) The effect of V_{SG} and foaming

ii) Shear stress in the gas bubble above the liquid film between slugs.

As well as affecting the slug characteristics, Laflin and Oglesby (1976) have proposed that in a three-phase slug flow, the gas fraction (and so V_{SG}) can modify the λ_w at which the oil-water mixture can invert. This is explained by the distribution of gas bubbles in the liquid slug interfering with the coalescense of the dispersed droplets and so the inversion mechanism.

Increased V_{SG} promotes increased foaming from previously. Water cut has a positive influence in this respect in that foaming may be lower than in an oil/gas flow, however this effect should in many cases be ignored: increased λ_w may affect slug holdup and the slug length, and these phenomena may prove more influential.

The point ii) above arises from visual observation of the present tests, though is not limited to three-phase flow. In high V_{SG} conditions the gas bubbles between slugs resulted in the upper wall being contacted with large globules and in some cases a continuous foam coating, and these effects may contribute a more significant pressure drop than has been hitherto considered in slug flow gas-liquid models.

Therefore, from a fundamental aspect, the first point to consider should be: liquids separated or mixed? In a stratified flow the mixing intensity should provide a pointer as to the increased energy provided to the liquid mixture. Such fluid property considerations may be less significant in a slug flow, where the prime factors are expected to be the water cut influence on the slug length, frequency and body holdup.

8.6.3 An Examination of Water-Fraction Influence on Oil/Water/Gas Pressure Loss - Inclined Flow

8.6.3.1 Uphill Flow

The presence of even a very small incline in the pipe has a series of repercussions which have been discussed earlier in this chapter; flow regime transitions and liquid holdup. These aspects will now briefly be considered with respect to the effect on pressure drop.

In all inclined flows, the pressure loss contains a component additional to those of friction and acceleration; the hydrostatic component. In two-phase flow this is often represented by:

$$(dp/dx)_{el} = \rho_m g \sin\theta \qquad (8.15)$$

The mixture density depends on the average liquid holdup. Oil-water slippage will then increase the mixture density somewhat, therefore the benefit of obtaining a three-phase stratified model is again evident.

As V_m is increased, as mentioned the point will be reached where the oil-water can be treated as no-slip, and indeed in such cases the friction loss may start to overwhelm the elevation pressure drop.

Uphill flow is dominated by the presence of slug flow, as already discussed. Whilst the elevation pressure drop is chiefly concerned with the in-situ holdup (and water cut), the friction loss will in some cases be more tied to the characteristics of the slug and the fluid physical properties. As a note, however, one should not forget that in large diameter flowlines the pressure drop due to friction may in certain cases be less than that due to elevation if sufficient terrain affects are present.

In a three-phase flow, any effect on the slug frequency, length and holdup will be expected to affect the pressure drop. Therefore the previous discussion on the effect of water cut on slug characteristics is applicable to this current aspect.

In terms of application of present two-phase methods, their extension to uphill flow of oil/water/gas carries the same attractions/drawbacks as already detailed in 8.5, with the additional complication of accounting for the hydrostatic loss (Eq (8.15)) component. It would be desirable to have a method to calculate this parameter, particularly for low-velocity hilly-terrain pipelines; if methods can be developed for oil/water/gas slippage this information would become one of the many useful products.

8.6.3.2 Downhill Flow

A downward incline results in a large portion of the flow map exhibiting stratified flow. When calculating pressure drop, much depends on the holdup and the character of the gas-liquid interface. The hydrostatic pressure component represents a theoretical gain of head in downhill flow : in two-phase systems, many investigators choose to ignore this pressure gain largely due to suggestions that it is not usually significant in field systems; it also then results in a more conservative prediction of head loss.

Again, much would seem to depend on the three-phase slippage if the oil and water are separated. In a two-phase flow, the liquid exhibits less slippage due to the effect of bouyancy on the gas and gravity on the film. The presence of an additional water film may produce still lower holdups if oil-water slippage takes place, which appears possible from current visual observation of downflow.

The change from a smooth to a wavy gas-liquid interface has been observed to give a sharp rise in the interfacial friction factor (Andreussi and Person (1985)) in both horizontal and inclined flow. One possible effect of oil-water slippage (and so lower holdup) could be that less damping will then be available to reduce the amplitude of these waves and therefore the reduced holdup may compete with this effect to a certain extent, which may be more important than in horizontal flow. Interfacial friction factors from two-phase studies could be modified for three-phase effects if considered appropriate.

Where the flow is gas-liquid stratified but with mixed liquids, the downhill pressure loss may be influenced more by the smooth-wavy transition then by any oil-water slippage due to the absence of water films. The earlier discussion on flow patterns touched on this matter: the characteristics may be inherently tied to the chemistry of the oil-water mixture.

As for the uphill flow, two-phase and three-phase slug flow may be expected to be dominated, so far as pressure loss is concerned, by the slug characteristics which result due to hydrodynamic and terrain effects.

Ideally, the effects of oil-water slippage should be taken into account for three-phase downhill slug flow, but it is debatable whether in many cases this will lead to much better predictions. This argument is based upon the fact that in downflow a relatively large V_{SL} is required to produce slug flow when compared to that required in uphill flow. This higher V_{SL} then provides increased mixing intensity to the oil-water combination. For low-to-moderate oil viscosity, the adoption of a simple mixing rule and oil-water no-slip assumption would seem, at first glance, to be a good approximation to take for pressure drop calculations. If however, a very viscous oil is being transported with water, oil-water slippage may be important and result in higher-than-expected in-situ oil ratios. In this case, the best approach may be to take the liquid as oil-only from the viewpoint of slug characteristics prediction, in combination with a simple mixing rule for the effective liquid viscosity, and use an existing two-phase method.

If one considers the scenario of a tightly-mixed water-in-oil dispersion, it seems unlikely that a small downward incline will have much effect on it's characteristics in terms of coalescense and breakup mechanisms, the energy of the flow being of more consequence than a weak gravity effect. The liquid density can be modified using the approach of Eq (8.15), and one would expect the dominance of the frictional component to depend on the effect of the viscosity/fluid chemistry on the total liquids holdup, the characteristics of the gas/liquid interfacial waves, and the transitions between flow regimes.

8.6.4 The Role of Fluid Physical Chemistry and Rheology in Three-Phase Pipeline Pressure Drop.

The previous section seeks to highlight what can be learned from work on gas/liquid pressure loss in terms of developing 3-phase pressure drop methods. In these discussions, and those concerning flow regime and liquid holdup, constant reference is made to the behaviour of the oil-water mixture. Whilst this is an area for fluid physical chemists, this section will briefly summarise the availability of methods and their applicability to the problem at hand.

When viewing the system microscopically rather than macroscopically, we are interested in the interaction of all 3 phases, but of primary interest here is that between the oil and water.

If the oil and water are expected to form separated layers with a third gas layer there is then minimal input from physical property modelling in that the fluids are pure. Where the oil and water form a mixture, homogeneous or otherwise ,the precise manner in which the oil and water co-exist is extremely difficult to predict, yet therein lies the best hope of achieving satisfactory predictions for dispersion transportation systems. The fundamental influences are:

i) The mixing intensity produced

ii) Droplet size and distribution

iii) Emulsion inversion behaviour.

The first aspect above will have crucial influence over the approach which will be taken for the oil-water property representation. If a low energy input is envisaged, then present data suggests that there is little benefit in disposing of a simple mixing law in favour of a more elaborate approach. It is true that the mixing intensity can be a function of the system of interest. For example, a subsea wellhead which is heavily choked back or gas-lifted may be envisaged as providing more energy to the oil-water mixture than a low flow rate, fully-open well. Therefore the development of a mixing intensity approach involving liquid flow rates, λ_w, flow regime and inherent losses (e.g chokes, formation) would be of value as a first step to ascertain the extent of these microscopic effects.

Two-phase, oil-water approaches have taken mixing intensity ϵ as:

$$\epsilon = \frac{2fV^3}{d_p} \qquad (8.16)$$

where f is the friction factor, which then requires the pressure loss to be known *a priori* unless simplifying assumptions are made.

The mixing intensity and chemistry of the oil-water system will to a large extent determine the droplet size and distribution which will equilibrate in the pipeflow, so determining the tightness of the dispersion. This is widely regarded as a fundamental influence in all forms of agitated liquid-liquid studies. Whilst being able to accurately predict mean droplet size would be useful, it is those droplets which contribute least to the average diameter which ultimately influence the mixture apparent viscosity. The work of Hinze (1955) and many workers since, on the maximum stable droplet size in a turbulent flow, can be used to give a first approximation. The drawback here is that it is uncertain how these methods will scale up to even an oil-water only flow with a large (greater than 2% (vol)) concentration of the dispersed phase. This is due to much higher droplet-droplet interactions, particularly the effects of coalescence. At a first pass, it seems plausible to neglect such interactions with the view that this will lead to the most conservative result i.e. smaller droplets.

The droplet coalescence and breakup mechanisms are also fundamental to the system inversion behaviour, and as such influence the mixture viscosity. The water fraction and presence of surfactants will to a large degree influence the inversion region, but in some systems the apparent viscosity may not rise appreciably until a large λ_w is attained: in such cases again the use of either a simple mixing law or the single phase oil viscosity may suffice. In saying this, it should also be remembered that the total V_{SL} may influence the inversion region.

The effect of a gas phase on the emulsion characteristics and pressure loss will be expected to be a function of flow pattern. In a gas-liquid stratified flow, the influence of the gas will be in the energy input to the oil-water mixture which then may alter the droplet size distribution. Therefore migration of gas bubbles through the liquid media is not expected, although obviously there is an influence at the gas-liquid interface. A greater effect on the dispersion characteristics may be expected in regimes such as intermittent or dispersed bubble, which are liquid dominant. The presence of gas in the dispersion can be envisaged as providing interference between the dispersed liquid droplets, where the following may be important:

i) Bubble size distribution

ii) Interfacial areas of the dispersed bubbles.

Again, methods are available to calculate the maximum stable droplet size in turbulent flow (e.g. Levich (1962)), and a turbulence intensity factor ϵ is required.

At the limit, if the bubbles provide sufficient interfacial area, the inversion λ_w may be different to the oil-water situation. This could then favour lower pressure drops if the inversion λ_w is shifted to considerably higher values.

Whilst it has largely been ignored by investigators of both two-phase and three-phase flow, the presence of another dispersion, foam, should not be overlooked. Foam generation is favoured by high V_{SG} flows, and so wavy flow, annular, mist and slug flows may well display considerable foam generation. Foammability from an oil-gas system would be expected to be higher than that for a water-cut system, presuming that the water contains no components which promote foaming. The possible effect on liquid holdup noticed by Minami and Brill (1987) should be borne in mind, as should the large-diameter slug flow calculations arising from Brill et al (1985) which were taken where the crude oil was reportedly very foamy. So, despite the fact that foaming has not received the attention warranted in gas/liquid flow, the presence of water may mitigate the foam generation. This will not always produce benefits e.g. the slug body holdup may well be higher in the water-cut case.

Finally, one should mention another effect which is a product of fluids chemistry and the fluid dynamics : rheology. Information exists, for both emulsions and foams, that points to the existance of non-Newtonian flow in many cases. It should be noted that existing two-phase methods, whether for pressure loss, holdup or flow regime, assume the fluid behaviour to be Newtonian. The literature suggests that water-in-oil emulsions will produce non-Newtonian behaviour more readily than do oil-in-water emulsions, although this depends on the base oil viscosity. Mostly pseudoplastic (shear-thinning) behaviour has been reported. Foams have also been reported as producing pseudoplastic behaviour (Assar et al (1988)).

Many workers have found that the flow of gas with a shear-thinning suspension produces a minimum in the pressure loss at particular values of V_{SG}. The above then indicates that the examination of the fluid rheology should be considered an important exercise in oil/water/gas systems.

CHAPTER 9

CONCLUSIONS AND RECOMMENDATIONS

The purpose of this study has been to experimentally extend the knowledge of oil/water/gas pipe flow and in the light of this new data and earlier data examine the important influences with respect to developing design methods for the various fluid-dynamic aspects. It is now appropriate to present the study conclusions and, importantly, suggest where future effort could be expanded, both experimentally and theoretically, in order to gain better understanding of what is a very complex fluid mechanics - fluid physical chemistry problem. The conclusions and suggestions for further work are assigned separate sections.

9.1 CONCLUSIONS

Chapter 2 has illustrated that, in the development of any marginal oilfield, any methods which can reduce the capital expenditure associated with the project can have a significant impact on the feasibility of development and the project profitability. The elimination of the requirement for a full fixed production platform to one where either a minimal-processing platform or subsea tie-back to an existing platform is installed can then influence field development feasibility: in each case the transporation of unprocessed well fluids is expected. An improvement of the knowledge of oil/water/gas pipeflow is then important in realising this objective.

Chapter 3, involving a review of many aspects involved in a three-phase flow, suggests the following:

i Whilst many of the aspects of three-phase flow have been investigated, they have involved non-pipe flows and been restricted to certain fluid physical properties where an insufficient departure from datum values has been performed.

ii Specific studies of oil/water/gas flow have been limited in the range of flow rates and fluid properly effects, and in some cases are of questionable accuracy.

The experiments performed on oil/water horizontal flow allows the following conclusions to be drawn:

i Flow regimes were similar to those observed in earlier experimental systems and were found to be similar for the two oil viscosities involved in the current study.

ii The oil-water behaviour was found to be no-slip in all of the flow regimes obtained.

iii For homogeneous flow, a slight decrease in the pressure drop was measured at an input water cut of approximately 40%.

iv In no case was a peak in the pressure drop obtained at any water fraction.

The experimental observations of oil-water flow were important prerequisites for the main study objective, that of experimental data collection of oil/water/gas flow in a horizontal 50mm i.d. test loop. This exercise produced the following conclusions:

i Bulk gas-liquid flow regimes are similar in oil/gas and oil/water/gas flows in the same pipe.

ii The degree of oil-water mixing obtained is dependent on several parameters, the most important of which appears to be the superficial gas velocity.

iii For a constant liquid superficial velocity, the flow pattern transitions in oil/water/gas flow were found to be similar to that of the oil/gas flow.

iv Comparison with an existing two-phase flow pattern model modified with a linear mixing law for the oil/water properties, gave similar accuracy for the oil/water/gas data as for the oil/gas data.

v The transition to annular flow was poorly predicted by the flow pattern model.

vi For constant liquid superficial velocity, total liquids holdup fraction was not found to be dependent upon input water fraction, although the data does exhibit some scatter.

vii Oil-water behaviour was found to be essentially no-slip for all flow conditions.

viii Comparison with a modified existing two-phase holdup correlation displayed good agreement except for low liquid superficial velocities.

ix Except for the lowest liquid superficial velocity, pressure drop was observed to show a slight decrease at input water fractions of approximately 40%.

x No peaks in the pressure drop-water cut relationship were obtained at any fixed liquid superficial velocity.

xi Comparison with two existing two-phase pressure loss correlations gave adequate results, with consistent overpredictions being obtained with one of the methods.

xii Based on a very small amount of data, it appears that slug frequency and length may be slightly modified when water is present with the oil as the liquid phase.

xiii The flow regime, liquid holdup and pressure drop trends were similar for oils of 1.8 cP and 4 cP viscosity.

From the experiments of oil/gas and oil/water/gas flow in a 50mm i.d. pipe, inclined in a 1-degree uphill-downhill combination, the following conclusions are drawn:

i The slug flow envelope uphill and stratified flow envelope downhill are both expanded to similar extents for the oil/gas and oil/water/gas tests.

ii The gas superficial velocity required to mix the oil and water was not significantly affected by the inclination involved.

iii Prediction of the flow regime by a modified existing two-phase model gave similar success for oil/gas and oil/water/gas tests.

iv Uphill total liquids fraction was slightly higher than the horizontal values at low liquid superficial velocities, but as these were increased the difference was diminished.

v In uphill flow oil-water slippage was obtained at the low superficial liquid velocities.

vi Inspection suggests that total liquid holdup in the uphill flow can be influenced by the input water fraction.

vii Satisfactory predictions for uphill total liquids holdup was obtained from two existing, modified two-phase correlations for moderate-to-high superficial liquid velocity.

viii Results suggest that uphill and downhill inclines can produce different effects regarding the influence of input water cut on the pressure drop.

ix The use of modified two-phase pressure loss correlations indicated that uphill they were less conservative than downhill, but this is partly due to the use of a horizontally-derived liquid holdup correlation.

x As for horizontal flow, pressure loss predictions by the Eaton-Oliemans method were generally less conservative than those from the Eaton-Dukler correlations.

xi For the uphill flow, slug frequencies were higher than for horizontal flow, with oil/water/gas slug frequency being slightly higher than that for oil/gas flow.

The final activity has entailed examining the different flow regimes possible in oil/water/gas flow, together with their associated liquid holdup and pressure drop characteristics. The current data and existing data has been viewed in perspective to where current two-phase modelling work can be utilised for particular three-phase calculations, and also to point out where new approaches are probably required. The following comments should be made as regards this activity:

i The choice of which approach to take is fundamentally dependent on whether the oil and water will mix or whether they will remain as essentially separated layers.

ii Input water fraction may be an important parameter in determining the location of the stratified-slug transition in a liquids-separated flow.

iii In all hydraulics aspects, the fluid properties of the oil-water mixture, particularly if they are dispersed, are expected to play a central role.

iv The formation of a homogeneous oil-water mixture need not necessarily imply the formation of a viscous emulsion, and the input mixing intensity is expected to be a governing parameter.

v The formation, transportation properties and microscopic properties of liquid-liquid and gas-liquid dispersions may play a critical role in high flow rate flow patterns.

vi Pipe inclination is a factor which must always be borne in mind, especially if low flow rates and separated liquids are involved.

9.2 RECOMMENDATIONS FOR FUTURE WORK

Following from the conclusions of the previous section, the present work, taken together with that already available from the literature, allows one to view more objectively the possible further fruitful areas of work which may be of high influence in gaining a better understanding of the problem, both qualitatively and quantatively. The following activities would be of great value:

i Measurement of in-situ water cut and pressure loss in horizontal and inclined oil-water flows where the water cut, oil viscosity, pipe diameter and material can be systematically varied.

ii Experimental work examining the effect of water cut on the stratified-intermittent transition and slug characteristics where pipe inclination and pipe diameter effects can also be investigated.

iii Measurement of droplet size distribution and pressure drop in an oil/water/gas flow where the liquid phase is a tight dispersion and where the effects of foaming can be controlled.

iv The development of a laboratory technique to simulate oil/water/gas flowing mixing intensities and application of effective viscosity data to modified two-phase correlations to obtain predictions for iii above.

v Development of a three-layer separated flow model for low-velocity, inclined oil/water/gas flow.

vi The collection of oil/water/gas field data.

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APPENDIX A

OIL SELECTION TESTS

A1.1 Initial Oil Screening

The change of test oil meant that the new oil had to fulfil the operational and technical requirements that Oil No 1 had passed, that is:

- i The oil should have a high flash point
- ii The oil should be non-hazardous
- iii The oil should not contain a high aromatics content
- iv The oil could be marked with the red tax powder

In addition, the Oil No 2, whilst being more viscous, ideally should have similar density, surface tension and interfacial tension over water when compared to the properties of Oil No 1. Finally, there was some influence of the cost of the oil, given that 2000 imp.gals would be purchased.
Approximate specifications were sent to oil suppliers and initially data sheets were obtained which gave the oil physical properties (except surface and interfacial tensions), environmental data and typical costs. Candidate oils, suitable from this first screening, were then supplied in 5 litre samples. This enabled in-house measurements of density,viscosity and, in particular, surface and interfacial tensions. It was found that several oils displayed the required behaviour in terms of density, viscosity and surface tension, but not interfacial tension. Finally, 2 oils were put forward for final screening, denoted Oil A and Oil B hereafter: these oils had interfacial tensions over water of approximately 40 mN/m as compared to the Oil No1 value of 32.5 mN/m.

A1.2 Final Selection Tests

A further series of tests were conducted on the candidate test oils. These tests were designed to show any practical problems which could arise from using the oil in the facility, and to give useful information on the new oils' behaviour in terms of foaming and oil-water dispersion and provide some comparison with existing Oil No1. An agitated liquid-liquid mixing facility was designed and constructed, which is depicted on Figs A1.1 and A1.2. The rig consists of a baffled 30cm diameter glass vessel, equipped with a 6-blade turbine impeller, electric motor and speed read-out. Gas injection to the mixer was provided via a pressurised gas bottle and a calibrated rotameter measured the gas input rate.

A1.3 Foaming Tests

The first series of tests involved foaming in the oil/gas system (i.e. no water). No attempt was made to conduct a large series of tests through changing a number of vessel parameters; the tests were simply intended to give comparisons with the Oil No1 benchmark and identify potential problems such as excessive foaming or stable foams.

The foaming tests consisted of filling the vessel with a fixed volume of the oil,which was usually about half the vessel volume, setting the impeller position and for a small number of impeller speeds passing a number of gas flow rates to the agitated liquid system. The equilibrium foam volume above the liquid-foam interface was then noted at each condition, where a sufficient period was allowed between each gas rate for the foam to attain steady-state conditions. This approach is consistent with numerous earlier studies such as Ohkawa et al (1984) and Callaghan et al (1985) which present foaminess data of a very wide range of fluid system properties. Fig A1.2 shows a typical test involving Oil No1. In addition to Oil No1 the test oils A and B were used as was dead BP Forties crude,the latter to give a preliminary idea of the test oil foaminess when compared to a typical crude oil. In all cases, air was used as the gas phase to preserve the consistency of the test with regard to scale-up to the flowing tests. Also, the oils were saturated with the microbiocide-dosed water before testing, for the same reason. Table A1-1 shows the data and Fig A1.3 a graphical plot of the data at impeller speed 400rpm. Examination of these data show that at all impeller speeds the crude oil was by far the most foamy of the oils tested. The Oil No1 is found to be more foamy than either of the oils A or B; it is considered that prolonged exposure to the tank water and so the microbiocide may in part be responsible for this difference. During pipe facility commissioning, it was noticed that Oil No1 was slightly less foamy in the oil-gas case than later when the water was introduced, supporting the above suggestion that longer exposure periods may actually produce a higher oil foaminess. Fortunately, in all cases the foams formed were unstable: this was inferred from the timing of the decay of the foam layer when the gas input and agitation had stopped. Therefore, so far as the candidate test oils were concerned, no operational problems due to foaming were expected.

A1.4 Oil-Water Dispersion Tests

Using the same apparatus, tests were conducted on stirred oil-water mixtures, mainly to identify any problems regarding separation of the oil and water, but also to obtain knowledge of the types of dispersions formed and inversion behaviour in a liquid/liquid mixing system.

The study of oil-water dispersions in mixing vessels has been the subject of a very large number of research studies. The present experiments were not intended to add to this knowledge base, but to give practical pointers to the convenience and suitability of the new test oil. In all tests the impeller location was fixed, and the vessel was filled with an initial oil volume of 6 litres. Test were conducted using Oil No1 and oils A and B initially to observe the ease of separation following agitation, although not severe agitation. The oils were mixed with water at 300rpm for 5 minutes duration, and the time for bulk oil-water separation was noted. It was observed that Oil B, even after 30 minutes did not give adequate oil-water separation. For this reason no further testing of this oil was made and therefore Oil A was chosen as the new facility oil. However, further tests were considered valuable to compare the oil-water emulsification of Oil No1 and Oil A.

Tests were made to establish the inversion behaviour of the oils in systems of identical geometry and construction, and to examine the feasibility of using a conductivity probe to determine the emulsion dispersed phase.

Visual observations can identify which of the oils is dispersed: after agitation ceases the manner in which the liquids settle and the appearance of the bulk layers is different depending on whether the dispersion is oil-in-water or water-in-oil. This method was used by a number of researchers including Selker and Sleicher (1965). The method was supplemented by the readings from a conductivity probe. This method is also commonly used to identify the dispersed phase; the workers above and the recent study by Guilinger et al (1989) employed the technique. When oil is the continuous phase, the mixture behaves essentially as an insulator. However, when water becomes the continuous phase following inversion, a sharp increase in conductivity is obtained. Table A1-2 gives the data for the Oil No1 and Oil A tests. Inversion for similar geometric conditions occurred at λ_w = 0.37 for Oil No1 and at λ_w = 0.34 for Oil A. It should also be remembered that the mixing vessel was made of glass: this was again deliberate to be consistent with the material of construction of the test pipe. The results suggest that the inversion behaviour of the different test oils will not be radically different.

A1.6 Concluding Comments

These tests have enabled several comments to be made regarding the adoption of the new test oil. The choice of Oil A as the new test oil (Oil No2) has given confidence as to the oils' technical and practical suitability. In addition, the following is relevant:

- i The new oil may be less foamy than Oil No 1.
- ii Oil-water dispersion behaviour is not expected to be greatly different to that of Oil No1.
- iii The use of a conductivity probe for dispersion inversion detection should be attempted in the oil-water and oil/water/gas flowing tests.

APPENDIX B

EQUIPMENT CALIBRATION CERTIFICATES

Turbine Flowmeter Calibrations
Horizontal, 2" Kerosene Tests
D80

QUADRINA LIMITED

FLUID FLOW MEASUREMENT AND CONTROL SYSTEMS

Flint Road, Letchworth, Hertfordshire, SG6 1HS.England
Telephone: Letchworth 673486, Telex: 826726 QDRINA G

TEST CERTIFICATE

Order no. 45945/JK Works Order no. 10170

Customer B.H.R.A.

Flowmeter type no. QEG13B/EP1 Serial no. 10839 Tag no. –

Pickup coil type ELECTRONIC Calibration fluid Air

Calibration conditions: T ....°C P S.T.P. psig Viscosity ......cSt s.g.........

Calculated flow rate m3/min	Meter Output Frequency Hz	Pulses per metre3
0.1360	1615.11	712548
0.1028	1223.78	714271
0.095071	1136.81	717446
0.079318	950.76	719202
0.071615	861.18	721509
0.064201	774.29	723623
0.055684	671.73	723794
0.039203	472.35	722929

$f(Hz) = 199\ Q(m^3/h)$

Mean factor = 719415 ppm3

Rebuild flowmeter

On Box 5V= 1622Hz

Date 10.2.89

QUADRINA LIMITED

FLUID FLOW MEASUREMENT AND CONTROL SYSTEMS

Flint Road, Letchworth, Hertfordshire, SG6 1HS.England
Telephone: Letchworth 673486, Telex: 826726 QDRINA G

TEST CERTIFICATE

Order no. 42843 Works Order no. 9412

Customer B.H.R.A.

Flowmeter type no. QFG/25B/B/EP1 Serial no. 10838 Tag no. -

Pickup coil type Electronic Calibration fluid Air

Calibration conditions: T ....°C P S.T.P. psig Viscosity ......cSt s.g.........

Calculated flow rate m^3/min	Meter Output Frequency Hz	Pulses per $metre^3$
0.794	1679.89	127199.9
0.5369	1140.43	127446.08
0.3964	843.22	127631.69
0.2848	605.63	127590.59
0.2069	441.01	127890.77
0.1454	308.10	127138.93
0.1252	265.36	127169.33
0.1025	216.62	126801.95

$f(Hz) = 35.377\ Q(m^3/h)$

Mean factor 127358.66 ppm^3

On scale, 5V = 2123 Hz

Date 7.12.87

QUADRINA LIMITED

FLUID FLOW MEASUREMENT AND CONTROL SYSTEMS

Flint Road, Letchworth, Hertfordshire, SG6 1HS.England
Telephone: Letchworth 673486, Telex: 826726 QDRINA G

TEST CERTIFICATE

Order no. 46493/JK Works Order no. 10279

Customer B.H.R.A.

Flowmeter type no. QFG/75B/EP1 Serial no. 10837 Tag no. -

Pickup coil type ELECTRONIC Calibration fluid AIR

Calibration conditions: T ....°C P S.T.P. psig Viscosity ......cSt s.g.........

Calculated flow rate M3/min	Meter Output Frequency Hz	Pulses per metre3
9.108	1685.75	11105.10
7.577	1403.63	11114.94
6.150	1135.47	11077.80
4.538	836.17	11055.62
3.140	580.44	11091.34
1.550	286.69	11097.86
1.482	273.73	11082.43
1.121	205.24	10985.13
0.7924	144.82	10965.80
0.5369	98.01	10952.62

$f\,(Hz) = 3.087\,Q\,(m^3/h)$

Date 20.4.89

Date ..12.87

(1)

QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QEL/10/MP2 Serial number; 10835

Customer; B.H.R.A. W.O. number; 9412

Order number; 42843 Tag number; -

Calibration conditions:
Fluid; WATER Pressure; 60 psig Temperature; 19 celsius
s.g.; 0.9982 Viscosity; 1 cSt Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	10.0	27012	21.70	12.575	1244.79	5938.94
2	10.0	26976	24.40	11.184	1105.57	5931.02
3	10.0	26952	29.60	9.219	910.54	5925.75
4	10.0	26984	34.10	8.002	791.31	5932.78
5	10.0	26977	41.10	6.639	656.37	5931.24
6	10.0	26925	53.20	5.129	506.10	5919.81
7	10.0	26963	60.80	4.488	443.47	5928.17
8	10.0	26978	73.50	3.712	367.04	5931.46
9	10.0	27048	105.20	2.594	257.11	5946.85
10	10.0	27066	141.70	1.925	191.00	5950.81

$f(Hz) = 1648\, Q\, (m^3/h)$

On Box, 5V = 1236 Hz

weight-lbs; time-seconds; rate-litre/min; factor-pulses per litre;

Mean factor is 5933.688 pulses per litre

Calibrated by; P.A.FARRELL : Checked; : Date; 15.12.87

Notes;

program in BASIC QF084/88

2

QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QEL/16B/MP2 Serial number; 10836

Customer; B.H.R.A. W.O. number; 9412

Order number; 42843 Tag number; -

Calibration conditions:
Fluid; WATER Pressure; 60 psig Temperature; 19 celsius
s.g.; 0.9982 Viscosity; 1 cSt Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	100.0	38324	30.20	5.421	1269.00	842603.64
2	100.0	38371	35.80	4.573	1071.81	843637.00
3	100.0	38334	46.10	3.551	831.54	842823.51
4	50.0	19142	26.00	3.148	736.23	841724.19
5	50.0	19122	31.50	2.599	607.04	840844.74
6	50.0	19116	38.20	2.143	500.41	840580.90
7	50.0	19096	44.70	1.831	427.20	839701.45
8	50.0	19120	60.20	1.359	317.60	840756.79
9	50.0	19148	80.80	1.013	236.98	841988.03
10	50.0	19150	131.60	.622	145.51	842075.97

$f\,(Hz) = 233.798\,Q\,(m^3/h)$

On Box, 5V = 1286 Hz

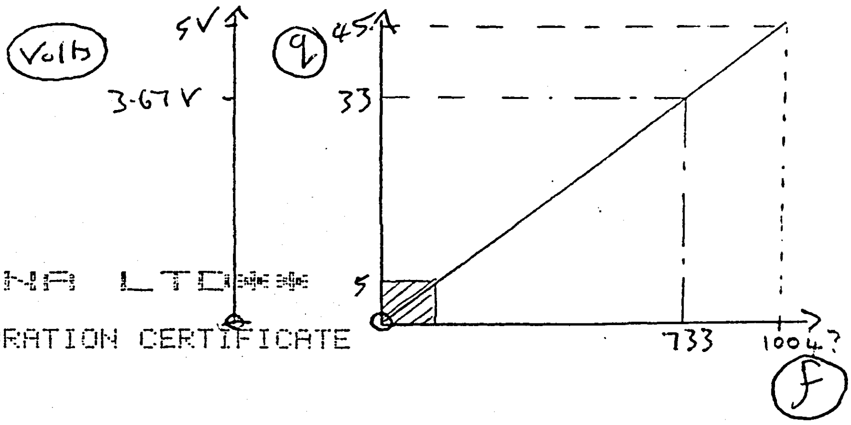
weight-lbs; time-seconds; rate-cu m/h; factor-pulses per cu metre;

Mean factor is 841673.626 pulses per cu metre

Calibrated by; P.A.FARRELL : Checked; : Date; 15.12.87

Notes;

3



QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QFL/38J/B/MP2 Serial number; 10834

Customer; B.H.R.A. W.O. number; 9412

Order number; 42843 Tag number; -

Calibration conditions:

Fluid; WATER	Pressure; 60 psig	Temperature; 19 celsius
s.g.; 0.9982	Viscosity; 1 cSt	Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	500.0	18319	25.00	32.747	732.75	80553.47
2	500.0	18309	29.70	27.565	616.46	80509.49
3	500.0	18241	33.20	24.659	549.42	80210.48
4	500.0	18241	37.20	22.007	490.34	80210.48
5	500.0	18232	43.30	18.907	421.06	80170.90
6	500.0	18252	50.50	16.211	361.42	80258.85
7	500.0	18269	63.60	12.872	287.24	80333.60
8	500.0	18260	78.50	10.429	232.61	80294.03
9	500.0	18287	110.40	7.415	165.64	80412.75
10	500.0	18295	167.20	4.896	109.41	80447.93
11	500.0	18215	215.40	3.800	84.56	80096.15

$f(Hz) = 22.311\ Q(m^3/h)$

On Box, 5V = 1004Hz

weight-lbs; time-seconds; rate-cu m/h; factor-pulses per cu metre;

Mean factor is 80318.017 pulses per cu metre

Calibrated by; P.A.FARRELL : Checked; : Date; 15.12.87

Notes;

CALIBRATION CERTIFICATE (1)

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	229251
Range:	1 Bar d
Supply:	10V dc
Sensitivity:	99.36mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

CALIBRATION CERTIFICATE (2)

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	229250
Range:	1 Bar d
Supply:	10V dc
Sensitivity:	99.40mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

CALIBRATION CERTIFICATE (6)

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	233191
Range:	700mBar d
Supply:	10V dc
Sensitivity:	99.98mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	0 to 50°C
Temperature error band:	

CALIBRATION CERTIFICATE (10) Air TM (high press)

Calibration date:	15.7.1987 WO: E02609
Transducer type:	PDCR 10/35L ✓
Serial Number:	206275
Range:	3.5 bar d
Supply:	10V D.C.
Sensitivity:	100.52mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

Temperature error band: ±1.5%

CALIBRATION CERTIFICATE (5)

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	233194
Range:	700mBar d
Supply:	10V dc
Sensitivity:	98.96mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	0 to 50°C
Temperature error band:	± 0.5%

CALIBRATION CERTIFICATE (9) w/w (N2) ✓

Calibration date:	24.3.1988 WO: E06257
Transducer type:	PDCR 120/35WL
Serial Number:	229647
Range:	2 bar d
Supply:	10 Volts
Sensitivity:	100.38mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE (10) Air TM (low press) ✗

unused

Calibration date:	15.12.1987 WO: E06257
Transducer type:	PDCR 10/35L
Serial Number:	212739
Range:	2 Bar d
Supply:	10V dc
Sensitivity:	101.41mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

Druck Pressure Transducer Calibrations

Horizontal, 2" Kerosene Tests

(= on loop

(= off - loop

[redacted]

ALTOMETER

Flowmeter productionfacility of Rheometron A.G. Basel

ALTOMETER
Krausstraat 14 -18
3364 AD Sliedrecht
HOLLAND

19232 H

calibration certificate | Kalibrierzertifikat

Com.nr: 874791 A

Primary head:		Converter:	
Type	: K280	Type	: SC80A
DN	: 10 mm	Field frequency	: 1/6*f line,code 0
Flanges	: Sandwich	Power supply	: 240 V, 50 Hz
Test pressure	: 60 Bar	Signal output	: 4 - 20 mA
Liner	: Al_2O_3		Rl<= 700 Ohm
Electrode constr.	: Standard	Measuring range	: 0- 0.2827 m3/h
Electrode material	: Platinum	Equals	: 1.0000 m/s
Iso class	: E	Setting PC	: L5.149
Protection class	: IP66	Frequency output	: 0- 3600000 pls/h
		SMU (I+F) set on 1 %	

The primary head has been calibrated against a fixed-volume tank.
This tank is guaranteed by the Dutch office of Measures and Weights
'De Dienst van het IJkwezen'.
Uncertainty in the volume of the tank is +/-0.02%.

The calibration fluid water, has a conductivity of about 300 uS/cm and a temperature of about 16 'C.
According to DIN 1944 an inlet section of 5D and an outlet section of 3D, measured from the electrode axis, with undisturbed flow is recommended.
The calibrations were carried out under these conditions, or better.

The calibrations:

Error limits:

Flow velocity	Max. error
Full scale range >=1 m/s:	
10 - 100%	+/-1% of actual flow
0 - 10%	+/-0.1% of full scale
Full scale range <1 m/s:	
0.1 m/s - full scale	+/-1% of actual flow
0 - 0.1 m/s	+/-0.001 m/s

INSTALLED
9/88

Sliedrecht, 27-1-1988

Cal. measuring range (=100%): 0.5655 m3/h = 2.0000 m/s

Range in %	Deviation in %
95	+/-0
50	-0.14

IJK AFDELING ALTOMETER SLIEDRECHT-HOLLAND

ALTOMETER
Flowmeter productionfacility of Rheometron A.G. Basel

ALTOMETER
Krausstraat 14 -18
3364 AD Sliedrecht
HOLLAND

19232 H

calibration certificate | Kalibrierzertifikat

Com. nr: 874791 B

Primary head:		Converter:	
Type	: K280	Type	: SC80A
DN	: 40 mm	Field frequency	: 1/6*f line, code 0
Flanges	: Sandwich	Power supply	: 240 V, 50 Hz
Test pressure	: 60 Bar	Signal output	: 4 - 20 mA
Liner	: Al_2O_3		Rl<= 700 Ohm
Electrode constr.	: Standard	Measuring range	: 0- 4.524 m3/h
El- trode material	: Platinum	Equals	: 1.000 m/s
Iso class	: E	Setting PC	: L5.285
Protection class	: IP66	Frequency output	: 0- 3600000 pls/h
		SMU (I+F) set on 1 %	

---

he primary head has been calibrated against a mastermeter, which is
roven regularly against a fixed-volume tank.
his tank is guaranteed by the Dutch office of Measures and Weights
De Dienst van het IJkwezen' (Uncertainty in the volume of the tank is
+/-0.02%).
ncertainty of the mastermeter is +/-0.1%.

he calibration fluid water, has a conductivity of about 300 uS/cm
nd a temperature of about 16 'C.
ccording to DIN 1944 an inlet section of 5D and an outlet section of
D, measured from the electrode axis, with undisturbed flow is re-
ommended.
he calibrations were carried out under these conditions, or better.

---

he calibrations:

rror limits:

low velocity	Max. error
ull scale range >=1 m/s:	
0 - 100%	+/-1% of actual flow
- 10%	+/-0.1% of full scale
ull scale range <1 m/s:	
1 m/s - full scale	+/-1% of actual flow
- 0.1 m/s	+/-0.001 m/s

INSTALLED
9/88

Sliedrecht, 21-1-1988

ull measuring range (=100%): 9.048001 m3/h = 2.0000 m/s

an	Deviation in %
	-0.01
ou	+0.01

AFDELING ALTOMETER
IJK
SLIEDRECHT-HOLLAND

Turbine Flowmeter Calibrations
2", +1° D80 Tests
(QE613, QFG75, QE110 as before)
NOT USED

QUADRINA LIMITED

FLUID FLOW MEASUREMENT AND CONTROL SYSTEMS

Flint Road, Letchworth, Hertfordshire, SG6 1HS. England
Telephone: Letchworth 673486, Telex: 826726 QDRINA G

TEST CERTIFICATE

Order no. 47974 Works Order no. 10568

Customer B.H.R.A.

Flowmeter type no. QFG/25B/B/EP1 Serial no. 10838 Tag no. -

Pickup coil type Electronic Calibration fluid AIR

Calibration conditions: T ....°C P S.T.P. psig Viscosity ......cSt s.g.........

Calculated flow rate m^3/min	Meter Output Frequency Hz	Pulses per m^3
0.7924	1706.47	129213
0.5369	1156.83	129279
0.3964	854.08	129275
0.2848	610.80	128679
0.2069	444.13	128796
0.1454	311.14	128394
0.1252	267.60	128246
0.1025	217.23	127160

Mean factor = 128630 ppm^3

Date 19.10.89

QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QEL/16B/MP2 Serial number; 10836

Customer; B.H.R.A. W.O. number; 10538

Order number; 47770/JK Tag number; -

Calibration conditions:
Fluid; WATER Pressure; 60 psig Temperature; 18.9 celsius
s.g.; 0.9982 Viscosity; 1 cSt Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	100.0	37754	31.45	5.206	1200.44	830083.90
2	50.0	18828	17.65	4.638	1066.74	827929.21
3	50.0	18829	20.47	3.999	919.83	827973.18
4	50.0	18812	23.28	3.516	808.07	827225.64
5	50.0	18824	26.56	3.082	708.73	827753.31
6	50.0	18836	34.51	2.372	545.81	828280.99
7	50.0	18832	40.31	2.030	467.17	828105.10
8	50.0	18843	50.85	1.609	370.56	828588.81
9	50.0	18895	82.19	.996	229.89	830875.42
10	50.0	18856	125.61	.651	150.11	829160.46

weight-lbs; time-seconds; rate-cu m/h; factor-pulses per cu metre;

Mean factor is 828597.606 pulses per cu metre

Calibrated by; P.A.FARRELL : Checked; : Date; 5.10.89

Notes;

 program in BASIC QRAP4/80

****QUADRINA LTD****

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QFL/38J/B/MP2 Serial number; 10834

Customer; B.H.R.A. W.O. number; 10538

Order number; 47770/JK Tag number; -

Calibration conditions:
Fluid; WATER Pressure; 60 psig Temperature; 18.6 celsius
s.g.; 0.9982 Viscosity; 1 cSt Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	300.0	11007	15.02	32.702	732.82	80672.64
2	300.0	11004	19.12	25.689	575.52	80650.65
3	300.0	11005	22.53	21.801	488.45	80657.98
4	300.0	11001	23.97	20.491	458.94	80628.66
5	300.0	10997	28.36	17.319	387.76	80599.35
6	300.0	11003	35.79	13.724	307.43	80643.32
7	300.0	10998	45.02	10.910	244.29	80606.68
8	300.0	11009	55.78	8.805	197.36	80687.30
9	300.0	11015	75.49	6.506	145.91	80731.27
10	300.0	11012	106.91	4.594	103.00	80709.28
11	300.0	10948	201.76	2.434	54.26	80240.21

weight-lbs; time-seconds; rate-cu m/h; factor-pulses per cu metre;

Mean factor is 80620.671 pulses per cu metre

Calibrated by; P.A.FARRELL : Checked; : Date; 5.10.89

Notes;

 program in BASIC QRAP4/80

CALIBRATION CERTIFICATE 1

Calibration date:	15.7.1987 WO: E02609
Transducer type:	PDCR 10/35L
Serial Number:	206277
Range:	3.5 bar d
Supply:	10V D.C.
Sensitivity:	98.98mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE 2

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	229251
Range:	1 Bar d
Supply:	10V dc
Sensitivity:	99.36mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

CALIBRATION CERTIFICATE 3

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	229250
Range:	1 Bar d
Supply:	10V dc
Sensitivity:	99.40mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

CALIBRATION CERTIFICATE 5

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	233194
Range:	700mBar d
Supply:	10V dc
Sensitivity:	98.96mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	0 to 50°C
Temperature error band:	± 0.5%

CALIBRATION CERTIFICATE 6

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	233191
Range:	700mBar d
Supply:	10V dc
Sensitivity:	99.98mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	0 to 50°C
Temperature error band:	[illegible]

CALIBRATION CERTIFICATE 9 w/w (N2)

Calibration date:	24.3.1988 WO: E06257
Transducer type:	PDCR 120/35WL
Serial Number:	229647
Range:	2 bar d
Supply:	10 Volts
Sensitivity:	100.38mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE 10

Air TM (high press)

Calibration date:	15.7.1987 WO: E02609
Transducer type:	PDCR 10/35L
Serial Number:	206275
Range:	3.5 bar d
Supply:	10V D.C.
Sensitivity:	100.52mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE

Unused

Air TM (low press)

Calibration date:	15.12.1987 WO: E06257
Transducer type:	PDCR 10/35L
Serial Number:	212739
Range:	2 Bar d
Supply:	10V dc
Sensitivity:	101.41mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	[illegible]

Druck Pressure Transducer Calibrations

Early +1°, 2" Kerosene Tests

= on-test loop

= off-loop

Turbine Flowmeter Calibrations
Horizontal, 2" Lube Oil Tests
(QEG13, QFG75 not used)

QUADRINA LIMITED

FLUID FLOW MEASUREMENT AND CONTROL SYSTEMS

Flint Road, Letchworth, Hertfordshire, SG6 1HS. England
Telephone: Letchworth 673486, Telex: 826726 QDRINA G

TEST CERTIFICATE

Order no. 49220/JK Works order no. 10885

Customer B.H.R.A.

Flowmeter type no. QFG/25B/B/EP1 Serial no. 10838 Tag no. -

Pickup coil type ELECTRONIC Calibration fluid AIR

Calibration conditions: T .... °C P S.T.P. psig Viscosity .....cSt s.g........

Calculated Flow Rate m^3/min	Meter Output Frequency Hz	Pulses per m^3
0.7924	1713.87	129773
0.5369	1161.85	129839
0.3964	857.76	129832
0.2848	614.29	129414
0.2069	445.39	129161
0.1454	312.02	128758
0.1252	268.36	128608
0.1025	218.16	127702

Mean factor = 129136 ppm^3

Date 10.5.90

QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QEL/10/MP2 Serial number; 10835

Customer; BHRA W.O. number; 10823

Order number; 48963/JK Tag number; -

Calibration conditions:
Fluid; WATER Pressure; 60 psig Temperature; 16.7 celsius
s.g.; 0.9982 Viscosity; 1 cSt Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	10.0	26904	22.29	12.238	1206.99	5917.24
2	10.0	26908	26.40	10.333	1019.24	5918.12
3	10.0	26907	29.37	9.288	916.13	5917.90
4	10.0	26905	32.83	8.309	819.52	5917.46
5	10.0	26903	37.32	7.309	720.87	5917.02
6	10.0	26920	43.62	6.254	617.14	5920.75
7	10.0	26927	50.70	5.380	531.10	5922.29
8	10.0	26965	63.99	4.263	421.39	5930.65
9	10.0	27051	89.04	3.063	303.80	5949.57
10	10.0	26910	117.51	2.321	229.00	5918.56
11	10.0	26736	134.96	2.021	198.10	5880.29

weight-lbs; time-seconds; rate-litre/min; factor-pulses per litre;

Mean factor is 5919.079 pulses per litre

Calibrated by; P.A.FARRELL : Checked; : Date; 29.3.90

Notes;

 program in BASIC QRAP4/80

QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QEL/16/MP2 　　　　Serial number; 10836

Customer; BHRA 　　　　W.O. number; 10823

Order number; 48963/JK 　　　　Tag number; -

Calibration conditions:

Fluid; WATER 　　Pressure; 60 psig 　　Temperature; 16.4 celsius

s.g.; 0.9982 　　Viscosity; 1 cSt 　　Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	100.0	37918	37.21	4.398	1019.02	834002.49
2	100.0	37893	41.43	3.950	914.62	833452.62
3	100.0	37893	46.08	3.551	822.33	833452.62
4	100.0	37874	51.69	3.166	732.71	833034.71
5	100.0	37889	63.26	2.587	598.94	833364.64
6	100.0	37882	74.06	2.210	511.50	833210.67
7	100.0	37884	93.97	1.741	403.14	833254.66
8	100.0	37902	118.86	1.377	318.87	833650.57
9	100.0	37897	157.04	1.042	241.32	833540.60
10	100.0	37791	258.93	.632	145.95	831209.14

weight-lbs; time-seconds; rate-cu m/h; factor-pulses per cu metre;

Mean factor is 833217.275 pulses per cu metre

Calibrated by; P.A.FARRELL 　: Checked; 　　　: Date; 29.3.90

Notes;

 　　　　program in BASIC QRAP4/80

QUADRINA LTD

Q-FLO WATER CALIBRATION CERTIFICATE

Flowmeter type; QFL/38B/MP2 Serial number; 10834

Customer; BHRA W.O. number; 10823

Order number; 48963/JK Tag number; -

Calibration conditions:
Fluid; WATER Pressure; 60 psig Temperature; 16.7 celsius
s.g.; 0.9982 Viscosity; 1 cSt Pickup type; MAGNETIC

Results:

run	weight	counts	time	rate	o/p Hz	factor
1	300.0	11039	15.21	32.284	725.77	80930.24
2	300.0	11041	17.02	28.851	648.70	80944.90
3	300.0	11034	18.16	27.039	607.59	80893.58
4	300.0	11036	21.75	22.576	507.40	80908.25
5	300.0	11043	25.61	19.173	431.19	80959.57
6	300.0	11057	30.00	16.368	368.56	81062.20
7	300.0	11060	37.39	13.133	295.80	81084.20
8	300.0	11074	46.73	10.508	236.97	81186.84
9	300.0	11085	77.29	6.353	143.42	81267.48
10	300.0	11073	110.99	4.424	99.76	81179.51
11	300.0	11031	210.74	2.330	52.34	80871.59

weight-lbs; time-seconds; rate-cu m/h; factor-pulses per cu metre;

Mean factor is 81026.218 pulses per cu metre

Calibrated by; P.A.FARRELL : Checked; : Date; 29.3.90

Notes;

 program in BASIC QRAP4/80

CALIBRATION CERTIFICATE ①

TULIP Ch: ∅

Calibration date:	15.7.1987 WO: E02609
Transducer type:	PDCR 10/35L
Serial Number:	206277
Range:	3.5 bar d
Supply:	10V D.C.
Sensitivity:	98.98mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE ②

TULIP Ch: 1

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	229251
Range:	1 Bar d
Supply:	10V dc
Sensitivity:	99.36mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

③

CALIBRATION CERTIFICATE

TULIP Ch: 2

Calibration date:	13.9.1988 WO: F05437
Transducer type:	PDCR 10/35L
Serial Number:	229250
Range:	1 Bar d
Supply:	10V dc
Sensitivity:	99.40mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

CALIBRATION CERTIFICATE

④ TULIP Ch: 3

Calibration date:	6.3.90 WO: H01416
Transducer type:	PDCR 10/7L
Serial Number:	283991
Range:	700 mbar d
Supply:	10 Volts
Sensitivity:	98.90mV
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	0 to 50°C

CALIBRATION CERTIFICATE

TULIP Ch: 4 ⑤

Calibration date:	6.3.90 WO: H01416
Transducer type:	PDCR 10/7L
Serial Number:	283993
Range:	700 mbar d
Supply:	10 Volts
Sensitivity:	99.27mV
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	0 to 50°C
Temperature error band:	±0.5%

CALIBRATION CERTIFICATE ⑩

TULIP Ch: 5

Air TM (high press)

Calibration date:	15.7.1987 WO: E02609 ✓
Transducer type:	PDCR 10/35L ✓
Serial Number:	206275
Range:	3.5 bar d
Supply:	10V D.C.
Sensitivity:	100.52mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE ⑨ w/w (N2) ✓

TULIP Ch: 6

Calibration date:	24.3.1988 WO: E06257
Transducer type:	PDCR 120/35WL
Serial Number:	229647
Range:	2 bar d
Supply:	10 Volts
Sensitivity:	100.38mV @ 23°C
Non-linearity & Hysteresis:	±0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	±1.5%

CALIBRATION CERTIFICATE ⑩ Air TM (low press) ✓

unused

Calibration date:	15.12.1987 WO: E06257
Transducer type:	PDCR 10/35L
Serial Number:	212739
Range:	2 Bar d
Supply:	10V dc
Sensitivity:	101.41mV @ 23°C
Non-linearity & Hysteresis:	± 0.1% BSL
Temperature operating range:	
Temperature compensated range:	-20 to +80°C
Temperature error band:	± 1.5%

DRUCK Pressure Transducer Calibrations

Horizontal, Lube oil 2" Tests

= On Test Loop

= Off Test Loop

Danfoss

Danfoss Flowmetering Ltd

Magflo House, Ebley Road
Stonehouse, Glos. GL10 2LU
Tel: Stonehouse (045382) 8891
Telex: 43692 MAGFLO G
Telefax: (045382) 4013

John Knopp
B.H.R.A.
Cranfield
Beds

SB/SM/0011

April 11, 1990

Dear John

This is to confirm that, in February 1990, two Krohne Electromagnetic Flowmeters (serial nos. 874791A - 10mm and 874791B - 40mm), supplied by you, were checked on the 100kg Calibration Rig of Danfoss Flowmetering Ltd.

The 4-20mA output signal from the flowmeter, was measured across a 100 ohm standard resistance by an averaging DVM, to provide the average measured flowrate.

The true flowrate was obtained by measuring the mass of water collected in a tank for a known period.

The difference between the measured and true flowrates is presented (as a percentage of the true flowrate) on the results sheets included.

It was observed that the flowrates obtained from both the frequency and analogue outputs of 874791B were exactly 10% that of the flowmeter display, which agreed with the true flowrate. This is reflected in the error of 90% obtained during the tests.

All the equipment used in the tests is traceable to national standards through an external NAMAS Accredited Calibration Laboratory.

Yours sincerely

SMans.

pp S C BRANDON
Flow Calibration Leader

DANFOSS
FLOWMETERING LTD.

: CALIBRATION RESULTS :

CUSTOMER NAME : DBRI SERVICE
CUSTOMER ORDER No.:
SERIAL No. : 874791A TAG No. :
TYPE : KROHNE SIZE : 10 mm
CONVERTER OUTPUT : 4 - 20 mA
MAXIMUM FLOW : .0785 Litres/Sec WATER TEMP : 20 Deg.C

Test No.	% of Max. Flow	Diversion Time (secs)	Collected Weight (Kg)	Output Current (Ma)	True Flowrate	Measured Flowrate	% Error
1	86	611.2390	41.07	17.6838	0.0674	0.0671	-0.370
2	19	1937.3158	28.50	6.9708	0.0147	0.0146	-1.200
3	33	1223.0175	31.32	9.2264	0.0257	0.0256	-0.150
4	48	750.4845	28.21	11.6734	0.0377	0.0376	-0.130
5	78	531.4123	32.32	16.3901	0.0610	0.0608	-0.330

Reference Current L 5.149 mA

Calibrated by .... Witnessed by ....Date 09.02.90

! CALIBRATION RESULTS !

DANFOSS
FLOWMETERING LTD.

CUSTOMER NAME : DBRI SERVICE/BHRA
CUSTOMER ORDER No.:
SERIAL No. : 874791B TAG No. :
TYPE : KROHNE SIZE : 40 mm
CONVERTER OUTPUT : 4 - 20 mA
MAXIMUM FLOW : 1.2567 Litres/Sec WATER TEMP : 16 Deg.C

Test No.	% of Max. Flow	Diversion Time (secs)	Collected Weight (Kg)	Output Current (Ma)	True Flowrate	Measured Flowrate	% Error
1	99	64.9840	80.86	5.5915	1.2469	0.1250	-89.980
2	74	55.1518	50.85	5.1803	0.9239	0.0927	-89.970
3	50	80.4164	50.80	4.8065	0.6330	0.0633	-89.990
4	32	102.7837	41.48	4.5168	0.4044	0.0406	-89.960
5	13	157.1951	25.17	4.2075	0.1604	0.0163	-89.840

Reference L 5.285

Calibrated by .... Witnessed by ....Date 12.02.90

TABLES AND FIGURES

HYPOTHETICAL FIELD PROPERTIES	
WATER DEPTH	150m
RECOVERABLE RESERVES	60mn bbl oil
WELLS	7 Prod, 3 Injn
PIPELINE	8–inch i.d, 18km long
CRUDE PROPERTIES	40°API, GOR=800scf/bbl

TABLE 2–1

CASE	Prod Profile	£-$ (-)	CapEx (£mn)	IRR Required (%)	Oil Price ($/bbl)
1	Met	1.6	180	10	15.4
		1.6	150	10	12.8
		1.6	100	10	8.6
		1.6	180	20	19.3
		1.6	150	20	16.0
		1.6	100	20	11.0
		1.6	180	30	23.6
		1.6	150	30	19.7
		1.6	100	30	13.1
2	Down	1.6	180	10	16.4
		1.6	150	10	13.7
		1.6	100	10	9.1
		1.6	180	20	20.2
		1.6	150	20	17.0
		1.6	100	20	11.3
		1.6	180	30	24.6
		1.6	150	30	20.5
		1.6	100	30	13.6

TABLE 2-2

CASE	Prod Profile	£-$ (-)	CapEx (£mn)	IRR Required (%)	Oil Price ($/bbl)
3	Met	1.2	180	10	11.5
		1.2	150	10	9.6
		1.2	100	10	6.4
		1.2	180	20	14.5
		1.2	150	20	12.0
		1.2	100	20	8.0
		1.2	180	30	17.7
		1.2	150	30	14.7
		1.2	100	30	9.8
4	Down	1.2	180	10	12.4
		1.2	150	10	10.3
		1.2	100	10	6.8
		1.2	180	20	15.2
		1.2	150	20	12.7
		1.2	100	20	8.5
		1.2	180	30	18.4
		1.2	150	30	15.4
		1.2	100	30	10.2

TABLE 2-3

OIL-WATER FLOW REGIME BY GUZHOV ET AL (1973)	
CODE	**DESCRIPTION**
1	Stratified Flow
2	Stratified flow with dense layer of emulsion at the interface (lower layer = water)
3	Stratified flow with dense layer of emulsion at the interface (lower layer = dilute oil-in-water emulsion)
4	Emulsion of water-in-oil and oil-in-water
5	Emulsion of water-in-oil
6	Dense emulsion of oil-in-water and water
7	Dense emulsion of oil-in-water and dispersed emulsion of oil-in-water
8	Emulsion of oil-in-water

TABLE 3-1

COMPARISON OF OIL PHYSICAL PROPERTIES AT 20 C		
PROPERTY	OIL No1 (kerosene)	OIL No2 (lube oil)
Density (kg/m3)	793	812
Viscosity (cP)	1.75	3.90
Oil/air surface tension (mN/m)	28.5	26.0
Oil/water interfacial tension (mN/m)	32.5	40.5

TABLE 5-1

TEST CODE	V_{sL} (m/s)	V_{so} (m/s)	λ_w (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 50m
OWK01B	0.125	0.1	0.2	0.26	A
OWK01C	0.167	0.1	0.4	0.42	A
OWK01D	0.200	0.1	0.5	0.50	B
OWK01E	0.250	0.1	0.6	0.58	B
OWK01F	0.330	0.1	0.7	0.68	B
OWK02B	0.188	0.15	0.2	0.23	A
OWK02C	0.250	0.15	0.4	0.40	A
OWK02D	0.300	0.15	0.5	0.48	B
OWK02E	0.380	0.15	0.6	0.55	B
OWK02F	0.500	0.15	0.7	0.65	D
OWK03B	0.375	0.3	0.2	0.22	C
OWK03D	0.600	0.3	0.5	0.48	C
OWK03E	0.750	0.3	0.6	0.60	D
OWK03F	1.000	0.3	0.7	0.69	D
OWK04B	0.630	0.5	0.2	–	C
OWK04C	0.830	0.5	0.4	–	C
OWK04D	1.000	0.5	0.5	–	D
OWK04E	1.250	0.5	0.6	–	E
OWK04F	1.670	0.5	0.7	–	F
OWK05B	1.000	0.8	0.2	0.22	C
OWKO5C	1.330	0.8	0.4	0.41	E
OWK05D	1.600	0.8	0.5	0.51	E
OWK05E	2.000	0.8	0.6	0.61	F
OWK06B	1.250	1.0	0.2	0.22	E
OWK06C	1.670	1.0	0.4	0.42	F
OWK06D	2.000	1.0	0.5	0.51	F

TABLE 5–2

TEST CODE	V_{sL} (m/s)	λ_w (–)	dp/dx (mB/m)	$(WC)_a$ (–)	Flow Regime Obsvd at 50m
OWD01B	0.1	0.2	–	0.22	B
OWD01C	0.1	0.4	–	0.41	B
OWD01D	0.1	0.5	–	0.50	B
OWD01E	0.1	0.7	–	0.65	A
OWD02A	0.3	0	0.23	0	
OWD02B	0.3	0.2	0.37	0.20	C
OWD02C	0.3	0.4	0.23	0.40	C
OWD02D	0.3	0.5	0.24	0.48	C
OWD02E	0.3	0.7	0.29	0.67	C
OWD03A	0.5	0	0.60	0	
OWD03B	0.5	0.2	0.82	0.20	C
OWD03C	0.5	0.4	0.78	0.39	C
OWD03D	0.5	0.5	0.75	0.50	C
OWD03E	0.5	0.7	0.76	0.68	C
OWD04A	1.0	0	2.28	0	
OWD04B	1.0	0.2	2.36	0.19	E
OWD04C	1.0	0.4	2.54	0.39	D
OWD04D	1.0	0.5	2.63	0.49	D
OWD04E	1.0	0.7	2.44	0.68	D
OWD05A	2.0	0	7.94	0	
OWD05B	2.0	0.2	7.11	0.18	F
OWDO5C	2.0	0.4	6.75	–	F
OWD05D	2.0	0.5	6.76	–	F
OWD05E	2.0	0.7	8.23	–	F

TABLE 5–3

TEST CODE	V_{sL} (m/s)	λ_w (-)	Pressure Gradient Measured (mB/m)	Pressure Gradient Charles-Lilleheht (mB/m)
OWL0501	2.0	0.25	7.75	6.32
OWL0502	2.0	0.28	7.57	6.27
OWL0503	2.0	0.3	7.50	6.37
OWL0504	2.0	0.32	7.40	6.21
OWL0505	2.0	0.35	6.67	5.97
OWL0506	2.0	0.38	7.09	5.71
OWL0507	2.0	0.42	6.59	5.97
OWL0508	2.0	0.45	6.59	5.95
OWL0509	2.0	0.48	6.79	5.89
OWL0510	2.0	0.52	7.20	6.40
OWL0511	2.0	0.55	7.37	6.22
OWL0512	2.0	0.6	7.73	6.49
OWL0513	2.0	0.7	8.07	6.08

TABLE 5-4

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Average Holdup Meas (–)	Eaton (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Obsvd at 50m	Taitel–Dukler
3P01AA	0.1	0	1.12	0.17	0.24	0.31	0.28	0.26	0	B	B	STSM
3P01AB	0.1	0	3.39	0.51	0.53	0.62	0.24	0.11	0	B	B	STWV
3P01AC	0.1	0	5.63	0.51	0.86	0.91	0.17	0.06	0	B	B	STWV
3P01AD	0.1	0	8.66	0.86	1.46	1.34	0.11	0.03	0	B	B	STWV
3P01BA	0.1	0.2	1.16	0.17	0.35	0.43	0.26	0.28	0.27	A	A	STSM
3P01BB	0.1	0.2	3.35	0.56	0.71	0.81	0.25	0.13	0.20	A	A	STWV
3P01BC	0.1	0.2	5.34	0.64	1.07	1.11	0.21	0.08	0.19	B	B	STWV
3P01BD	0.1	0.2	8.39	1.37	1.78	1.63	0.14	0.04	0.21	B	B	STWV
3P01CA	0.1	0.4	1.20	0.51	0.51	0.58	0.25	0.30	0.44	C	C	STSM
3P01CB	0.1	0.4	3.49	0.86	1.04	1.11	0.19	0.14	0.42	C	C	STWV
3P01CC	0.1	0.4	5.46	1.20	1.56	1.53	0.21	0.09	0.48	H	H	STWV
3P01CD	0.1	0.4	8.28	1.37	2.36	2.13	0.17	0.05	0.47	H	K	ANDP
3P01DA	0.1	0.5	1.30	0.68	0.68	0.72	0.28	0.30	0.57	C	C	STWV
3P01DB	0.1	0.5	3.56	1.20	1.32	1.34	0.21	0.14	0.48	C	C	STWV
3P01DC	0.1	0.5	6.69	1.54	2.56	2.30	0.18	0.08	0.56	H	H	ANDP
3P01DD	0.1	0.5	10.50	2.39	4.24	3.35	0.13	0.04	0.54	K	K	ANDP
3P01EA	0.1	0.7	1.09	0.86	1.29	1.19	0.26	0.39	0.73	D	D	INTM
3P01EB	0.1	0.7	3.97	2.22	3.01	2.59	0.21	0.17	0.76	D	F	INTM
3P01EC	0.1	0.7	8.00	4.44	5.40	4.27	0.16	0.09	0.75	K	K	ANDP
3P01ED	0.1	0.7	9.35	4.62	5.78	4.49	0.13	0.07	0.77	K	K	ANDP

TABLE 6–1

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E–D (mB/m)	Pressure Gradient E–O (mB/m)	Average Holdup Meas (–)	Average Holdup Eaton (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Flow Regime Obsvd at 50m	Flow Regime Taitel–Dukler
3P02AA	0.15	0	1.30	0.17	0.44	0.51	0.25	0.28	0	E	E	STSM
3P02AB	0.15	0	3.87	0.68	0.94	1.02	0.18	0.12	0	K	K	STWV
3P02AC	0.15	0	6.33	1.20	1.48	1.46	0.15	0.07	0	K	K	STWV
3P02AD	0.15	0	8.76	1.37	2.12	1.91	0.13	0.05	0	K	K	ANDP
3P02BA	0.15	0.2	1.27	0.34	0.60	0.66	0.26	0.30	0.23	C	C	STSM
3P02BB	0.15	0.2	3.46	1.03	1.14	1.19	0.16	0.15	0.19	C	C	STWV
3P02BC	0.15	0.2	5.75	1.20	1.77	1.69	0.15	0.09	0.20	E	E	STWV
3P02BD	0.15	0.2	8.44	1.70	2.58	2.29	0.11	0.06	–	K	J	ANDP
3P02CA	0.15	0.4	1.04	0.68	0.80	0.81	0.28	0.37	0.46	C	C	STWV
3P02CB	0.15	0.4	3.07	1.03	1.58	1.50	0.19	0.18	0.38	E	E	STWV
3P02CC	0.15	0.4	5.68	1.71	2.50	2.28	0.16	0.11	0.38	E	E	ANDP
3P02CD	0.15	0.4	9.10	2.22	3.98	3.31	0.15	0.06	0.40	K	K	ANDP
3P02DA	0.15	0.5	1.08	0.86	1.04	0.99	0.30	0.38	0.54	D	D	STWV
3P02DB	0.15	0.5	3.47	1.54	2.19	1.96	0.23	0.18	0.52	F	F	INTM
3P02DC	0.15	0.5	5.97	2.05	3.30	2.86	0.16	0.11	0.56	F	F	ANDP
3P02DD	0.15	0.5	9.24	2.91	5.13	4.02	0.16	0.07	0.56	K	K	ANDP
3P02EA	0.15	0.7	1.37	2.05	2.42	1.95	0.36	0.39	0.75	D	D	INTM
3P02EB	0.15	0.7	3.36	3.59	4.17	3.32	0.33	0.23	0.75	D	F	INTM
3P02EC	0.15	0.7	6.19	5.64	4.86	4.86	0.30	0.15	0.80	F	F	INTM
3P02ED	0.15	0.7	9.66	8.89	9.36	6.53	0.11	0.10	0.82	K	K	ANDP

TABLE 6–2

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Average Holdup Meas (–)	Eaton (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Obsvd at 50m	Taitel–Dukler
3P03AA	0.30	0	0.87	0.86	0.87	0.87	0.48	0.44	0	F	F	INTM
3P03AB	0.30	0	3.81	1.71	2.34	2.12	0.23	0.17	0	F	F	INTM
3P03AC	0.30	0	6.17	2.39	3.35	2.96	0.19	0.12	0	K	K	ANDP
3P03AD	0.30	0	8.36	3.59	4.51	3.69	0.12	0.09	0	K	K	ANDP
3P03BA	0.30	0.2	1.08	0.51	1.44	1.31	0.32	0.42	0.19	D	D	INTM
3P03BB	0.30	0.2	3.29	1.37	2.84	2.51	0.19	0.22	0.16	D	F	INTM
3P03BC	0.30	0.2	5.95	2.74	4.41	3.72	0.20	0.14	0.15	K	K	ANDP
3P03BD	0.30	0.2	8.43	3.59	6.09	4.75	0.13	0.10	0.23	K	K	ANDP
3P03CA	0.30	0.4	1.07	1.54	2.18	1.82	0.43	0.45	0.44	D	D	INTM
3P03CB	0.30	0.4	3.22	1.88	3.46	4.26	0.30	0.24	0.38	D	F	INTM
3P03CC	0.30	0.4	5.81	3.08	6.44	4.94	0.19	0.16	0.37	D	K	INTM
3P03CD	0.30	0.4	9.65	5.47	10.09	7.11	0.12	0.10	0.35	K	K	ANDP
3P03DA	0.30	0.5	1.19	2.39	2.98	2.33	0.46	0.45	0.54	D	D	INTM
3P03DB	0.30	0.5	3.21	3.76	5.47	4.12	0.29	0.26	0.52	D	F	INTM
3P03DC	0.30	0.5	5.90	5.64	8.48	6.06	0.14	0.16	0.50	F	F	INTM
3P03DD	0.30	0.5	9.35	6.33	12.28	8.38	0.15	0.12	0.47	K	K	ANDP
3P03EA	0.30	0.7	1.21	4.44	5.98	4.05	0.49	0.51	0.81	D	F	INTM
3P03EB	0.30	0.7	3.35	7.86	11.20	7.05	0.35	0.30	0.71	F	F	INTM
3P03EC	0.30	0.7	6.51	13.70	17.33	10.50	0.25	0.20	0.72	K	K	INTM
3P03ED	0.30	0.7	10.34	20.50	24.00	14.00	0.21	0.14	0.71	K	K	INTM

TABLE 6–3

TEST CODE	V_{so}	λ_w	V_{sg} at 39m	Pressure Gradient			Average Holdup		$(WC)_a$	Flow Regime		
	(m/s)	(–)	(m/s)	Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Meas (–)	Eaton (–)	(–)	Obsvd at 25m	Obsvd at 50m	Taitel–Dukler
3P04AA	0.50	0	1.04	1.54	2.00	1.71	0.46	0.47	0	F	F	INTM
3P04AB	0.50	0	3.24	3.25	3.98	3.28	0.27	0.25	0	F	F	INTM
3P04AC	0.50	0	5.79	5.30	5.98	4.67	0.22	0.16	0	F	F	INTM
3P04AD	0.50	0	9.00	8.00	8.32	6.24	0.12	0.12	0	K	K	ANDP
3P04BA	0.50	0.2	1.02	1.54	2.83	2.25	0.44	0.50	0.21	D	D	INTM
3P04BB	0.50	0.2	3.00	3.25	5.36	4.06	0.39	0.28	0.18	D	F	INTM
3P04BC	0.50	0.2	6.08	5.30	8.72	6.24	0.21	0.17	0.17	K	K	INTM
3P04BD	0.50	0.2	8.97	7.18	11.53	8.13	0.17	0.13	0.15	K	K	INTM
3P04CA	0.50	0.4	1.15	2.05	4.49	3.25	0.48	0.51	0.42	D	F	INTM
3P04CB	0.50	0.4	3.35	3.76	8.69	5.87	0.34	0.28	0.37	D	F	INTM
3P04CC	0.50	0.4	5.83	6.15	12.66	8.30	0.25	0.20	0.37	D	F	INTM
3P04CD	0.50	0.4	8.88	9.23	17.12	10.89	0.18	0.15	0.37	K	K	INTM
3P04DA	0.50	0.5	1.11	4.10	5.63	3.90	0.55	0.54	0.55	D	F	INTM
3P04DB	0.50	0.5	3.58	8.21	11.46	7.28	0.32	0.30	0.53	F	F	INTM
3P04DC	0.50	0.5	6.64	12.30	17.47	10.66	0.20	0.20	0.46	F	F	INTM
3P04DD	0.50	0.5	10.67	14.20	25.66	14.84	0.22	0.14	0.50	K	K	INTM
3P04EA	0.50	0.7	1.20	10.10	12.02	7.55	0.66	0.59	0.73	D	F	INTM
3P04EB	0.50	0.7	2.55	13.85	18.48	10.23	0.44	0.42	0.75	D	F	INTM
3P04EC	0.50	0.7	3.79	17.57	23.47	12.52	0.45	0.35	0.77	D	F	INTM
3P04ED	0.50	0.7	6.82	26.79	34.69	17.52	0.28	0.25	0.76	F	K	INTM

TABLE 6–4

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E–D (mB/m)	Pressure Gradient E–O (mB/m)	Average Holdup Meas (–)	Average Holdup Eaton (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Flow Regime Obsvd at 50m	Flow Regime Taitel–Dukler
3P05AA	1.00	0	1.32	4.10	5.85	4.10	0.53	0.50	0	F	F	INTM
3P05AB	1.00	0	3.78	7.69	11.04	7.21	0.31	0.29	0	F	F	INTM
3P05AC	1.00	0	6.56	12.14	15.42	9.92	0.21	0.21	0	F	F	INTM
3P05AD	1.00	0	10.40	17.09	21.51	13.17	0.16	0.15	0	K	K	INTM
3P05BA	1.00	0.2	1.30	4.44	8.03	5.32	0.54	0.54	0.16	D	F	INTM
3P05BB	1.00	0.2	3.72	8.21	14.95	9.05	0.37	0.32	0.17	D	F	INTM
3P05BC	1.00	0.2	7.95	13.16	26.69	14.52	0.21	0.19	0.14	K	K	INTM
3P05BD	1.00	0.2	10.37	17.44	30.09	16.57	0.15	0.17	–	K	K	INTM
3P05CA	1.00	0.4	1.19	8.03	11.47	7.36	0.62	0.60	0.39	D	F	INTM
3P05CB	1.00	0.4	2.59	9.74	18.17	10.20	0.46	0.42	0.39	D	F	INTM
3P05CC	1.00	0.4	3.99	12.31	23.71	12.77	0.34	0.33	0.38	D	F	INTM
3P05CD	1.00	0.4	7.20	18.97	35.25	17.99	0.23	0.24	0.39	K	K	INTM
3P05DA	1.00	0.5	1.27	13.16	15.11	9.72	0.64	0.62	0.57	D	F	INTM
3P05DB	1.00	0.5	2.65	17.61	23.33	12.55	0.52	0.45	0.56	F	F	INTM
3P05DC	1.00	0.5	4.15	23.08	30.62	15.56	0.47	0.36	0.57	F	F	INTM
3P05DD	1.00	0.5	7.48	34.70	44.72	21.48	0.32	0.26	0.54	K	K	INTM

TABLE 6–5

TEST CODE	V_{sL} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Slug Length $L_{S\,min}$ (m)	Slug Length $L_{S\,max}$ (m)	Slug Length $L_{S\,av}$ (m)	Slug Front Velocity V_S (m/s)	Slug Front Velocity V_S / V_m (–)	Slug Frequency Obsvd (1/s)	Slug Frequency Greg–Scott (1/s)
SLUG H1	0.3	0	1.10	0.39	3.00	1.35	1.95	1.40	0.171	0.336
SLUG H2	0.3	0.5	1.10	0.39	3.53	1.64	2.03	1.45	0.134	0.336
SLUG H3	1.0	0	1.26	0.54	2.44	1.24	3.12	1.38	0.794	0.950
SLUG H4	1.0	0.5	1.24	0.54	2.60	1.32	3.33	1.49	0.629	0.950
SLUG H5	1.67	0.4	1.27	0.47	1.88	0.95	3.70	1.25	1.410	1.490
SLUG H6	1.67	0.7	1.25	0.47	1.52	0.89	3.55	1.21	1.410	1.490

TABLE 6–6

TEST CODE	V_{sL} (m/s)	λ_w (-)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E-D (mB/m)	Pressure Gradient E-O (mB/m)	Average Holdup Meas (-)	Average Holdup Eaton (-)	$(WC)_a$ (-)	Flow Regime Obsvd at 25m	Flow Regime Obsvd at 50m	Flow Regime Taitel-Dukler
3D01AA	0.1	0	1.10	0.21	0.34	0.46	0.35	0.29	0	B	E	STSM
3D01AB	0.1	0	3.13	0.42	0.72	0.83	0.24	0.14	0	B	J	STSM
3D01AC	0.1	0	5.87	0.78	1.31	1.28	0.18	0.07	0	J	J	STWV
3D01AD	0.1	0	8.14	1.11	1.91	1.67	0.13	0.04	0	M	J	ANDP
3D01BA	0.1	0.2	1.08	0.23	0.31	0.42	0.34	0.28	0.23	C	C	STSM
3D01BB	0.1	0.2	3.02	0.45	0.66	0.76	0.25	0.14	0.24	C	C	STSM
3D01BC	0.1	0.2	5.70	1.23	1.23	1.22	0.18	0.07	0.22	G	G	STWV
3D01BD	0.1	0.2	7.92	1.06	1.76	1.57	0.13	0.04	0.21	A	A	ANDP
3D01CA	0.1	0.4	1.08	0.13	0.31	0.42	0.34	0.28	0.43	C	C	STSM
3D01CB	0.1	0.4	3.25	0.36	0.69	0.79	0.26	0.12	0.48	G	G	STWV
3D01CC	0.1	0.4	5.87	0.58	1.23	1.22	0.19	0.06	0.43	A	A	STWV
3D01CD	0.1	0.4	8.04	0.92	1.77	1.59	0.14	0.04	0.43	A	A	STWV
3D01DA	0.1	0.5	1.15	0.29	0.31	0.41	0.33	0.27	0.51	C	C	STSM
3D01DB	0.1	0.5	3.11	0.52	0.64	0.75	0.27	0.13	0.55	C	A	STWV
3D01DC	0.1	0.5	5.93	0.71	1.23	1.23	0.19	0.06	0.53	A	A	STWV
3D01DD	0.1	0.5	8.02	1.07	1.74	1.57	0.15	0.04	0.52	A	A	STWV
3D01EA	0.1	0.7	1.07	0.30	0.30	0.39	0.34	0.28	0.71	C	C	STSM
3D01EB	0.1	0.7	3.25	0.42	0.67	0.76	0.25	0.12	0.73	G	G	STWV
3D01EC	0.1	0.7	6.15	0.81	1.25	1.25	0.16	0.06	0.72	A	A	STWV
3D01ED	0.1	0.7	8.35	1.26	1.81	1.62	0.14	0.04	0.73	A	A	STWV

TABLE 6-7

TEST CODE	V_{sL} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E–D (mB/m)	Pressure Gradient E–O (mB/m)	Average Holdup Meas (–)	Average Holdup Eaton (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Flow Regime Obsvd at 50m	Flow Regime Taitel–Dukler
3D02AA	0.30	0	1.06	0.72	1.39	1.45	0.36	0.42	0	F	F	INTM
3D02AB	0.30	0	3.28	1.75	2.72	2.65	0.31	0.22	0	F	F	INTM
3D02AC	0.30	0	6.15	3.19	4.22	3.84	0.24	0.13	0	K	K	ANDP
3D02AD	0.30	0	9.00	4.63	6.12	5.01	0.15	0.09	0	K	K	ANDP
3D02BA	0.30	0.2	1.08	0.61	1.40	1.44	0.38	0.42	0.21	D	D	INTM
3D02BB	0.30	0.2	3.43	1.69	2.82	2.70	0.31	0.21	0.19	D	D	INTM
3D02BC	0.30	0.2	5.69	2.71	4.10	3.73	0.24	0.14	0.16	H	K	ANDP
3D02BD	0.30	0.2	8.86	3.56	5.91	4.86	0.17	0.06	–	K	K	ANDP
3D02CA	0.30	0.4	1.13	0.67	1.38	1.40	0.41	0.40	0.41	D	D	INTM
3D02CB	0.30	0.4	3.44	1.42	2.76	2.60	0.30	0.20	0.39	D	D	INTM
3D02CC	0.30	0.4	5.95	2.07	4.04	3.65	0.23	0.13	0.40	H	H	ANDP
3D02CD	0.30	0.4	8,83	2.88	5.88	4.81	0.15	0.09	0.40	K	K	ANDP
3D02DA	0.30	0.5	1.11	0.99	1.38	1.39	0.40	0.40	0.50	D	D	INTM
3D02DB	0.30	0.5	3.39	1.35	2.69	2.54	0.33	0.20	0.51	D	F	INTM
3D02DC	0.30	0.5	5.83	2.10	4.00	3.62	0.22	0.13	0.51	K	K	ANDP
3D02DD	0.30	0.5	8.74	3.03	5.80	4.75	0.18	0.09	0.51	K	K	ANDP
3D02EA	0.30	0.7	1.02	0.84	1.26	1.25	0.42	0.41	0.71	D	D	INTM
3D02EB	0.30	0.7	3.23	1.78	2.56	2.40	0.29	0.20	0.77	D	F	INTM
3D02EC	0.30	0.7	5.79	2.83	3.90	3.50	0.24	0.13	0.77	K	K	ANDP
3D02ED	0.30	0.7	8.80	4.32	5.76	4.68	0.16	0.08	0.76	K	K	ANDP

TABLE 6–8

TEST CODE	V_{sL} (m/s)	λ_w (-)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E-D (mB/m)	Pressure Gradient E-O (mB/m)	Average Holdup Meas (-)	Average Holdup Eaton (-)	$(WC)_a$ (-)	Flow Regime Obsvd at 25m	Flow Regime Obsvd at 50m	Flow Regime Taitel-Dukler
3D03AA	0.5	0	1.08	1.63	2.66	2.44	0.49	0.48	0	F	F	INTM
3D03AB	0.5	0	3.29	3.52	5.24	4.50	0.34	0.26	0	F	F	INTM
3D03AC	0.5	0	5.89	5.65	7.70	6.20	0.24	0.17	0	K	F	INTM
3D03AD	0.5	0	8.95	8.16	10.47	8.17	0.18	0.12	0	K	K	ANDP
3D03BA	0.5	0.2	1.13	1.51	2.74	2.46	0.44	0.47	0.20	D	F	INTM
3D03BB	0.5	0.2	3.27	2.96	5.15	4.38	0.33	0.26	0.17	D	F	INTM
3D03BC	0.5	0.2	5.95	4.79	7.76	6.17	0.23	0.17	0.18	K	F	INTM
3D03BD	0.5	0.2	8.79	6.95	10.24	7.95	0.16	0.13	0.14	K	K	ANDP
3D03CA	0.5	0.4	1.11	1.30	2.64	2.34	0.49	0.47	0.41	D	D	INTM
3D03CB	0.5	0.4	3.20	2.60	4.95	4.16	0.33	0.26	0.39	D	D	INTM
3D03CC	0.5	0.4	5.89	4.53	7.56	5.98	0.27	0.17	0.40	K	F	INTM
3D03CD	0.5	0.4	8.94	6.38	10.30	7.87	0.19	0.12	0.39	K	K	ANDP
3D03DA	0.5	0.5	1.04	1.87	2.55	2.24	0.46	0.48	0.50	D	D	INTM
3D03DB	0.5	0.5	3.31	3.14	5.08	4.24	0.39	0.25	0.51	D	D	INTM
3D03DC	0.5	0.5	5.85	4.99	7.53	5.93	0.28	0.17	0.50	K	K	INTM
3D03DD	0.5	0.5	9.05	7.16	10.36	7.83	0.21	0.12	0.52	K	K	INTM
3D03EA	0.5	0.7	1.08	1.89	2.52	2.16	0.41	0.46	0.74	D	D	INTM
3D03EB	0.5	0.7	3.34	3.69	4.96	4.10	0.36	0.24	0.74	D	D	INTM
3D03EC	0.5	0.7	6.16	6.10	7.62	5.95	0.26	0.16	0.75	K	K	INTM
3D03ED	0.5	0.7	8.95	9.46	10.33	7.68	0.20	0.12	0.74	K	K	ANDP

TABLE 6-9

TEST CODE	V_{sL} (m/s)	λ_w (-)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E-D (mB/m)	E-O (mB/m)	Average Holdup Meas (-)	Eaton (-)	$(WC)_a$ (-)	Flow Regime Obsvd at 25m	Obsvd at 50m	Taitel-Dukler
3D04AA	1.0	0	1.11	4.20	6.78	5.09	0.61	0.56	0	F	F	INTM
3D04AB	1.0	0	3.47	7.98	13.13	9.02	0.38	0.32	0	F	F	INTM
3D04AC	1.0	0	6.45	13.78	19.68	13.26	0.24	0.23	0	K	K	INTM
3D04AD	1.0	0	9.56	18.10	25.44	16.38	0.20	0.17	0	K	K	INTM
3D04BA	1.0	0.2	1.08	3.93	6.73	5.00	0.62	0.57	0.19	D	F	INTM
3D04BB	1.0	0.2	3.48	7.02	13.10	8.90	0.36	0.32	0.20	F	F	INTM
3D04BC	1.0	0.2	6.51	12.11	19.33	12.79	0.29	0.23	0.18	F	F	INTM
3D04BD	1.0	0.2	9.37	15.51	25.10	15.66	0.18	0.17	0.15	K	K	INTM
3D04CA	1.0	0.4	1.15	3.79	6.70	4.88	0.64	0.55	0.41	D	F	INTM
3D04CB	1.0	0.4	3.49	6.24	12.82	8.59	0.38	0.32	0.40	F	F	INTM
3D04CC	1.0	0.4	6.34	10.86	18.57	12.18	0.28	0.23	0.38	F	F	INTM
3D04CD	1.0	0.4	9.19	14.17	24.40	15.00	0.24	0.17	0.38	K	F	INTM
3D04DA	1.0	0.5	1.10	4.29	6.48	4.75	0.61	0.56	0.53	D	F	INTM
3D04DB	1.0	0.5	3.50	8.08	12.86	8.55	0.36	0.31	0.51	D	F	INTM
3D04DC	1.0	0.5	6.52	14.21	19.07	12.27	0.30	0.21	0.50	F	F	INTM
3D04DD	1.0	0.5	9.67	18.88	25.00	15.37	0.21	0.16	0.50	K	K	INTM
3D04EA	1.0	0.7	1.19	4.70	6.70	4.73	0.58	0.53	0.72	D	F	INTM
3D04EB	1.0	0.7	3.52	8.79	12.86	8.39	0.37	0.30	0.74	D	F	INTM
3D04EC	1.0	0.7	6.69	14.92	19.96	12.32	0.29	0.21	0.72	F	F	INTM
3D04ED	1.0	0.7	9.63	21.17	25.24	15.24	0.23	0.16	0.74	K	K	INTM

TABLE 6-10

TEST CODE	V_{sL} (m/s)	λ_w (-)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E-D (mB/m)	E-O (mB/m)	Average Holdup Meas (-)	Eaton (-)	$(WC)_a$ (-)	Flow Regime Obsvd at 25m	Obsvd at 50m	Taitel-Dukler
3D05AA	2.0	0	1.33	12.90	19.31	12.64	0.70	0.63	0	F	F	INTM
3D05AB	2.0	0	2.71	17.92	28.47	16.23	0.53	0.46	0	F	F	INTM
3D05AC	2.0	0	4.27	23.67	37.28	20.10	0.42	0.36	0	F	F	INTM
3D05AD	2.0	0	6.69	31.35	50.00	25.50	0.31	0.29	0	F	F	INTM
3D05BA	2.0	0.2	1.28	11.00	18.51	12.10	0.69	0.63	0.20	F	F	INTM
3D05BB	2.0	0.2	2.65	15.35	27.82	15.71	0.54	0.47	0.19	F	F	INTM
3D05BC	2.0	0.2	4.15	20.65	36.51	19.49	0.41	0.37	0.19	F	F	INTM
3D05BD	2.0	0.2	6.57	28.50	49.21	24.85	0.34	0.29	0.18	F	F	INTM
3D05CA	2.0	0.4	1.27	13.13	18.16	11.83	0.70	0.63	0.41	F	F	INTM
3D05CB	2.0	0.4	2.71	17.81	27.70	15.44	0.55	0.46	0.40	F	F	INTM
3D05CC	2.0	0.4	4.33	22.19	37.20	19.50	0.41	0.36	0.40	F	F	INTM
3D05CD	2.0	0.4	6.77	28.26	49.70	24.70	0.35	0.28	0.41	F	F	INTM
3D05DA	2.0	0.5	1.29	13.65	18.10	11.60	0.69	0.63	0.53	D	F	INTM
3D05DB	2.0	0.5	2.73	19.35	27.90	15.36	0.53	0.45	0.53	F	F	INTM
3D05DC	2.0	0.5	4.34	25.44	36.92	19.22	0.41	0.36	0.50	F	F	INTM
3D05DD	2.0	0.5	6.41	33.68	47.70	23.80	0.36	0.29	0.52	F	F	INTM
3D05EA	2.0	0.7	1.27	14.28	17.57	11.17	0.68	0.62	0.73	D	F	INTM
3D05EB	2.0	0.7	2.68	20.08	27.08	14.74	0.53	0.45	0.73	F	F	INTM
3D05EC	2.0	0.7	4.31	26.73	36.39	18.62	0.45	0.35	0.76	F	F	INTM
3D05ED	2.0	0.7	6.51	36.09	47.70	23.40	0.33	0.28	0.73	F	F	INTM

TABLE 6-11

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 14m (m/s)	Pressure Gradient			Average Holdup			$(WC)_a$ (–)	Flow Regime	
				Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Meas (–)	Eaton (–)	Muk–Br (–)		Obsvd at 25m	Taitel–Dukler
3P11AA	0.1	0	1.22	0.56	0.94	1.00	0.30	0.26	0.31	0	F	INTM
3P11AB	0.1	0	3.56	1.05	0.83	0.90	0.24	0.11	0.15	0	F	INTM
3P11AC	0.1	0	5.78	0.91	1.05	1.06	0.20	0.07	0.09	0	K	INTM
3P11AD	0.1	0	8.88	1.08	1.53	1.37	0.14	0.04	0.05	0	K	INTM
3P11BA	0.1	0.2	1.15	0.61	1.08	1.15	0.35	0.29	0.35	0.24	D	INTM
3P11BB	0.1	0.2	3.16	1.14	1.06	0.98	0.33	0.14	0.19	0.26	D	INTM
3P11BC	0.1	0.2	5.55	1.66	1.23	1.25	0.24	0.08	0.11	0.32	H	INTM
3P11BD	0.1	0.2	8.00	1.56	1.62	1.52	0.16	0.05	0.07	0.30	H	ANDP
3P11CA	0.1	0.4	1.02	1.44	1.27	1.34	0.36	0.34	0.40	0.42	D	INTM
3P11CB	0.1	0.4	3.29	1.92	1.25	1.32	0.26	0.15	0.21	0.46	D	INTM
3P11CC	0.1	0.4	5.47	2.51	1.58	1.57	0.23	0.09	0.13	0.49	H	INTM
3P11CD	0.1	0.4	8.04	1.95	2.09	1.94	0.19	0.06	0.09	0.52	H	INTM
3P11DA	0.1	0.5	1.09	1.72	1.39	1.45	0.35	0.35	0.41	0.53	D	INTM
3P11DB	0.1	0.5	3.02	1.88	1.44	1.47	0.27	0.17	0.24	0.54	D	INTM
3P11DC	0.1	0.5	5.64	2.31	1.92	1.87	0.21	0.10	0.14	0.55	H	INTM
3P11DD	0.1	0.5	8.18	1.94	2.56	2.33	0.18	0.07	0.10	0.59	H	INTM
3P11EA	0.1	0.7	1.09	1.72	1.91	1.88	0.43	0.40	0.46	0.73	D	INTM
3P11EB	0.1	0.7	3.35	2.14	2.44	2.28	0.29	0.20	0.27	0.75	D	INTM
3P11EC	0.1	0.7	5.58	2.70	3.21	2.91	0.24	0.13	0.18	0.74	K	INTM
3P11ED	0.1	0.7	7.82	3.37	4.24	3.57	0.22	0.09	0.14	0.75	K	INTM

TABLE 7–1

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 14m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E–D (mB/m)	Pressure Gradient E–O (mB/m)	Average Holdup Meas (–)	Average Holdup Eaton (–)	Average Holdup Muk–Br (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Flow Regime Taitel–Dukler
3P12AA	0.15	0	1.11	1.09	1.11	1.18	0.31	0.32	0.38	0	F	INTM
3P12AB	0.15	0	3.26	1.04	1.08	1.16	0.24	0.15	0.20	0	F	INTM
3P12AC	0.15	0	5.41	1.21	1.35	1.35	0.21	0.09	0.13	0	K	INTM
3P12AD	0.15	0	8.03	1.37	1.84	1.71	0.16	0.06	0.08	0	K	INTM
3P12BA	0.15	0.2	1.10	1.02	1.28	1.35	0.36	0.34	0.40	0.24	D	INTM
3P12BB	0.15	0.2	3.31	1.37	1.34	1.39	0.23	0.16	0.21	0.27	D	INTM
3P12BC	0.15	0.2	5.67	1.87	1.71	1.68	0.21	0.10	0.14	0.29	K	INTM
3P12BD	0.15	0.2	8.35	2.15	2.37	2.16	0.16	0.06	0.09	0.30	K	ANDP
3P12CA	0.15	0.4	1.09	1.52	1.55	1.59	0.38	0.38	0.43	0.44	D	INTM
3P12CB	0.15	0.4	3.34	2.19	1.82	1.79	0.27	0.18	0.24	0.43	D	INTM
3P12CC	0.15	0.4	5.44	2.99	2.24	2.16	0.20	0.12	0.16	0.46	K	INTM
3P12CD	0.15	0.4	8.10	3.09	3.16	2.78	0.18	0.08	0.11	0.43	K	INTM
3P12DA	0.15	0.5	1.05	1.82	1.73	1.73	0.41	0.40	0.46	0.56	D	INTM
3P12DB	0.15	0.5	3.23	2.93	2.13	2.05	0.29	0.20	0.26	0.54	D	INTM
3P12DC	0.15	0.5	5.61	3.39	2.84	2.63	0.23	0.13	0.18	0.54	K	INTM
3P12DD	0.15	0.5	8.04	3.50	3.82	3.28	0.17	0.09	0.13	0.55	K	INTM
3P12EA	0.15	0.7	1.03	1.90	2.57	2.36	0.47	0.46	0.51	0.75	D	INTM
3P12EB	0.15	0.7	3.19	2.34	3.70	3.20	0.28	0.24	0.32	0.78	F	INTM
3P12EC	0.15	0.7	5.15	3.55	4.92	4.00	0.22	0.17	0.23	0.74	K	INTM
3P12ED	0.15	0.7	7.90	5.34	6.87	5.41	0.19	0.12	0.17	0.73	K	INTM

TABLE 7–2

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 14m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E–D (mB/m)	Pressure Gradient E–O (mB/m)	Average Holdup Meas (–)	Average Holdup Eaton (–)	Average Holdup Muk–Br (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Flow Regime Taitel–Dukler
3P13AA	0.30	0	1.09	1.42	1.64	1.62	0.37	0.40	0.45	0	F	INTM
3P13AB	0.30	0	3.12	1.87	1.97	1.94	0.27	0.21	0.27	0	F	INTM
3P13AC	0.30	0	5.54	2.55	2.68	2.50	0.21	0.14	0.18	0	F	INTM
3P13AD	0.30	0	7.92	3.29	3.59	3.09	0.15	0.09	0.13	0	K	INTM
3P13BA	0.30	0.2	1.02	1.90	1.94	1.90	0.40	0.44	0.48	0.23	D	INTM
3P13BB	0.30	0.2	3.30	2.43	2.60	2.43	0.32	0.22	0.28	0.23	D	INTM
3P13BC	0.30	0.2	5.56	3.33	3.52	3.13	0.25	0.15	0.20	0.27	D	INTM
3P13BD	0.30	0.2	8.22	4.42	4.78	3.96	0.15	0.10	0.14	0.27	K	INTM
3P13CA	0.30	0.4	1.05	2.82	2.51	2.31	0.47	0.47	0.51	0.41	D	INTM
3P13CB	0.30	0.4	3.32	3.62	3.76	3.25	0.32	0.25	0.31	0.44	D	INTM
3P13CC	0.30	0.4	5.59	4.93	5.08	4.15	0.27	0.16	0.22	0.46	K	INTM
3P13CD	0.30	0.4	8.20	6.56	6.75	5.33	0.16	0.12	0.16	0.49	K	INTM
3P13DA	0.30	0.5	1.00	3.14	2.93	2.60	0.54	0.50	0.53	0.55	D	INTM
3P13DB	0.30	0.5	3.16	4.10	4.59	3.74	0.37	0.27	0.34	0.54	F	INTM
3P13DC	0.30	0.5	5.52	6.99	6.42	4.95	0.28	0.18	0.24	0.54	K	INTM
3P13DD	0.30	0.5	8.23	9.30	8.40	6.39	0.21	0.13	0.18	0.58	K	INTM
3P13EA	0.30	0.7	1.03	3.75	5.22	4.06	0.58	0.55	0.57	0.74	D	INTM
3P13EB	0.30	0.7	3.33	5.63	9.11	6.32	0.34	0.30	0.37	0.73	F	INTM
3P13EC	0.30	0.7	5.60	8.34	12.44	8.51	0.30	0.22	0.29	0.73	F	INTM
3P13ED	0.30	0.7	8.27	12.20	16.86	10.94	0.20	0.16	0.22	0.71	K	INTM

TABLE 7–3

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 14m (m/s)	Pressure Gradient Meas (mB/m)	Pressure Gradient E–D (mB/m)	Pressure Gradient E–O (mB/m)	Average Holdup Meas (–)	Average Holdup Eaton (–)	Average Holdup Muk–Br (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Flow Regime Taitel–Dukler
3P14AA	0.50	0	1.04	1.78	2.40	2.23	0.46	0.48	0.51	0	F	INTM
3P14AB	0.50	0	2.95	2.57	3.31	2.91	0.29	0.26	0.32	0	F	INTM
3P14AC	0.50	0	5.46	3.76	4.95	4.05	0.17	0.17	0.22	0	F	INTM
3P14AD	0.50	0	8.34	5.56	6.62	5.27	0.13	0.12	0.16	0	K	INTM
3P14BA	0.50	0.2	1.00	2.79	3.02	2.65	0.47	0.51	0.54	0.21	D	INTM
3P14BB	0.50	0.2	3.14	3.35	4.69	3.82	0.32	0.27	0.34	0.24	D	INTM
3P14BC	0.50	0.2	5.66	5.06	6.71	5.16	0.23	0.18	0.24	0.24	F	INTM
3P14BD	0.50	0.2	8.29	6.89	8.71	6.61	0.17	0.13	0.18	0.29	K	INTM
3P14CA	0.50	0.4	1.06	4.24	4.21	3.42	0.56	0.52	0.55	0.44	D	INTM
3P14CB	0.50	0.4	3.07	5.58	6.62	4.93	0.36	0.30	0.37	0.42	F	INTM
3P14CC	0.50	0.4	5.43	7.39	9.48	6.87	0.24	0.21	0.27	0.40	F	INTM
3P14CD	0.50	0.4	7.94	10.38	12.26	8.58	0.21	0.15	0.21	0.44	K	INTM
3P14DA	0.50	0.5	1.07	4.79	5.16	4.02	0.62	0.55	0.57	0.53	D	INTM
3P14DB	0.50	0.5	3.00	6.81	8.43	5.95	0.40	0.33	0.40	0.54	F	INTM
3P14DC	0.50	0.5	5.63	10.78	12.24	8.45	0.31	0.22	0.28	0.51	F	INTM
3P14DD	0.50	0.5	7.65	13.49	15.45	10.09	0.25	0.16	0.23	0.51	K	INTM
3P14EA	0.50	0.7	1.08	6.14	9.53	6.74	0.65	0.61	0.61	0.73	D	INTM
3P14EB	0.50	0.7	2.16	7.62	13.54	8.51	0.47	0.44	0.50	0.73	D	INTM
3P14EC	0.50	0.7	3.22	9.50	17.00	10.28	0.43	0.36	0.43	0.72	F	INTM
3P14ED	0.50	0.7	5.28	13.23	23.28	13.58	0.32	0.26	0.30	0.73	K	INTM

TABLE 7–4

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 14m (m/s)	Pressure Gradient Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Average Holdup Meas (–)	Eaton (–)	Muk–Br (–)	$(WC)_a$ (–)	Flow Regime Obsvd at 25m	Taitel–Dukler
3P15AA	1.00	0	1.05	3.57	4.91	3.89	0.60	0.56	0.58	0	F	INTM
3P15AB	1.00	0	3.16	5.32	8.12	5.83	0.38	0.32	0.39	0	F	INTM
3P15AC	1.00	0	6.20	8.75	12.39	8.58	0.23	0.21	0.27	0	F	INTM
3P15AD	1.00	0	8.31	10.79	15.58	10.33	0.15	0.16	0.22	0	K	INTM
3P15BA	1.00	0.2	1.08	4.71	6.62	4.96	0.62	0.58	0.59	0.20	D	INTM
3P15BB	1.00	0.2	3.12	6.81	10.93	7.33	0.37	0.35	0.41	0.19	F	INTM
3P15BC	1.00	0.2	5.83	10.24	16.18	10.53	0.28	0.24	0.30	0.21	F	INTM
3P15BD	1.00	0.2	8.04	13.65	20.60	12.81	0.23	0.18	0.24	0.21	K	INTM
3P15CA	1.00	0.4	1.09	7.69	9.58	6.82	0.66	0.62	0.61	0.42	D	INTM
3P15CB	1.00	0.4	2.14	10.42	13.27	8.42	0.50	0.46	0.51	0.42	F	INTM
3P15CC	1.00	0.4	3.26	11.90	16.89	10.28	0.40	0.37	0.44	0.46	F	INTM
3P15CD	1.00	0.4	5.77	16.88	24.60	14.32	0.32	0.26	0.34	0.42	K	INTM
3P15DA	1.00	0.5	1.09	9.87	12.10	8.46	0.70	0.64	0.63	0.52	D	INTM
3P15DB	1.00	0.5	2.14	11.50	16.88	10.29	0.57	0.48	0.52	0.53	F	INTM
3P15DC	1.00	0.5	3.19	14.27	21.36	12.38	0.44	0.38	0.45	0.53	F	INTM
3P15DD	1.00	0.5	4.85	19.43	27.73	15.58	0.38	0.30	0.38	0.52	K	INTM

TABLE 7–5

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Flow Regime Obsvd at 50m	Flow Regime Taitel–Dukler
3P11AA	0.1	0	1.24	B	STSM
3P11AB	0.1	0	3.60	B	STWV
3P11AC	0.1	0	5.89	B	STWV
3P11AD	0.1	0	9.10	B	STWV
3P11BA	0.1	0.2	1.18	A	STSM
3P11BB	0.1	0.2	3.21	A	STWV
3P11BC	0.1	0.2	5.67	A	STWV
3P11BD	0.1	0.2	8.29	B	STWV
3P11CA	0.1	0.4	1.03	A	STSM
3P11CB	0.1	0.4	3.36	A	STWV
3P11CC	0.1	0.4	5.61	B	STWV
3P11CD	0.1	0.4	8.28	B	ANDP
3P11DA	0.1	0.5	1.10	A	STSM
3P11DB	0.1	0.5	3.09	A	STWV
3P11DC	0.1	0.5	5.83	B	STWV
3P11DD	0.1	0.5	8.47	B	ANDP
3P11EA	0.1	0.7	1.11	A	STWV
3P11EB	0.1	0.7	3.49	B	STWV
3P11EC	0.1	0.7	5.88	B	ANDP
3P11ED	0.1	0.7	8.46	B	ANDP

TABLE 7–6

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Flow Regime Obsvd at 50m	Flow Regime Taitel–Dukler
3P12AA	0.15	0	1.13	B	STSM
3P12AB	0.15	0	3.32	B	STWV
3P12AC	0.15	0	5.53	B	STWV
3P12AD	0.15	0	8.30	B	ANDP
3P12BA	0.15	0.2	1.12	A	STSM
3P12BB	0.15	0.2	3.39	A	STWV
3P12BC	0.15	0.2	5.84	B	STWV
3P12BD	0.15	0.2	8.69	B	ANDP
3P12CA	0.15	0.4	1.12	A	STSM
3P12CB	0.15	0.4	3.43	G	STWV
3P12CC	0.15	0.4	5.66	J	STWV
3P12CD	0.15	0.4	8.53	J	ANDP
3P12DA	0.15	0.5	1.08	A	STSM
3P12DB	0.15	0.5	3.34	G	STWV
3P12DC	0.15	0.5	5.89	J	ANDP
3P12DD	0.15	0.5	8.55	J	ANDP
3P12EA	0.15	0.7	1.06	D	STWV
3P12EB	0.15	0.7	3.38	F	ANDP
3P12EC	0.15	0.7	5.61	K	ANDP
3P12ED	0.15	0.7	8.98	K	ANDP

TABLE 7–7

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Flow Regime Obsvd at 50m	Taitel–Dukler
3P13AA	0.30	0	1.11	0.13	0.46	0.43	E	STSM
3P13AB	0.30	0	3.22	1.30	1.97	1.94	E	STWV
3P13AC	0.30	0	5.84	2.67	3.25	2.87	J	ANDP
3P13AD	0.30	0	8.50	3.88	4.79	3.86	K	ANDP
3P13BA	0.30	0.2	1.05	0.15	0.74	0.63	D	STSM
3P13BB	0.30	0.2	3.44	1.91	2.76	2.38	F	STWV
3P13BC	0.30	0.2	5.90	2.85	4.46	3.74	K	ANDP
3P13BD	0.30	0.2	8.92	4.22	6.69	5.12	K	ANDP
3P13CA	0.30	0.4	1.09	0.27	1.47	1.13	D	STWV
3P13CB	0.30	0.4	3.52	2.23	4.39	3.51	F	INTM
3P13CC	0.30	0.4	6.07	3.93	6.79	5.23	K	ANDP
3P13CD	0.30	0.4	9.14	5.41	10.07	7.17	K	ANDP
3P13DA	0.30	0.5	1.04	0.70	1.98	1.43	D	STWV
3P13DB	0.30	0.5	3.41	3.61	5.56	4.18	F	INTM
3P13DC	0.30	0.5	6.12	4.57	8.89	6.44	K	ANDP
3P13DD	0.30	0.5	9.39	6.25	13.07	8.87	K	ANDP
3P13EA	0.30	0.7	1.11	3.88	5.19	3.36	F	INTM
3P13EB	0.30	0.7	3.78	8.59	12.64	7.96	F	INTM
3P13EC	0.30	0.7	6.70	13.10	19.70	11.86	F	INTM
3P13ED	0.30	0.7	10.54	20.00	29.05	16.74	K	INTM

TABLE 7–8

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Flow Regime Obsvd at 50m	Taitel–Dukler
3P14AA	0.50	0	1.07	1.20	1.48	1.18	F	STWV
3P14AB	0.50	0	3.10	2.84	3.70	3.02	F	INTM
3P14AC	0.50	0	5.95	5.28	6.58	5.10	K	ANDP
3P14AD	0.50	0	9.40	8.33	9.85	7.09	K	ANDP
3P14BA	0.50	0.2	1.07	0.68	2.32	1.70	D	STWV
3P14BB	0.50	0.2	3.34	3.14	5.72	4.29	F	INTM
3P14BC	0.50	0.2	6.26	5.53	9.32	6.73	K	ANDP
3P14BD	0.50	0.2	9.51	8.27	13.41	9.16	K	ANDP
3P14CA	0.50	0.4	1.05	1.02	3.69	2.51	F	INTM
3P14CB	0.50	0.4	3.35	3.47	8.04	5.88	F	INTM
3P14CC	0.50	0.4	6.22	6.70	14.30	9.26	K	INTM
3P14CD	0.50	0.4	9.66	9.98	20.07	12.69	K	ANDP
3P14DA	0.50	0.5	1.10	3.35	5.20	3.41	F	INTM
3P14DB	0.50	0.5	3.42	7.71	11.60	7.40	F	INTM
3P14DC	0.50	0.5	6.63	9.22	19.45	11.80	K	INTM
3P14DD	0.50	0.5	9.30	12.23	25.13	15.12	K	INTM
3P14EA	0.50	0.7	1.22	8.72	11.72	7.09	F	INTM
3P14EB	0.50	0.7	2.54	13.24	19.25	10.72	F	INTM
3P14EC	0.50	0.7	3.94	17.39	20.25	14.17	K	INTM
3P14ED	0.50	0.7	6.88	26.47	39.90	20.65	K	INTM

TABLE 7–9

TEST CODE	V_{so} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Pressure Gradient Meas (mB/m)	E–D (mB/m)	E–O (mB/m)	Flow Regime Obsvd at 50m	Taitel–Dukler
3P15AA	1.0	0	1.11	3.26	5.07	3.43	F	INTM
3P15AB	1.0	0	3.55	7.45	11.17	7.29	F	INTM
3P15AC	1.0	0	7.54	14.28	19.76	12.23	K	INTM
3P15AD	1.0	0	10.42	17.90	25.74	15.64	K	INTM
3P15BA	1.0	0.2	1.18	3.96	7.56	4.86	F	INTM
3P15BB	1.0	0.2	3.59	7.47	15.73	7.54	K	INTM
3P15BC	1.0	0.2	7.14	13.84	26.72	15.29	F	INTM
3P15BD	1.0	0.2	10.22	18.31	35.43	19.84	K	INTM
3P15CA	1.0	0.4	1.24	5.93	12.06	7.37	F	INTM
3P15CB	1.0	0.4	2.51	9.00	19.10	10.74	F	INTM
3P15CC	1.0	0.4	3.96	11.95	26.37	14.32	F	INTM
3P15CD	1.0	0.4	7.50	20.72	43.90	22.17	K	INTM
3P15DA	1.0	0.5	1.30	13.00	16.00	9.76	F	INTM
3P15DB	1.0	0.5	2.66	18.47	25.60	13.61	F	INTM
3P15DC	1.0	0.5	4.14	23.67	34.90	17.83	F	INTM
3P15DD	1.0	0.5	6.58	30.90	50.00	24.40	K	INTM

TABLE 7–10

TEST CODE	INCLINE (Deg)	V_{sL} (m/s)	λ_w (–)	V_{sg} at 39m (m/s)	Slug Length $L_{S\,min}$ (m)	$L_{S\,max}$ (m)	$L_{S\,av}$ (m)	Slug Front Velocity V_S (m/s)	V_S / V_m (–)	Slug Frequency Obsvd (1/s)	Greg–Scott (1/s)
SLUG I1	+1	0.3	0	1.04	0.43	2.46	0.92	1.83	1.37	0.235	0.330
SLUG I2	+1	0.3	0.5	1.06	0.44	2.12	0.92	2.08	1.53	0.254	0.330
SLUG I3	+1	1.0	0	1.11	0.51	1.10	0.89	2.81	1.33	0.910	0.920
SLUG I4	−1	1.0	0.5	1.20	0.53	1.05	0.84	2.72	1.24	0.910	0.920
SLUG I5	+1	1.0	0.5	1.02	0.48	0.97	0.71	3.00	1.49	1.030	0.910
SLUG I6	−1	1.0	0.5	1.10	0.50	1.18	0.68	2.71	1.29	0.960	0.910

TABLE 7–11

TEST CODE	OIL	INITIAL OIL VOL (L)	IMPR SPEED (rpm)	GAS FLOW (L/MIN)	FOAM VOL (L)
FOAM01	No 1	9.0	400	11.2	3.5
FOAM02				20.0	4.0
FOAM03				27.0	4.2
FOAM04			500	11.2	3.9
FOAM05				20.0	4.8
FOAM06				27.0	5.2
FOAM07			600	11.2	4.2
FOAM08				20.0	5.0
FOAM09				27.0	5.4
FOAM10	A	9.0	400	11.2	1.3
FOAM11				20.0	2.0
FOAM12				27.0	3.0
FOAM13			500	11.2	1.7
FOAM14				20.0	2.6
FOAM15				27.0	3.4
FOAM16			600	11.2	1.9
FOAM17				20.0	2.8
FOAM18				27.0	3.5
FOAM19	B	9.0	400	11.2	1.4
FOAM20				20.0	2.0
FOAM21				27.0	2.5
FOAM22			500	11.2	1.8
FOAM23				20.0	2.8
FOAM24				27.0	3.5
FOAM25			600	11.2	2.3
FOAM26				20.0	3.3
FOAM27				27.0	4.0
FOAM28	FORTIES CRUDE	9.0	400	11.2	3.8
FOAM29				20.0	6.8
FOAM30				27.0	9.5
FOAM31			500	11.2	3.8
FOAM32				20.0	7.0
FOAM33				27.0	9.9
FOAM34			600	11.2	4.0
FOAM35				20.0	6.5
FOAM36				27.0	9.6

FOR ALL TESTS:

Impeller Position = 60mm From Base
Gas Injection Point = 40mm From Base

TABLE A1–1

TEST CODE	OIL	INITIAL OIL VOL (L)	INPUT WATER CUT (-)	IMPR SPEED (rpm)	COND. PROBE VALUE (mV)	DISP PHASE (O/W)
DISP01	No 1	6.0	0.33	300	0	W
DISP02			0.36		0	W
DISP03			0.38		>50	O
DISP04			0.39		>50	O
DISP05			0.41		>50	O
DISP06	A	6.0	0.25	300	0	W
DISP07			0.32		0	W
DISP08			0.33		>50	O
DISP09			0.35		>50	O
DISP10			0.38		>50	O
FOR ALL TESTS: Impeller Position = 60mm From Base						

TABLE A1-2

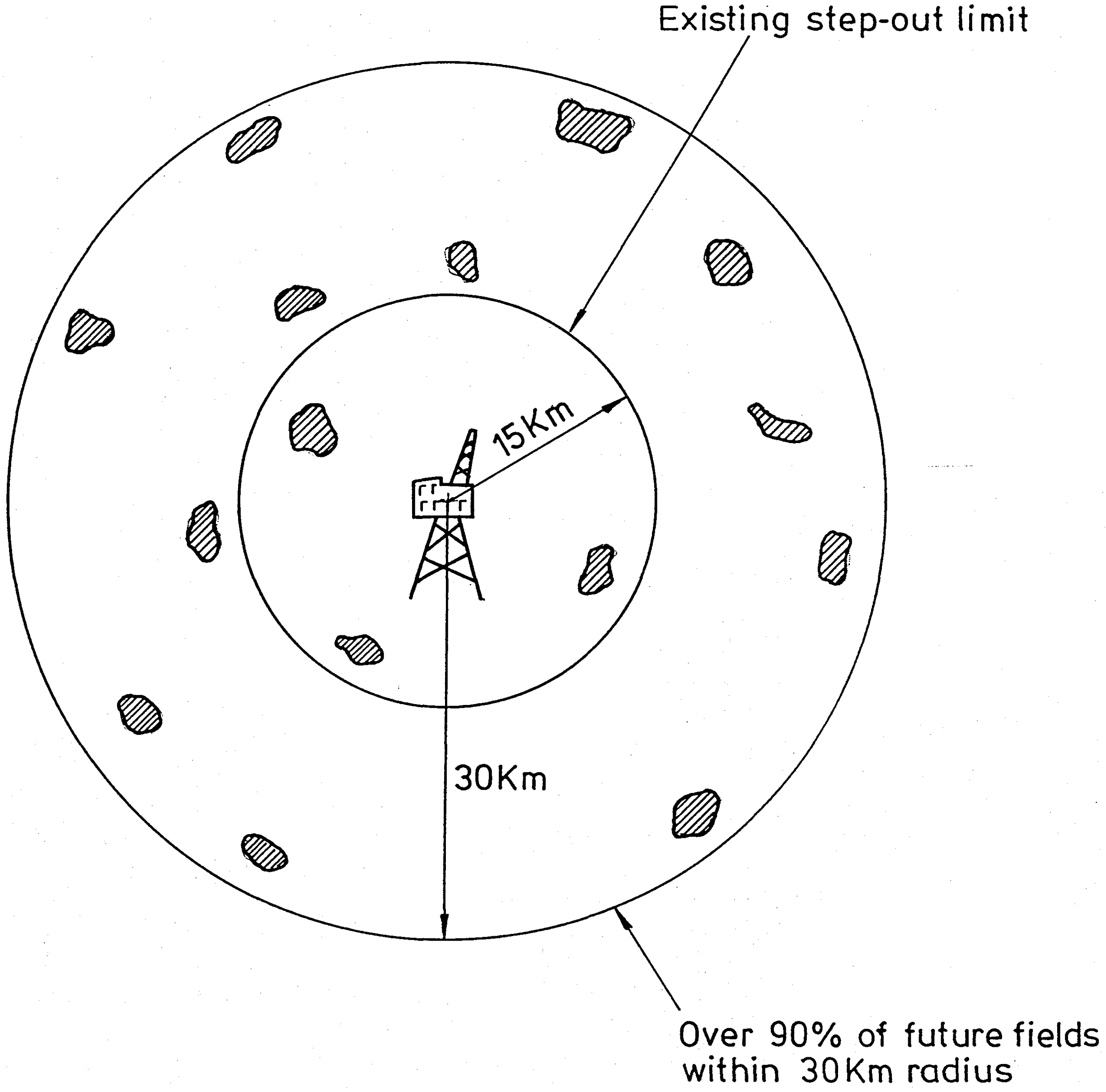


Fig 2.1 Tie-in of Small Fields to Existing Infrastructure

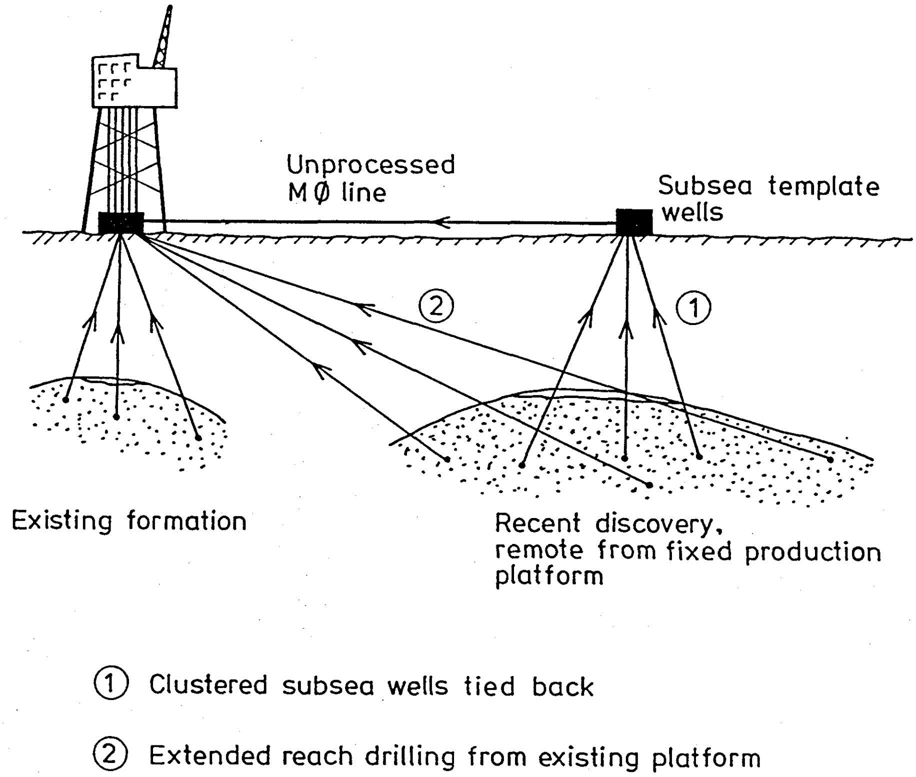


Fig 2.2 Subsea ERD Solutions

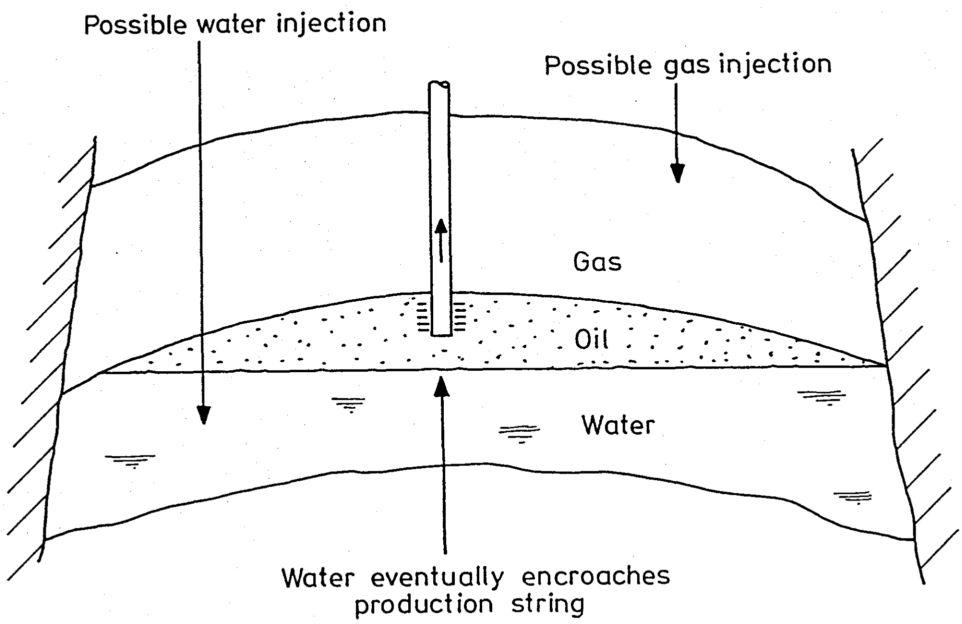


Fig 2.3 Simplified Reservoir Structure

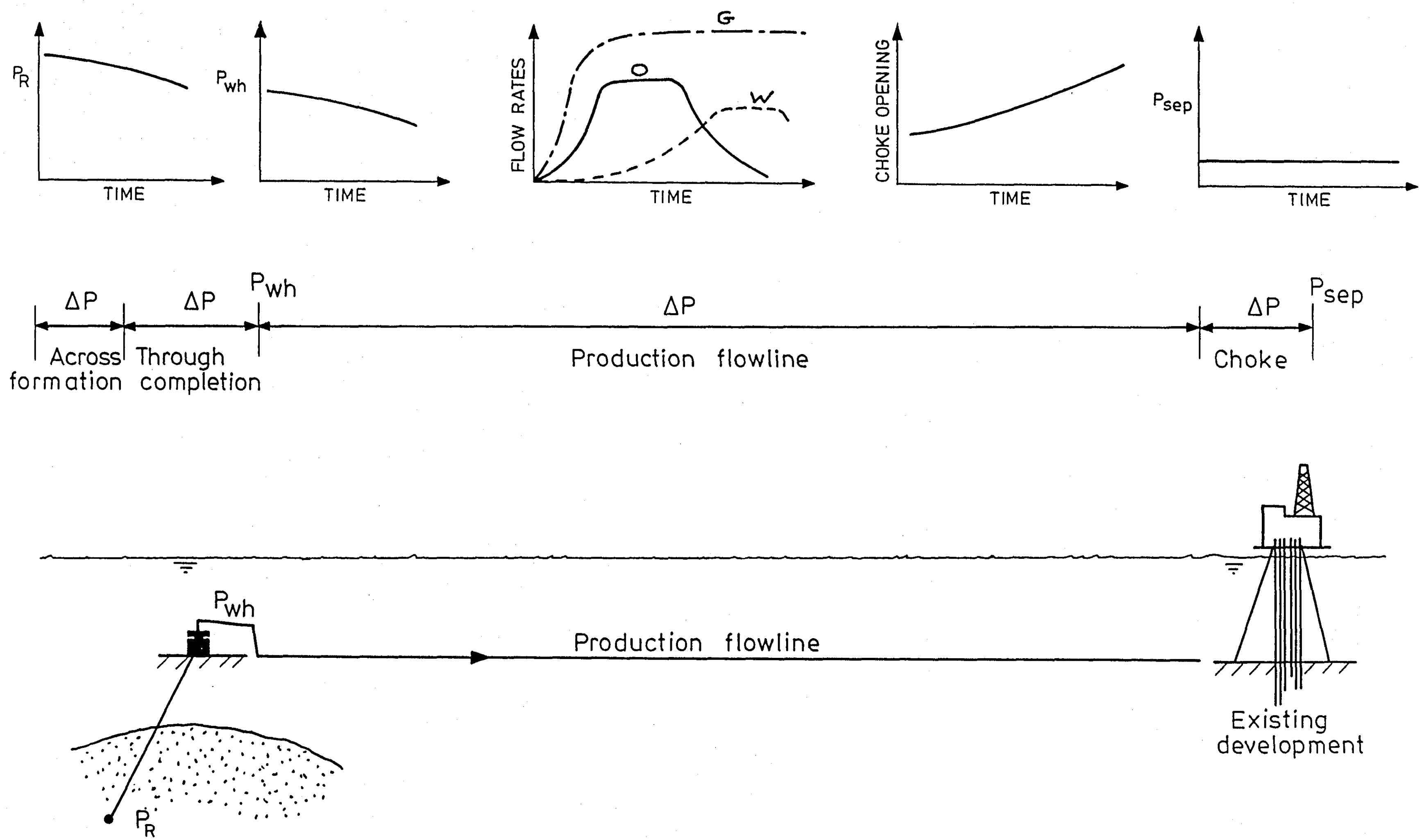


Fig 2.4 Production Constraints for a Subsea Tie-back

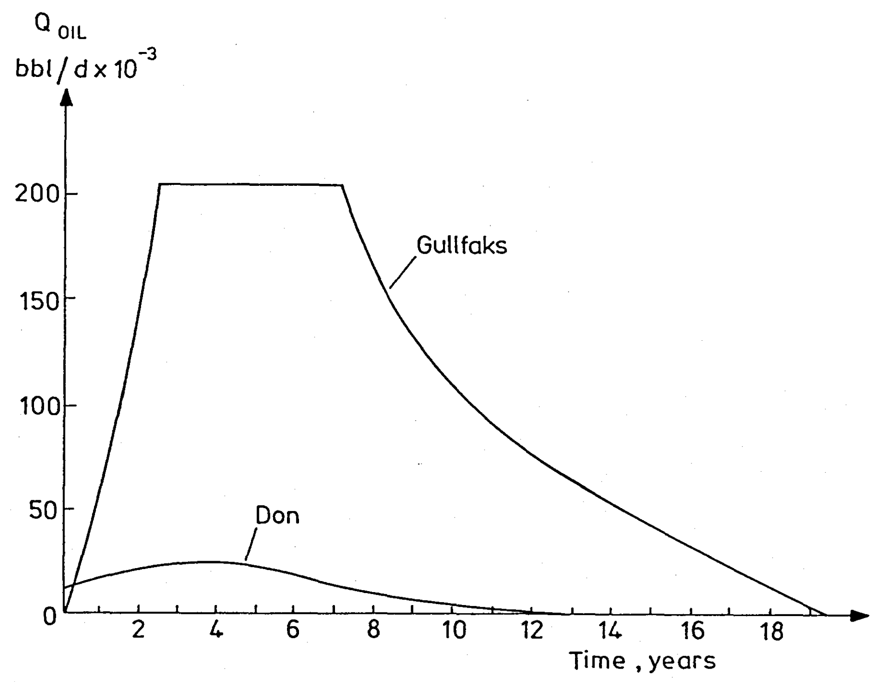


Fig 2.5 Production Profile for 2 Oil Fields

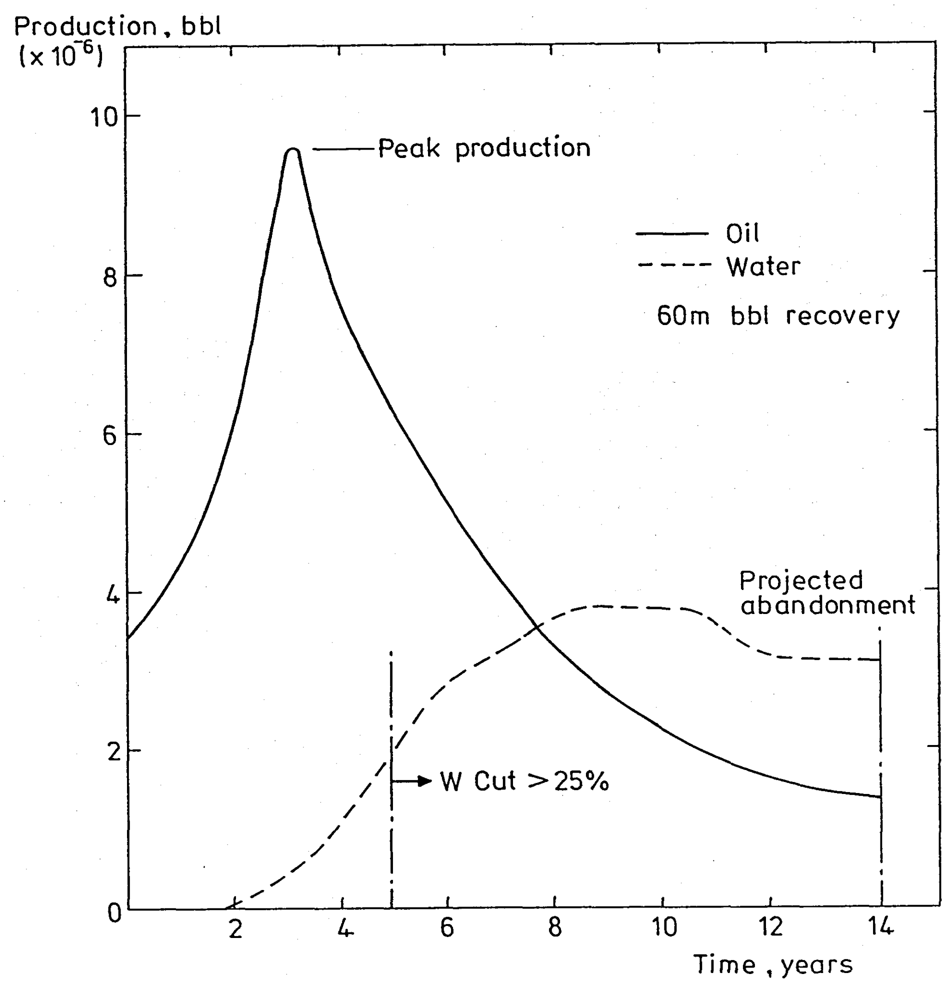


Fig 2.6 Typical Oil and Water Production Profiles

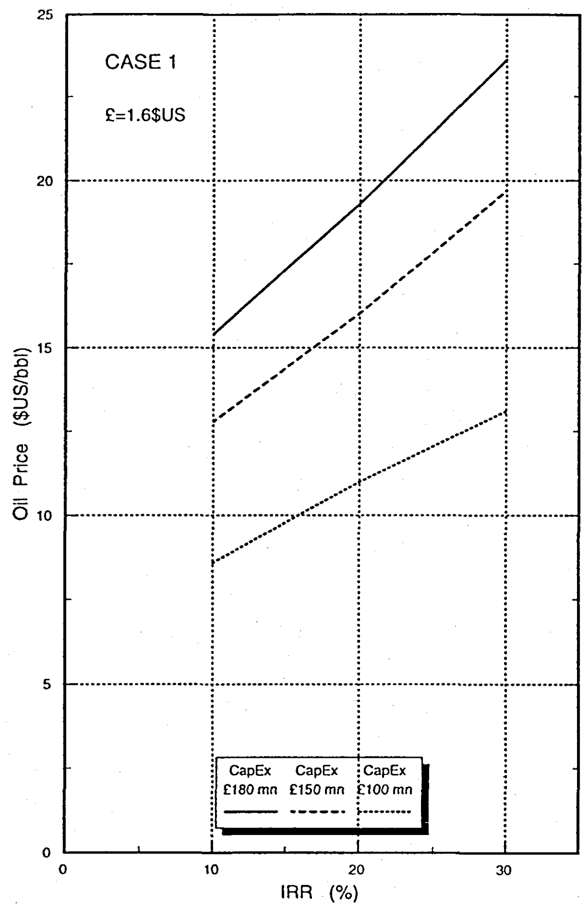


Fig 2.7 Economic Analysis Case 1

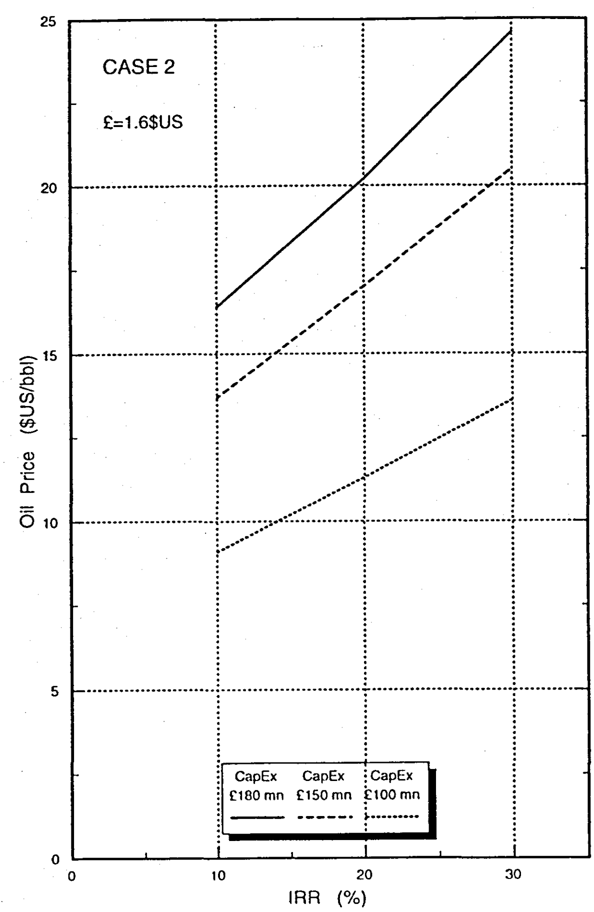


Fig 2.8 Economic Analysis Case 2

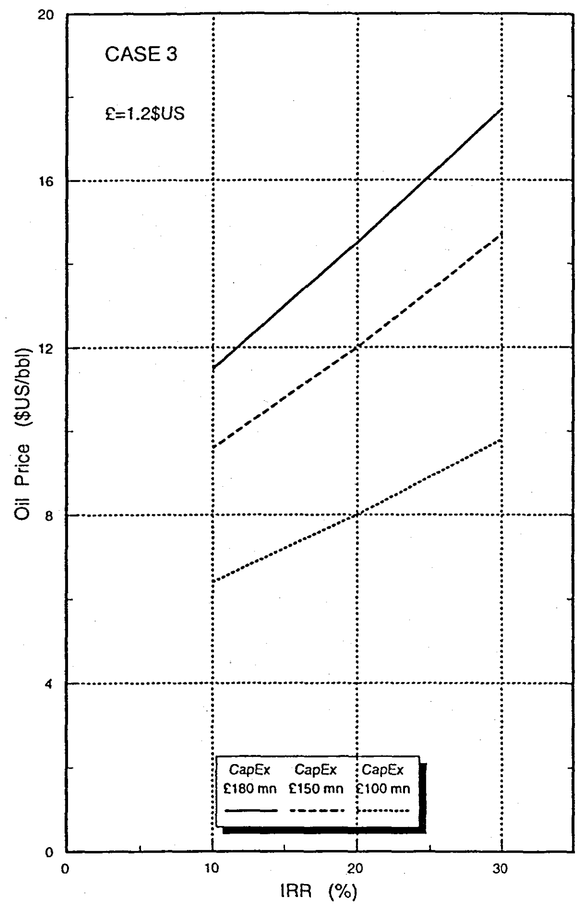


Fig 2.9 Economic Analysis Case 3

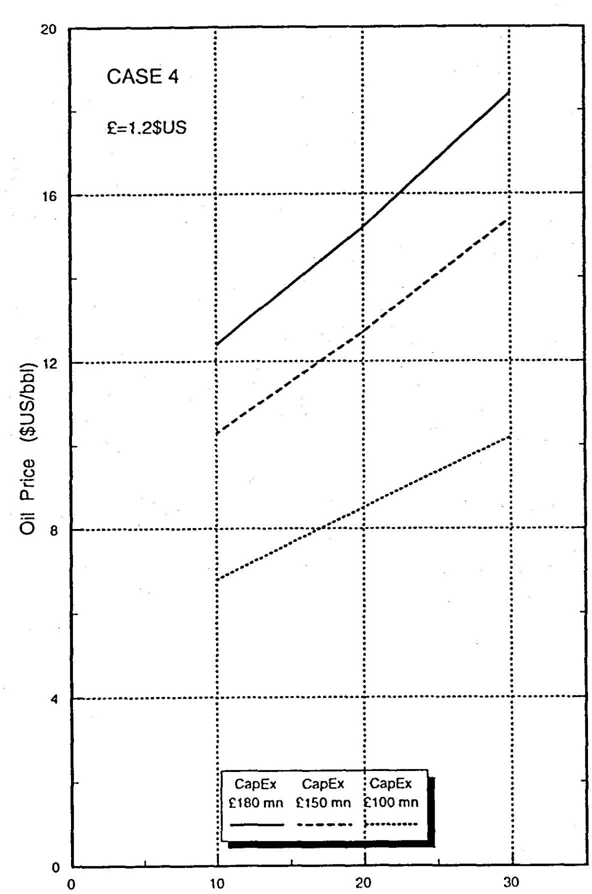


Fig 2.10 Economic Analysis Case 4

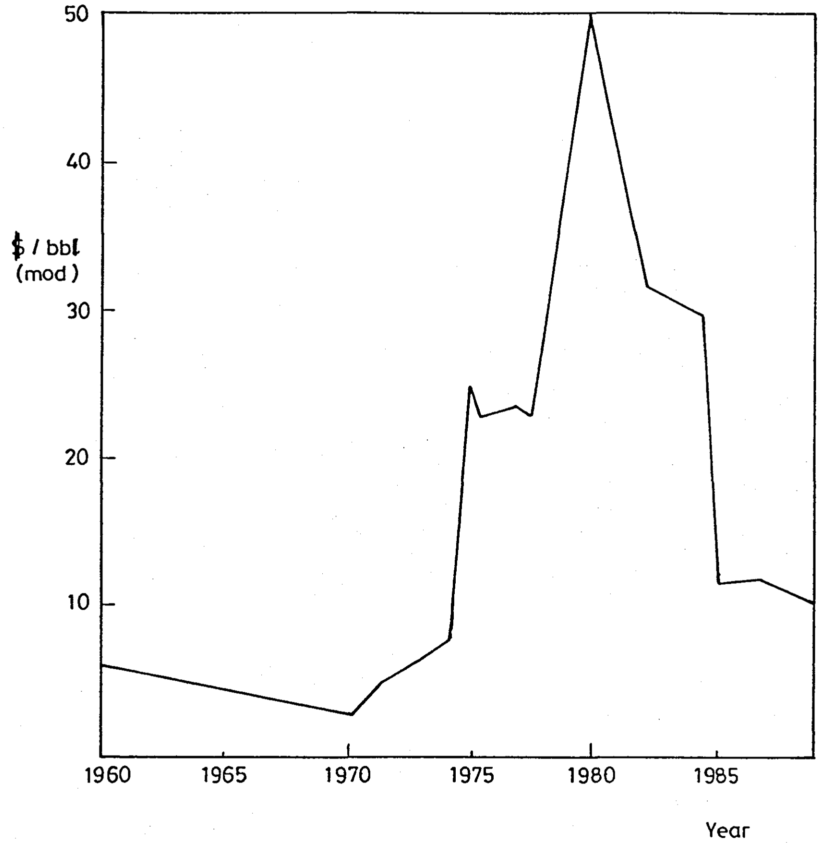


Fig 2.11 Crude Oil Price History

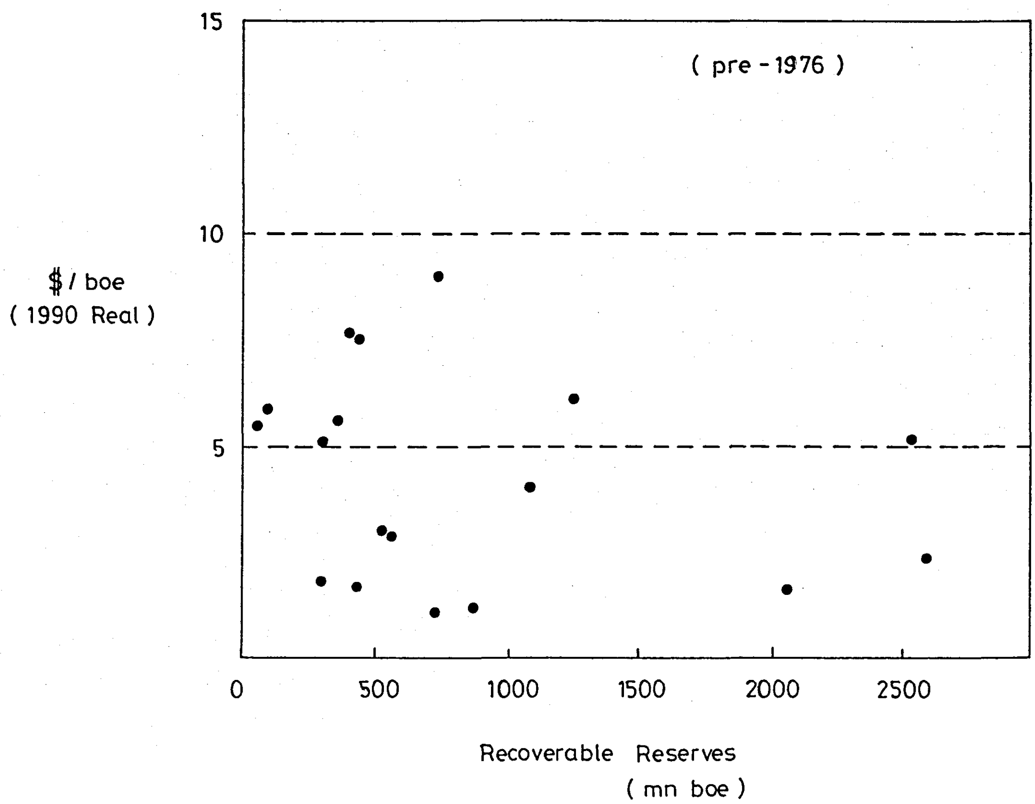


Fig 2.12 Economic Indicator of Early N.Sea Fields

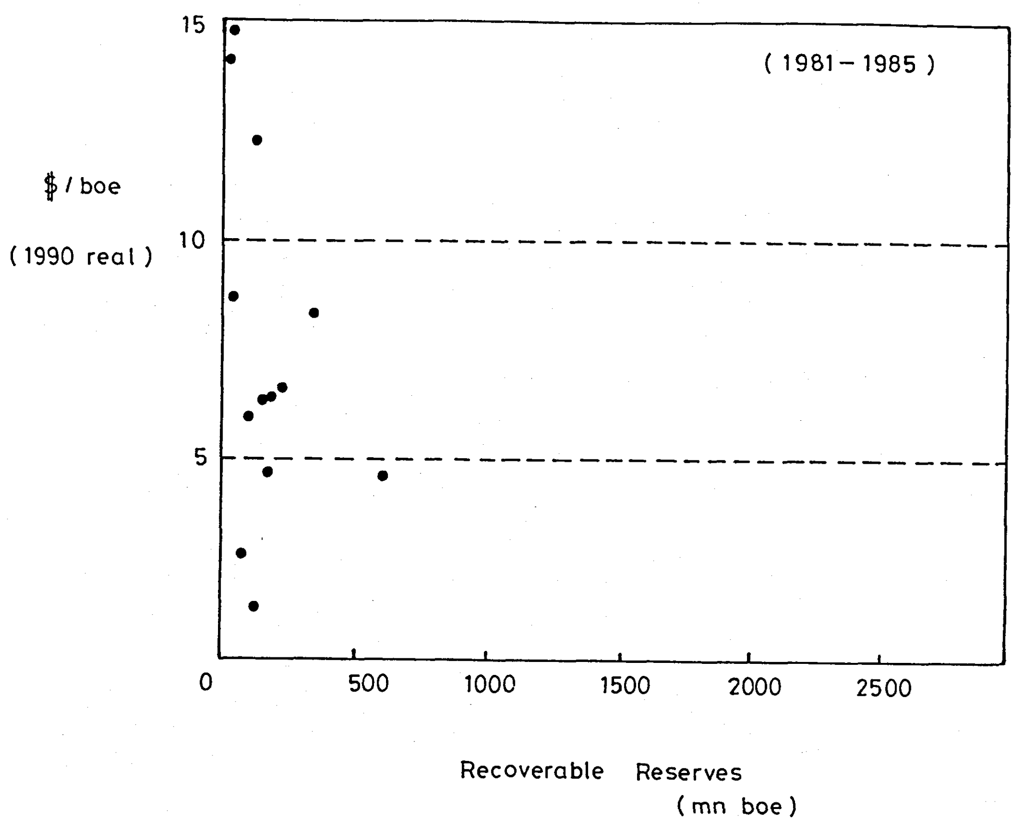


Fig 2.13 Economic Indicators Prior to Oil Price Fall

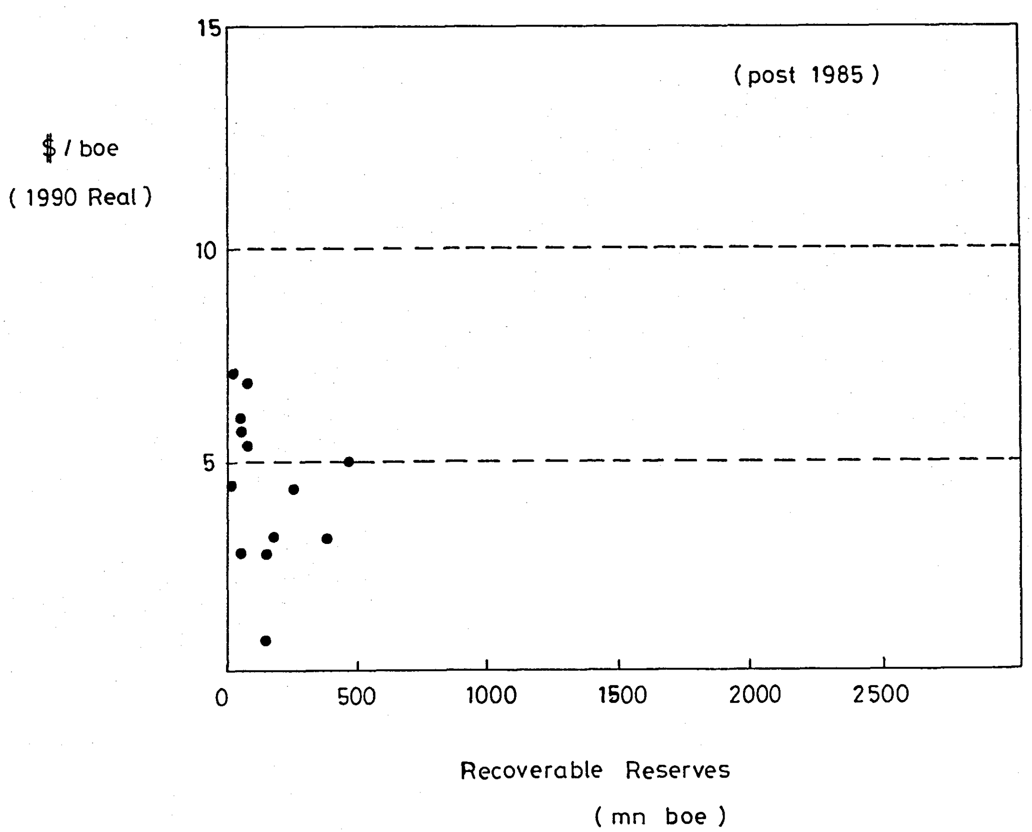


Fig 2.14 Economics Following Oil Price Collapse

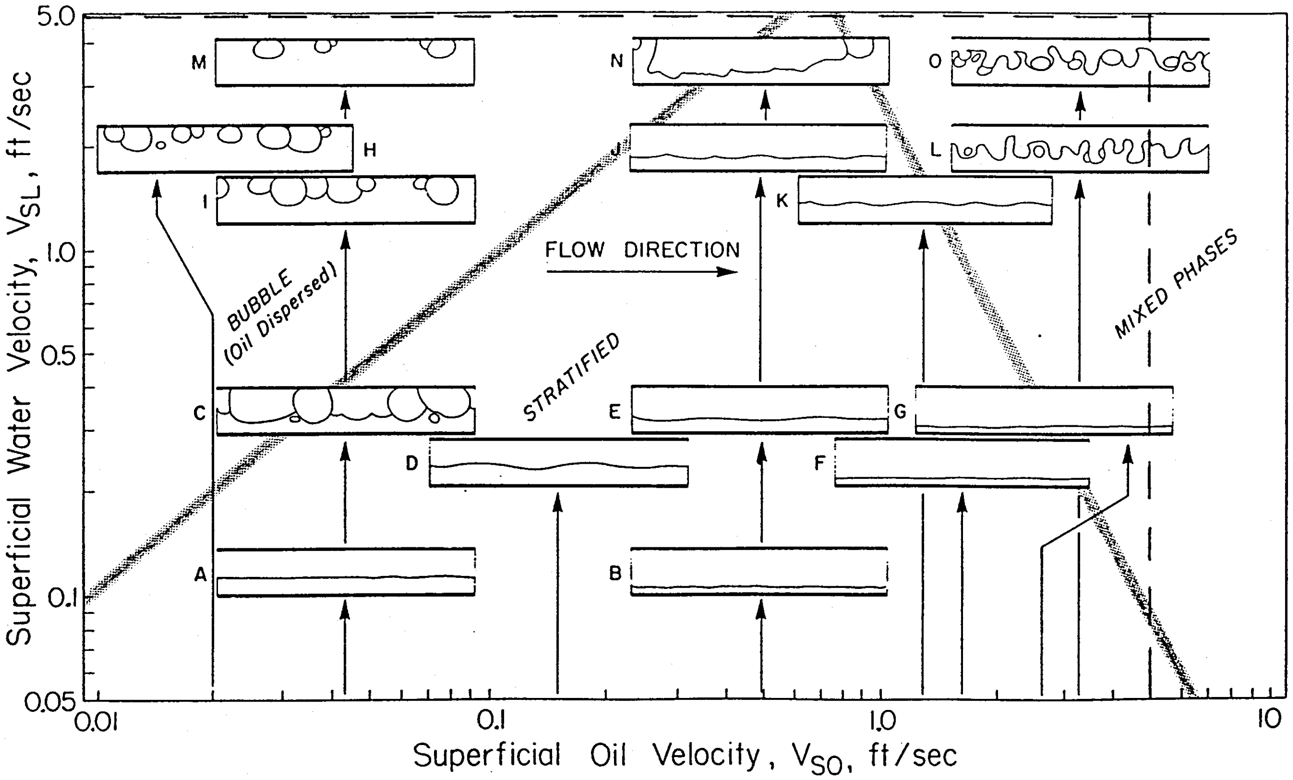


Fig 3.1 Oil–Water Flow Regime Map of Russell et al(1959)

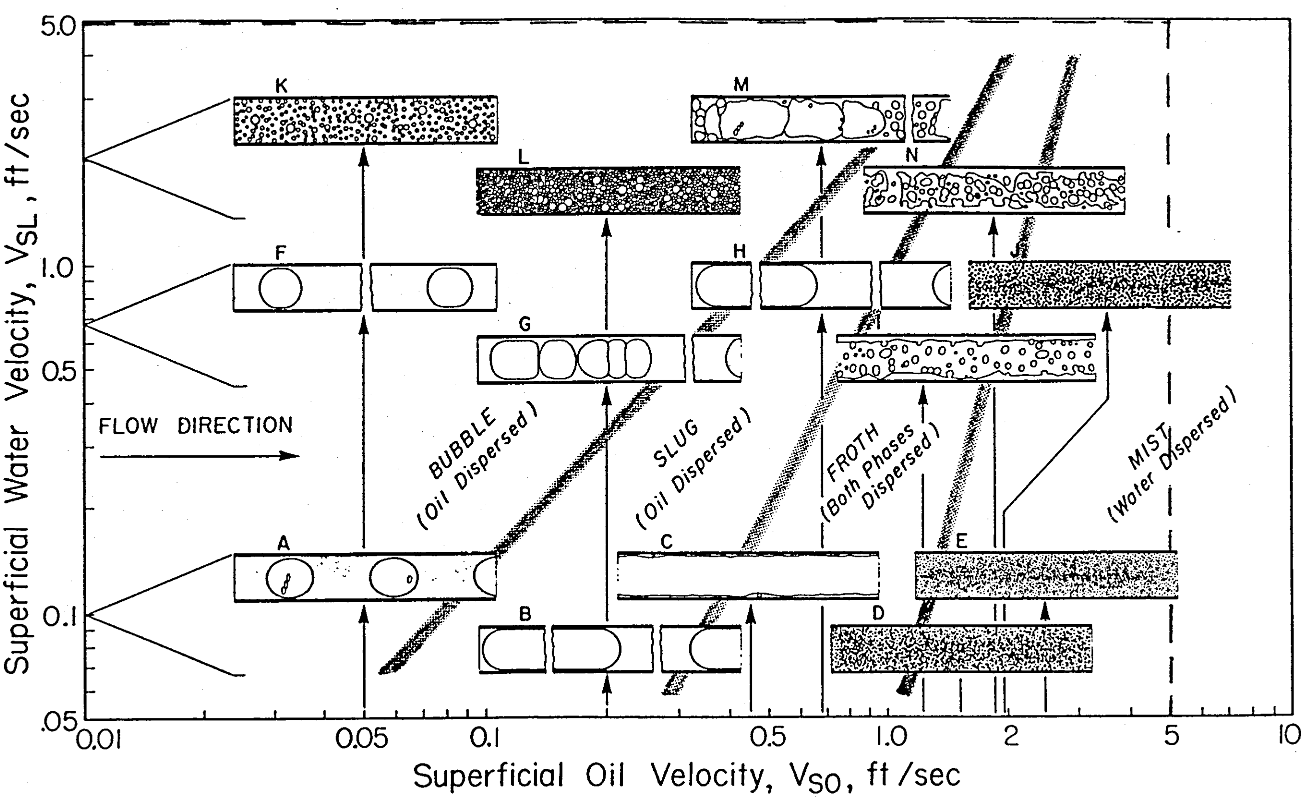


Fig 3.2 Oil–Water Flow Regime Map of Charles et al (1961)

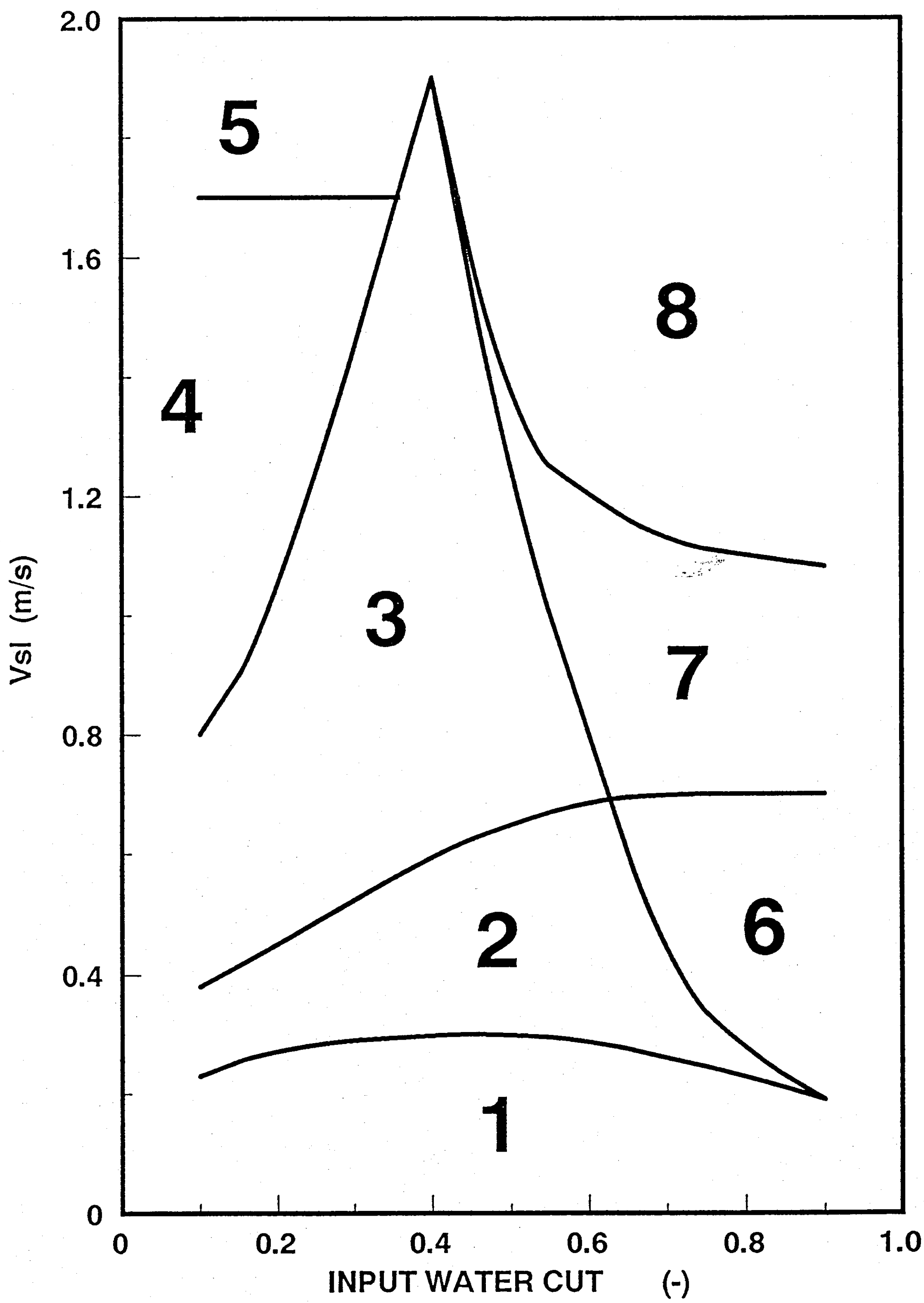


Fig 3.3 Oil–Water Flow Regime Map of Guzhov et al(1973)

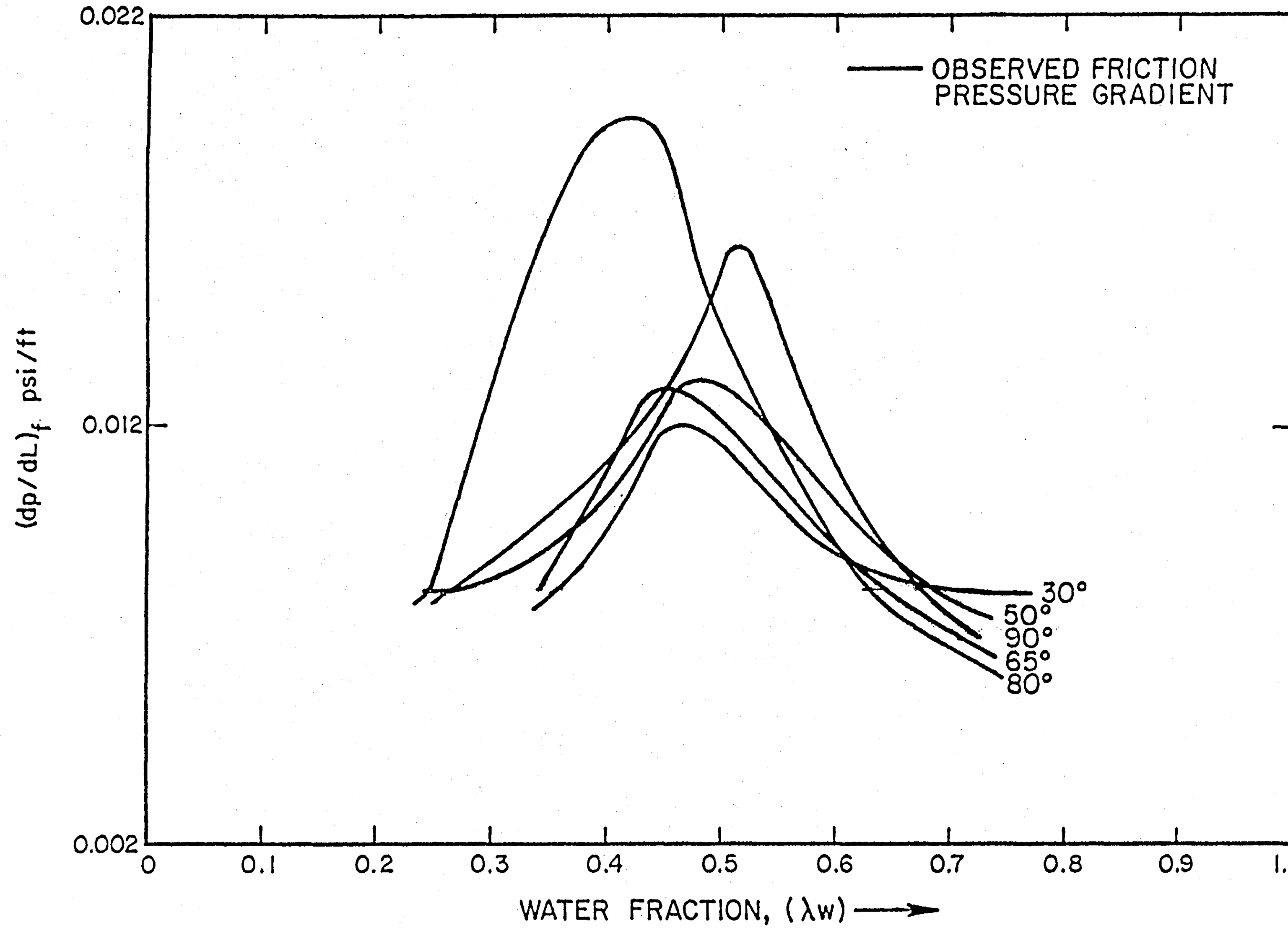


Fig 3.4 Oil–Water Pressure Inclined Pressure Drop Data of Mukhopadhyay(1977)

FLOW PATTERN CODE

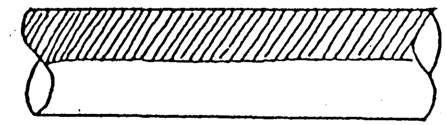
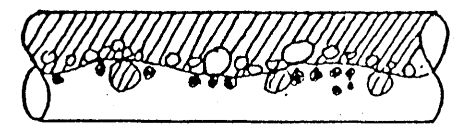
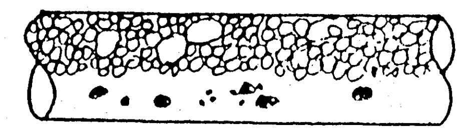
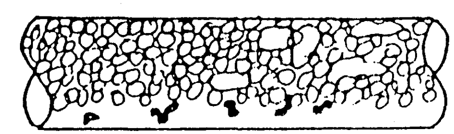
Oil Dominant	Water Dominant	Description	Sketch
	A	Segregated - no mixing at the interface.	
	B	Semi-segregated - some mixing at the interface.	
C	K	Semi-mixed - segregated flow of a dispersion and "free" phase. Bubbly interface. Dispersion volume less than $\frac{1}{2}$ the total pipe volume.	 Example: water-in-oil dispersion with a "free" water phase
D	L	Mixed - same as the above coding but with the dispersion occupying more than $\frac{1}{2}$ the pipe volume.	 Example: water-in-oil dispersion with a "free" water phase

Fig 3.5 (a) Oil–Water Flow Regime Classifications of Oglesby(1979)

FLOW PATTERN CODE

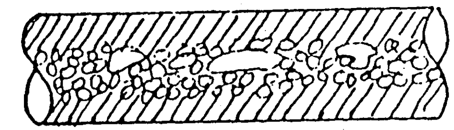
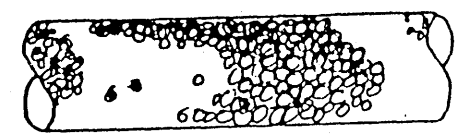
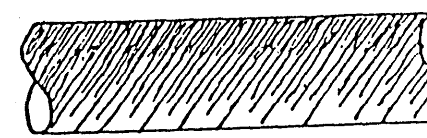
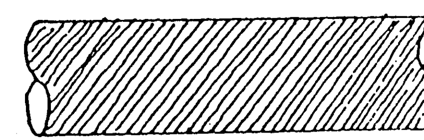
Oil Dominant	Water Dominant	Description	Sketch
G	J	Annular or concentric - core of one phase within the other phase.	 Example: water-core in an oil layer
H	I	Slug - phases alternately occupying the pipe volume as a free phase or as a dispersion.	
E	M	Semi-dispersed - some vertical gradient of fluid concentrations in the mixture.	
F	N	Fully dispersed homogeneous flow.	

Fig 3.5 (b) Oil–Water Flow Regime Classifications of Oglesby(1979)

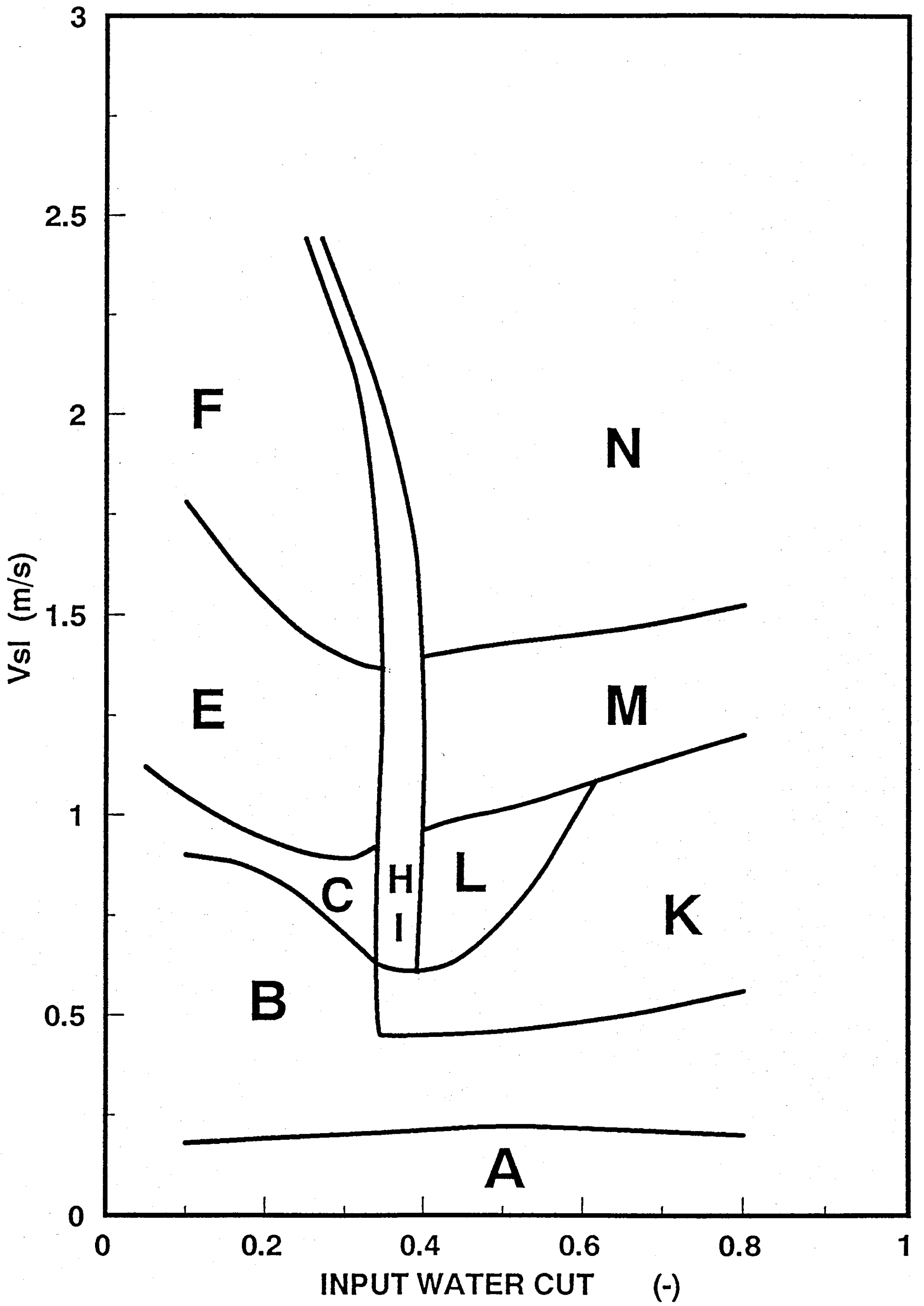


Fig 3.6 Oil–Water Flow Regime Map for 32cP Oil, Oglesby(1979)

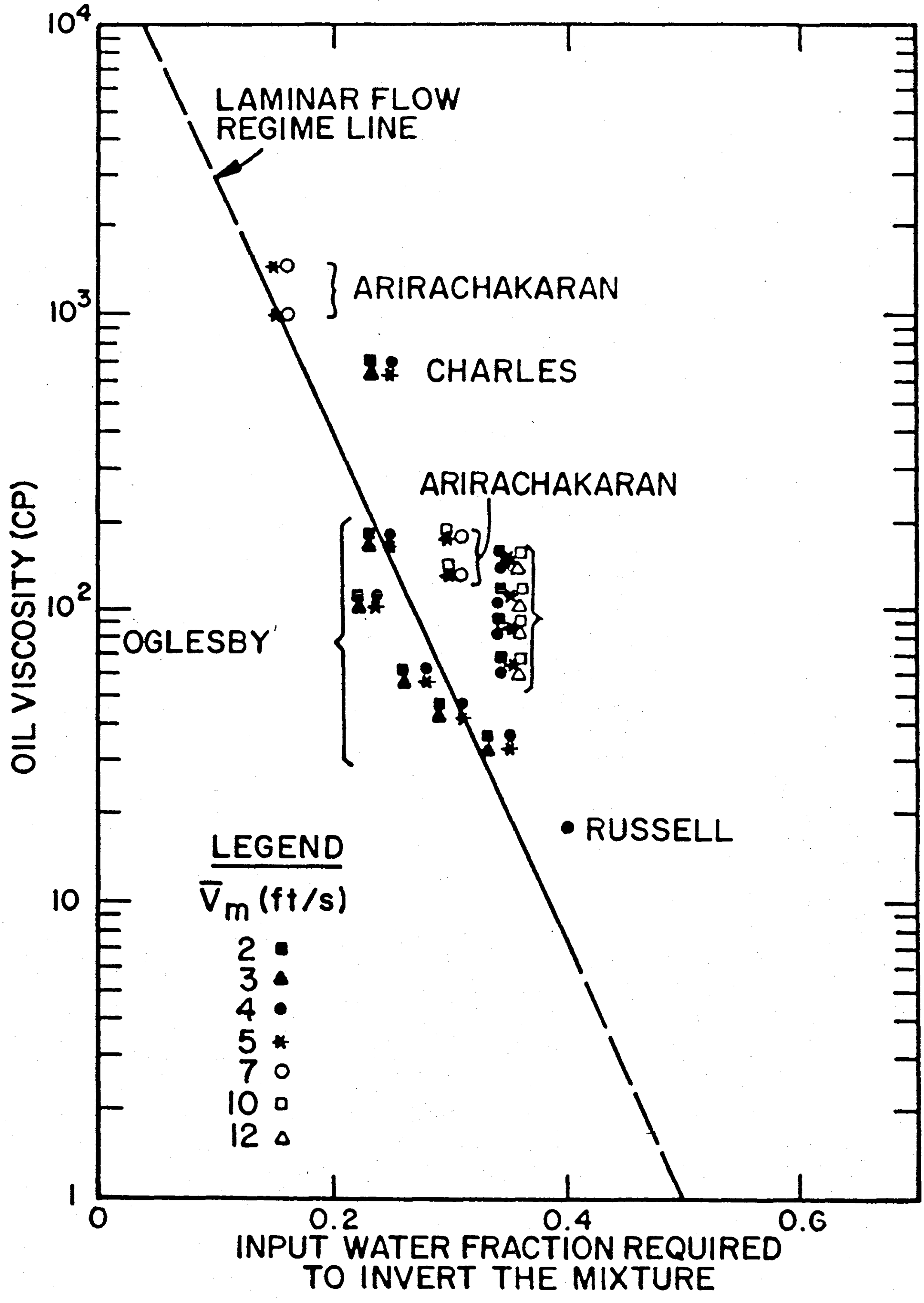


Fig 3.7 Oil–Water Emulsion Inversion Relation of Arirachakaran et al(1989)

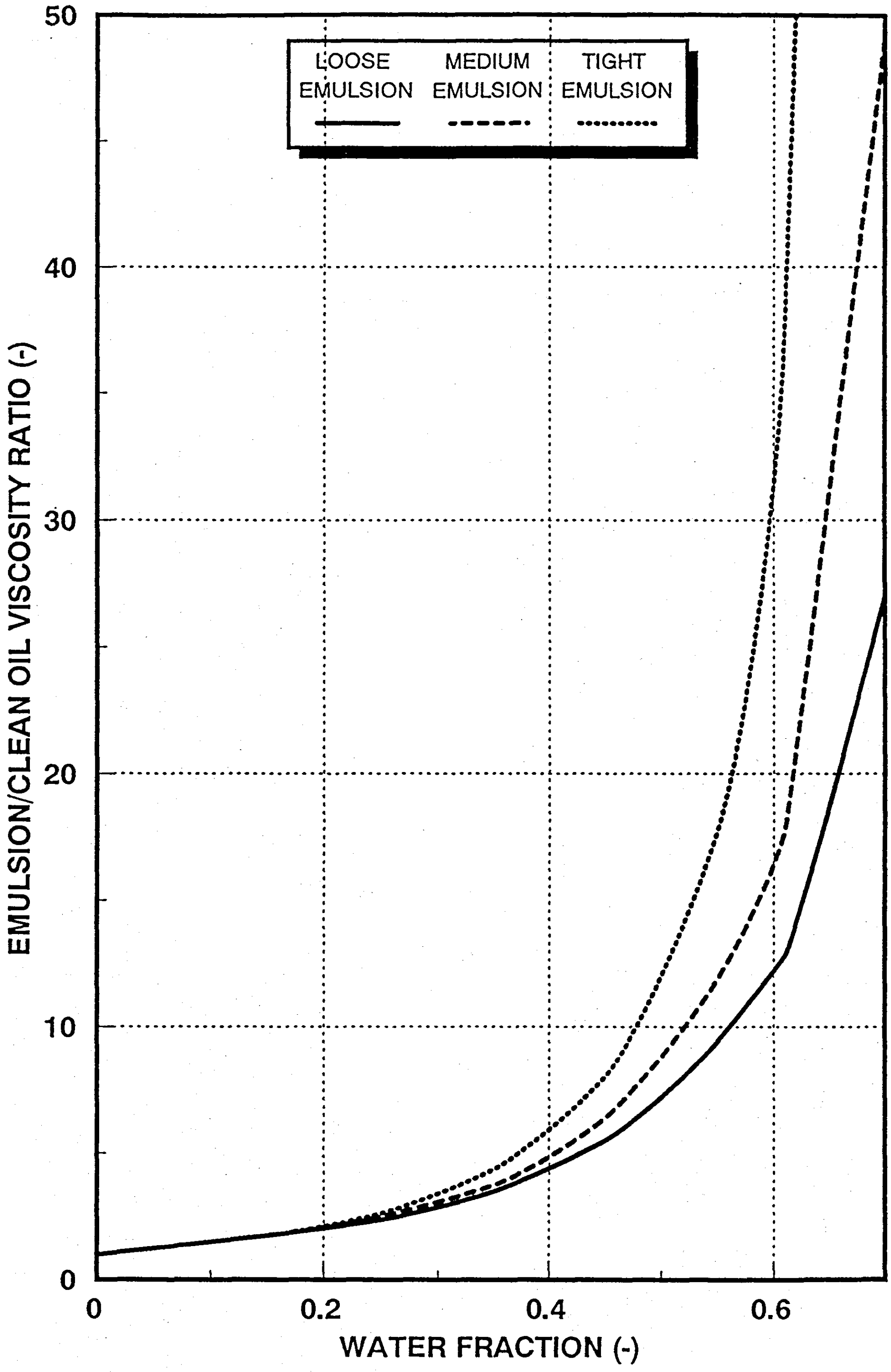


Fig 3.8 Water–Oil Emulsion Viscosity Relation of Woelflin(1947)

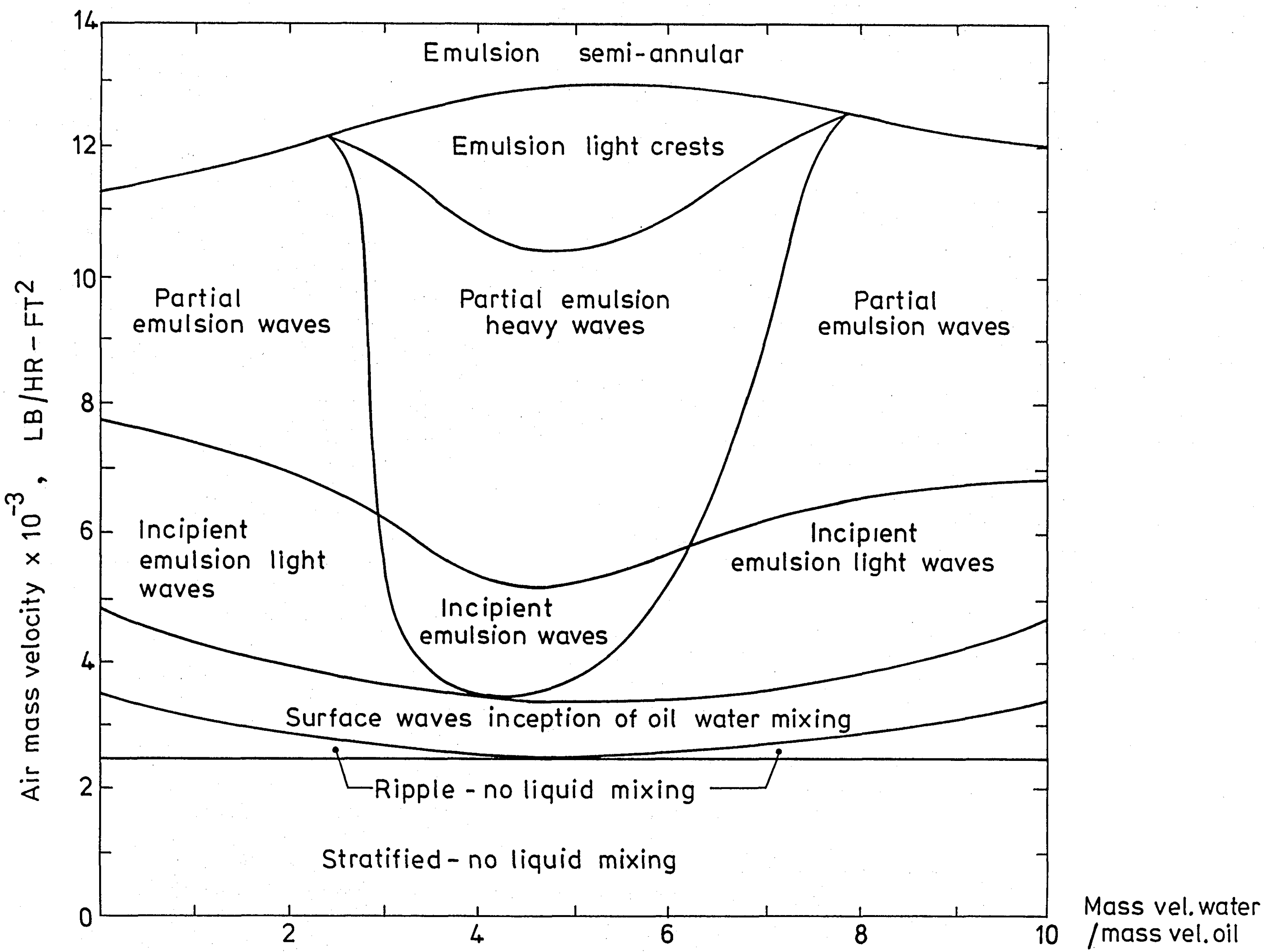


Fig 3.9 Oil/Water/Gas Flow Regime Map of Sobocinski(1955)

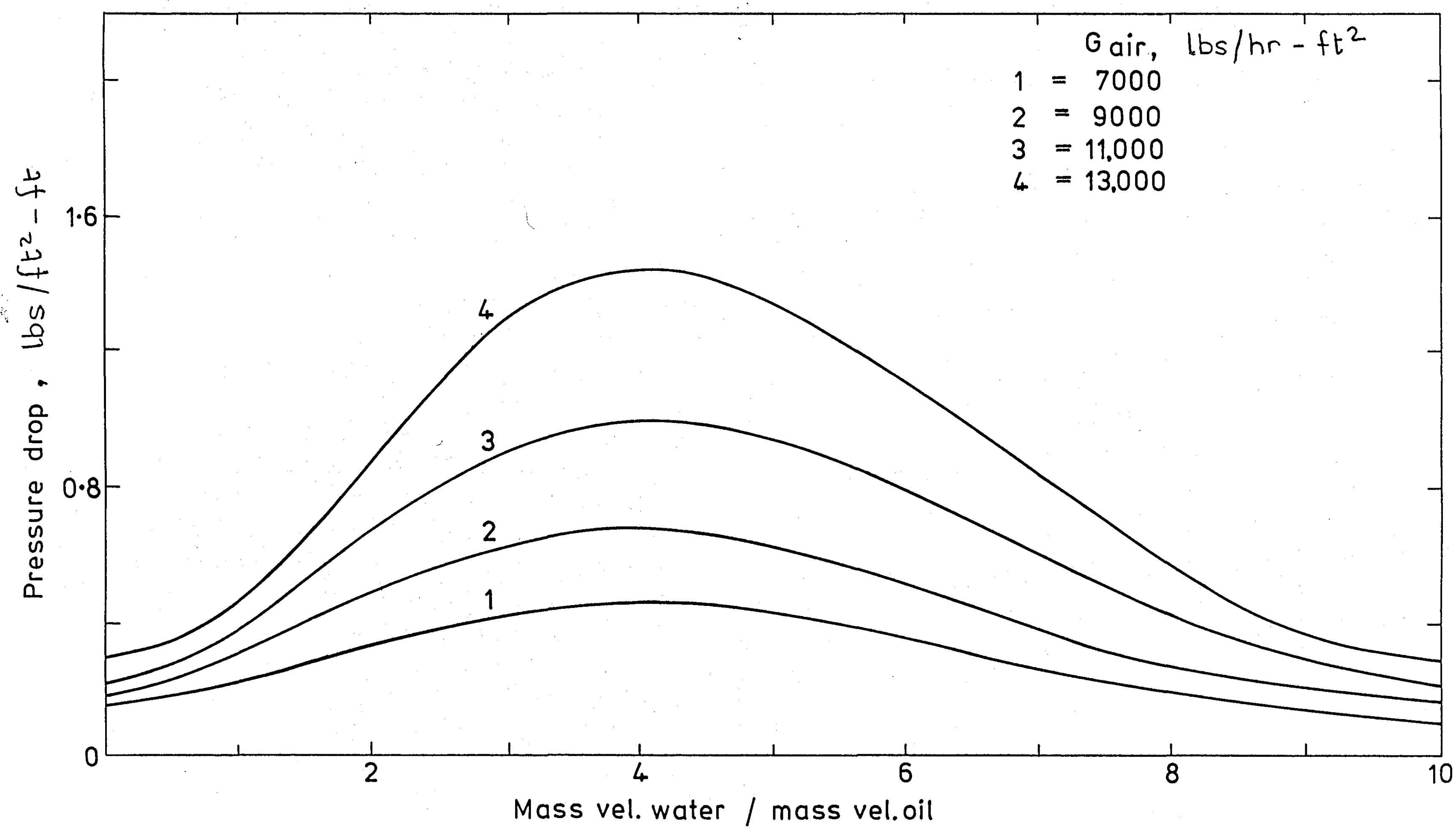


Fig 3.10 Oil/Water/Gas Pressure Drop Data of Sobocinski(1955)

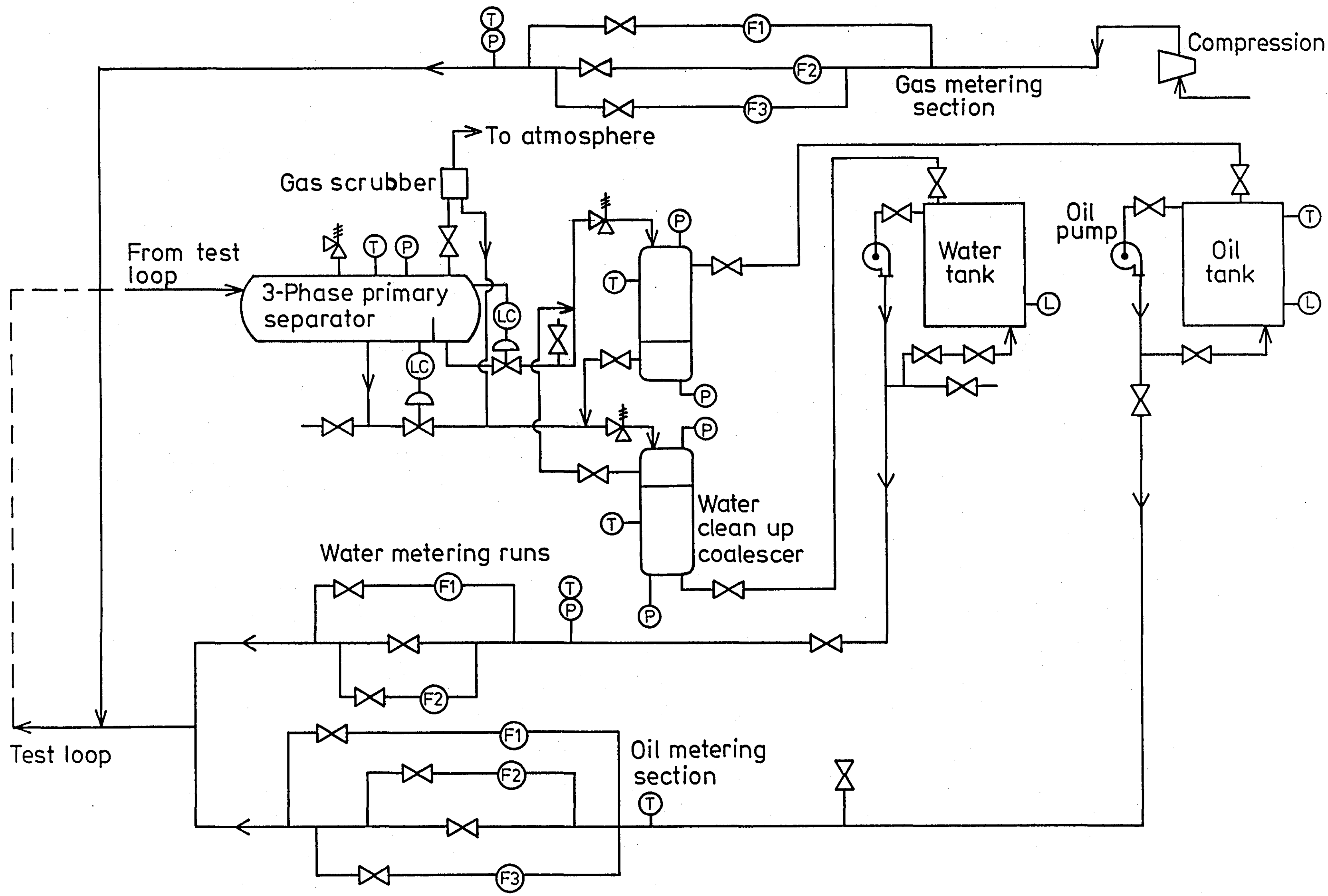


Fig 4.1 Overall System Flowchart

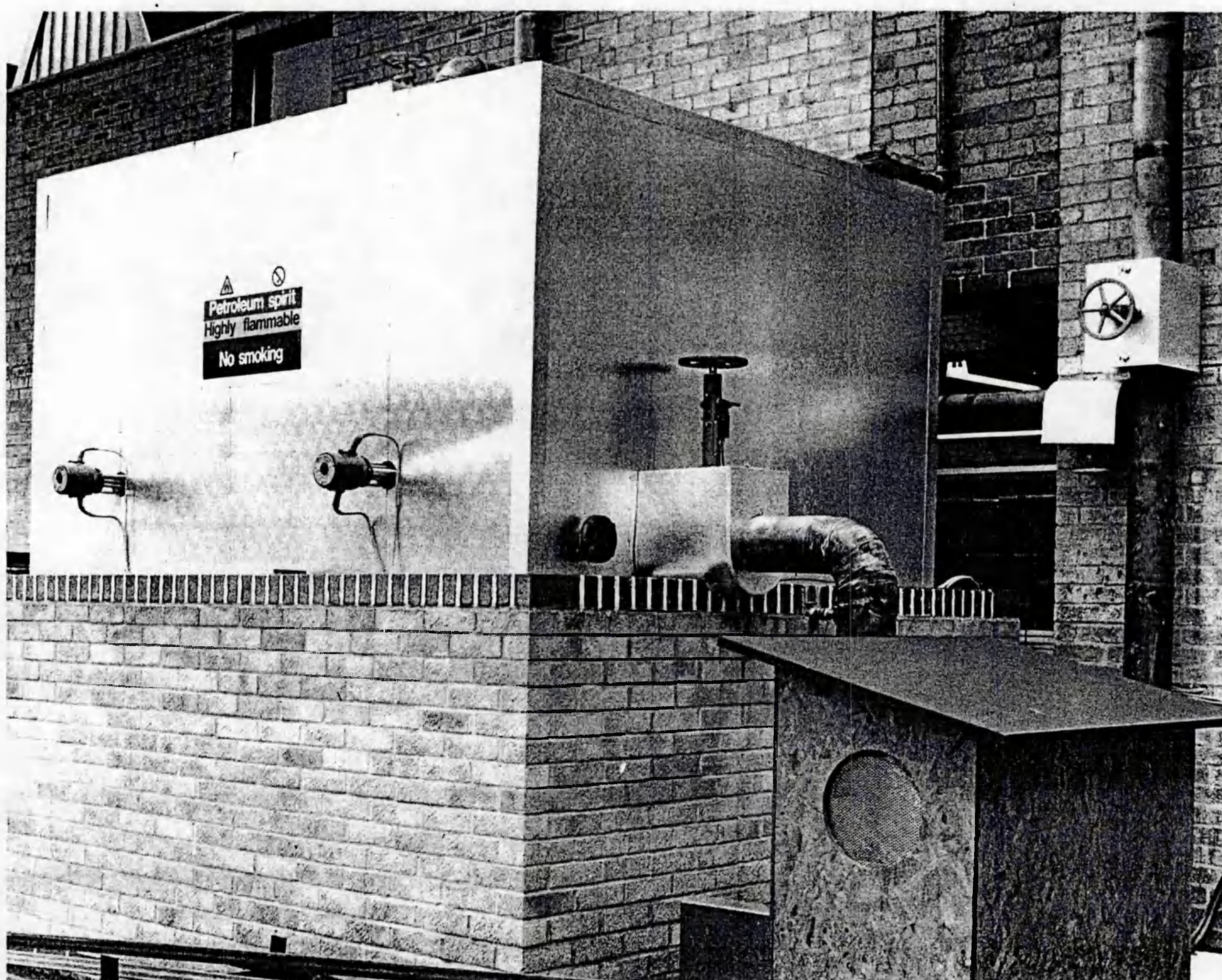


Fig 4.2 Oil Storage Tank

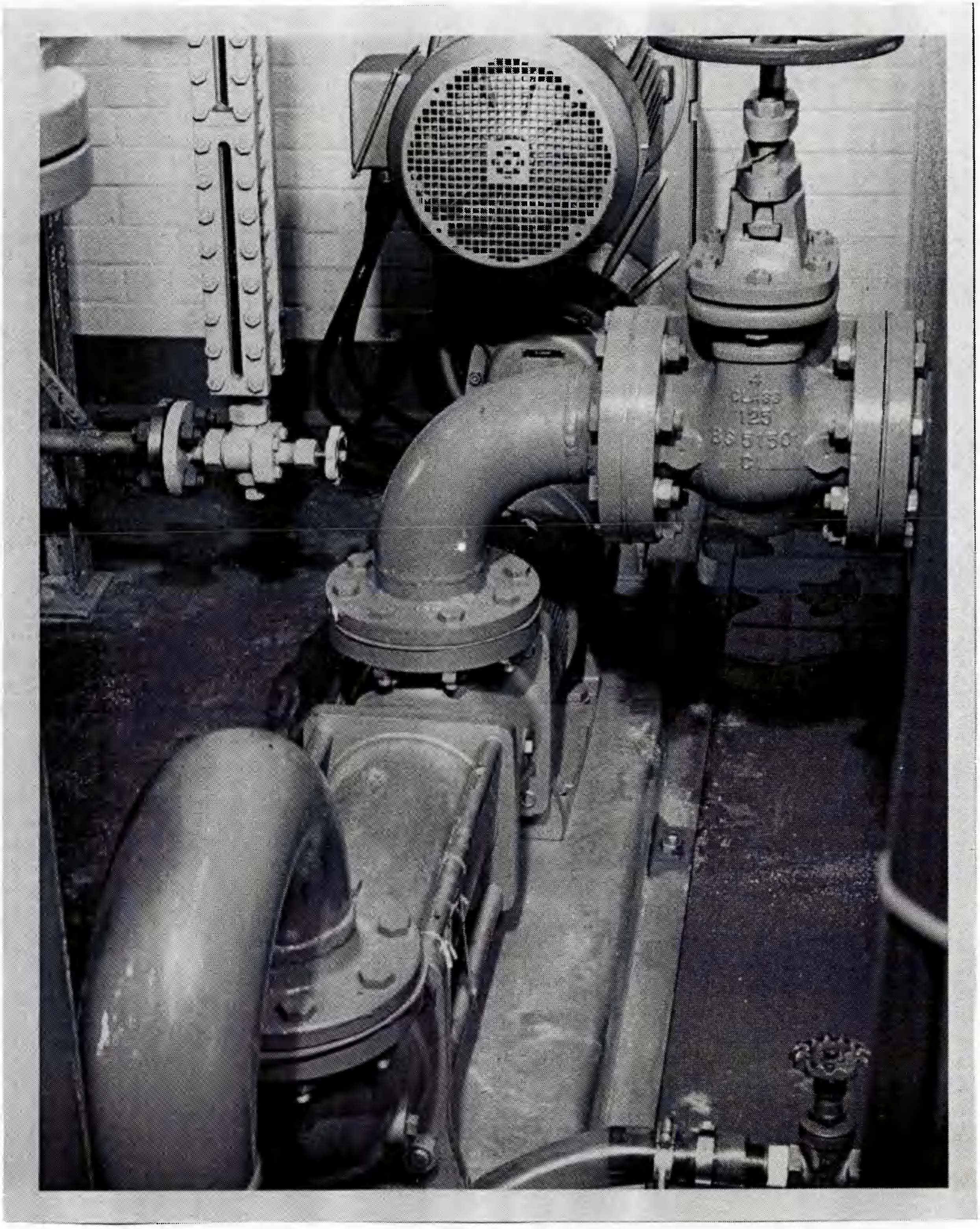


Fig 4.3 Water Mono–Pump

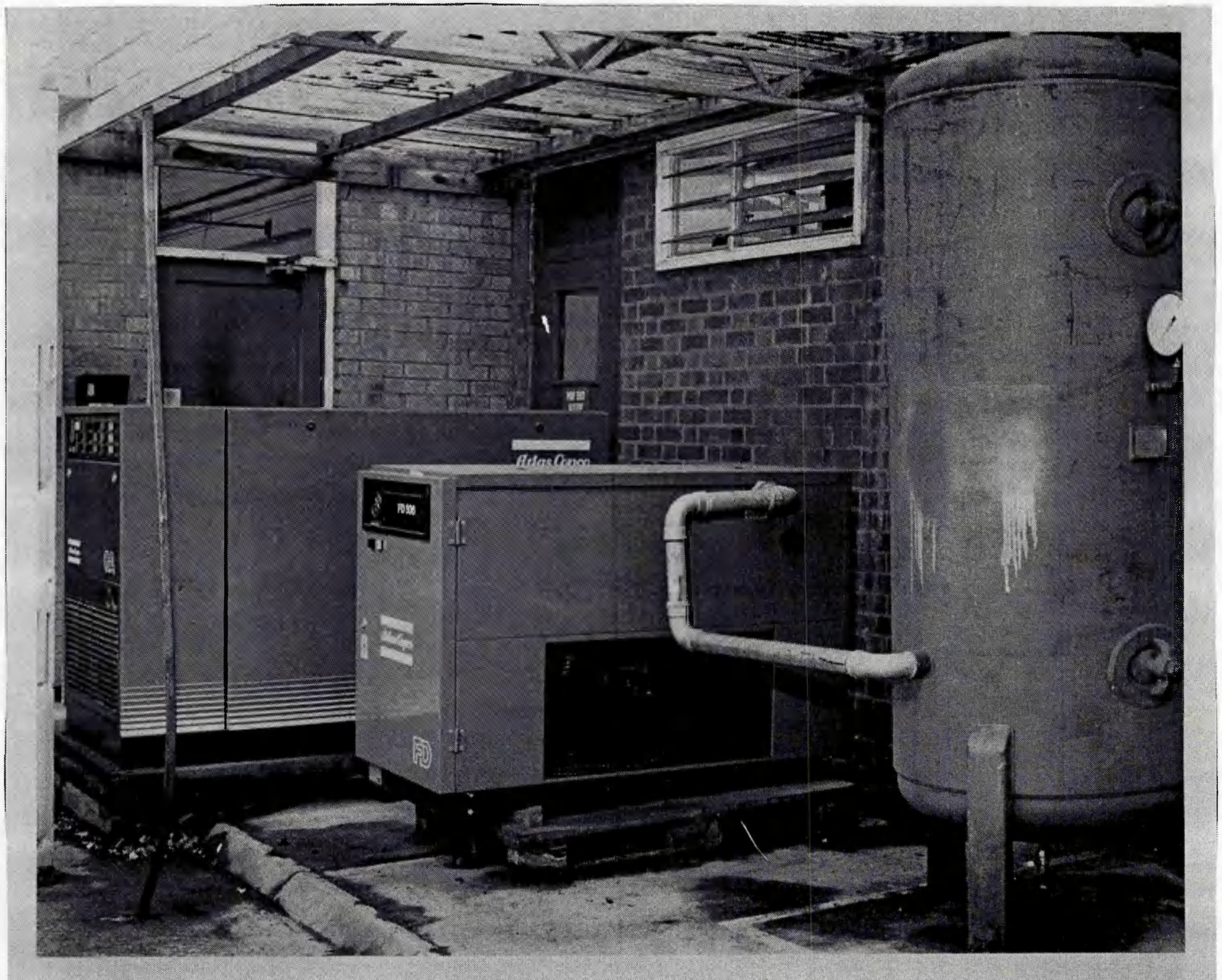


Fig 4.4 Air Compressor System

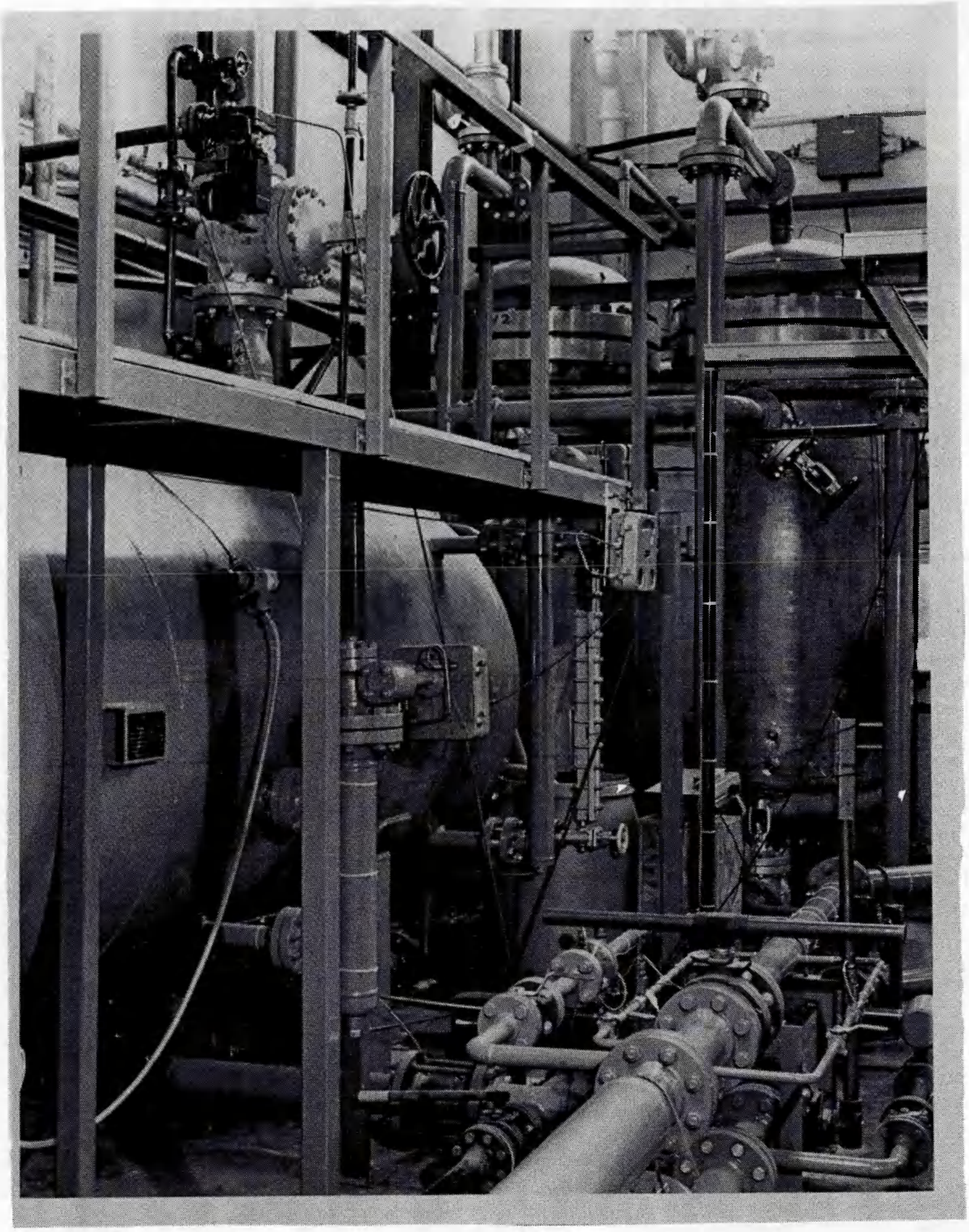


Fig 4.5 Oil/Water Metering,Mixing and Separation Area

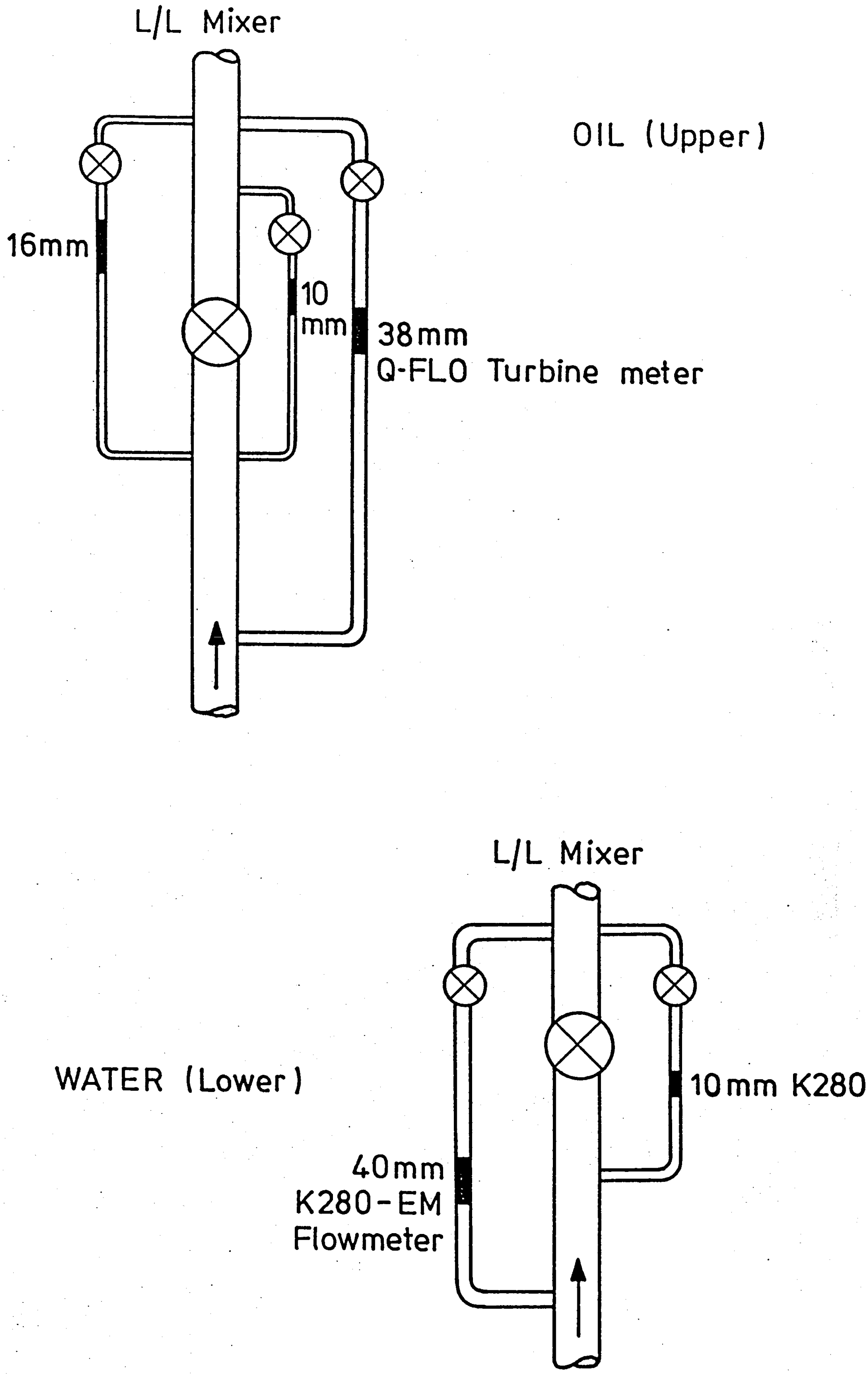


Fig 4.6 Liquid Flowmeter Runs

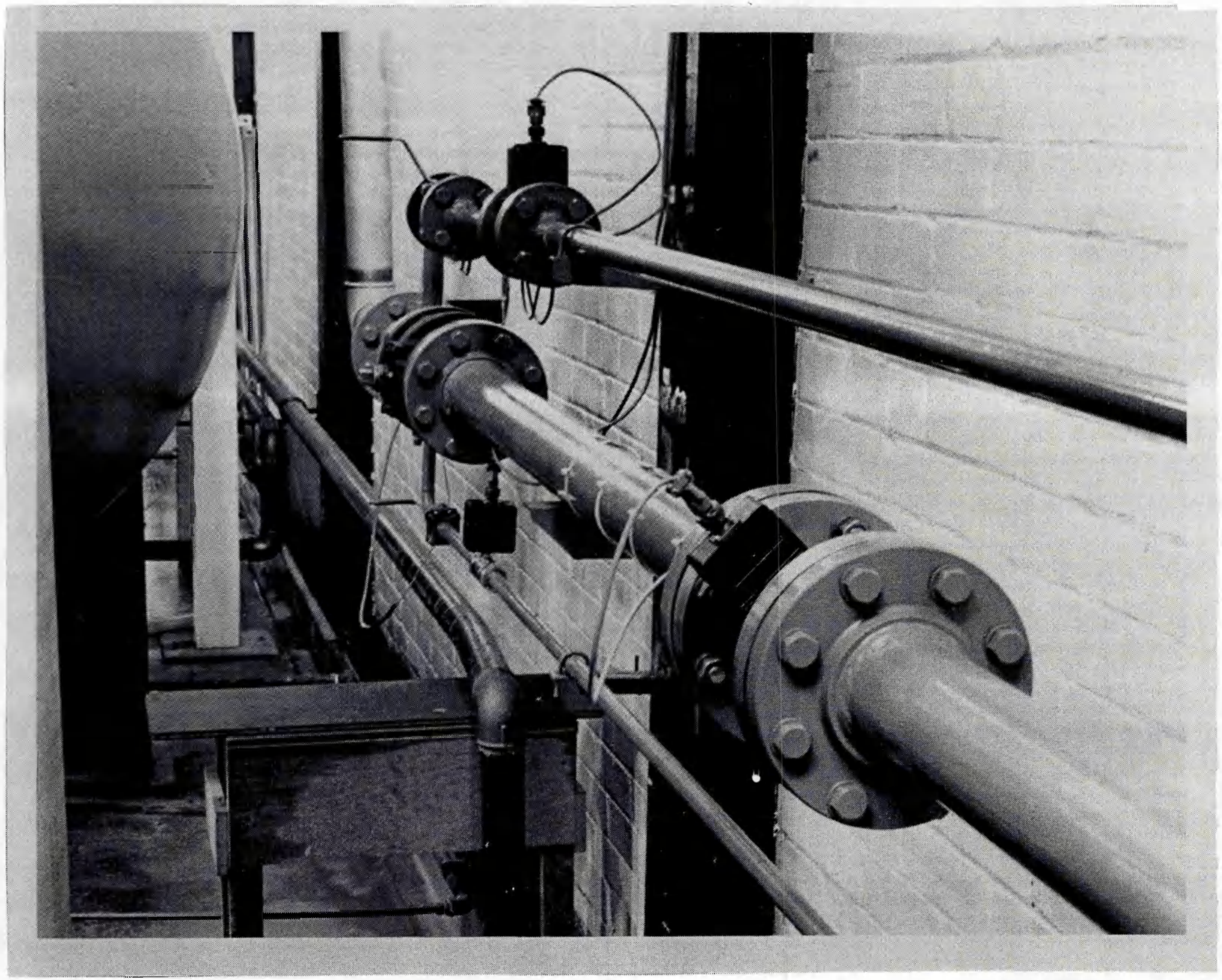


Fig 4.7 Gas Metering Section

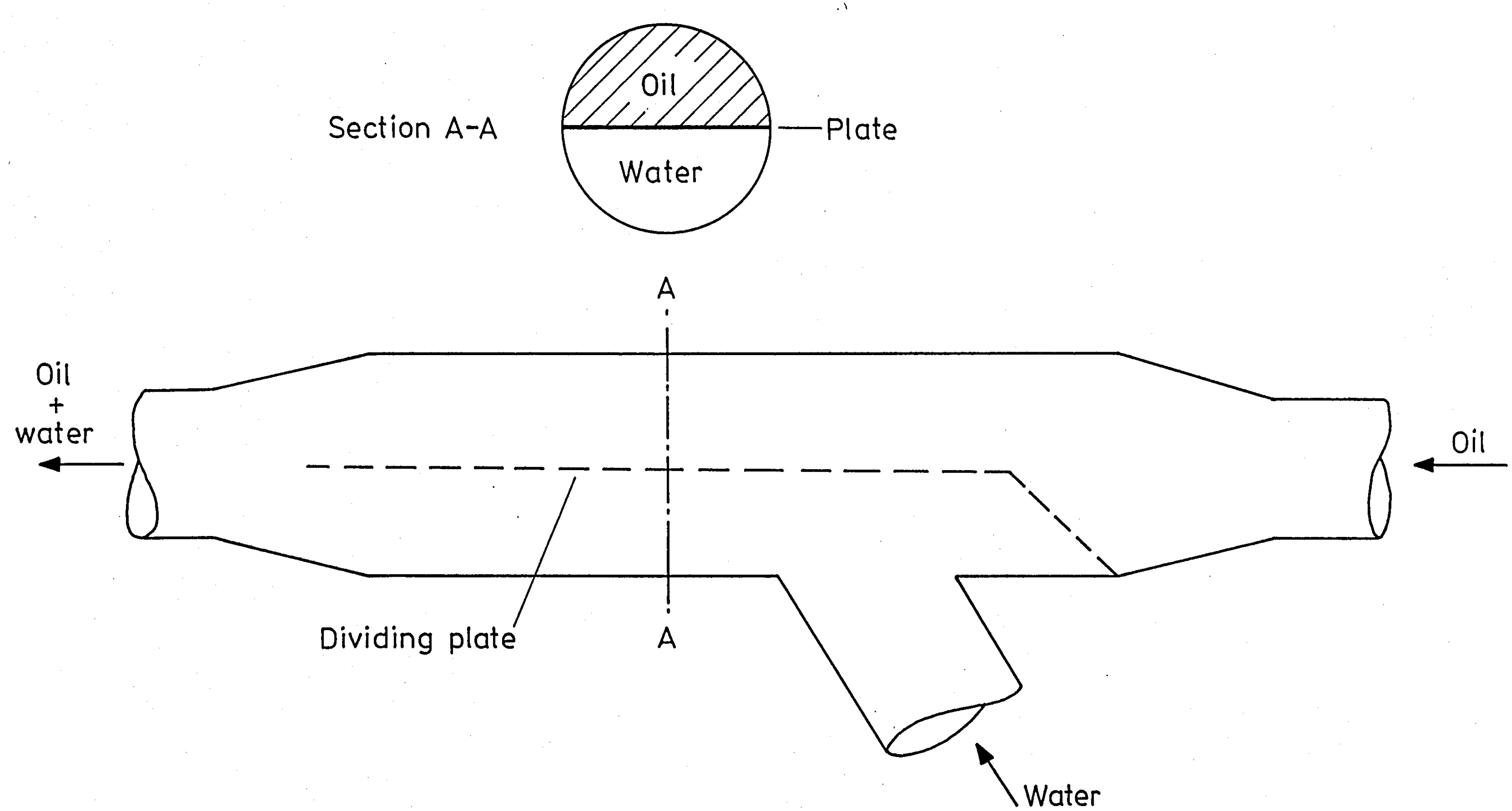


Fig 4.8 Oil–Water Mixer

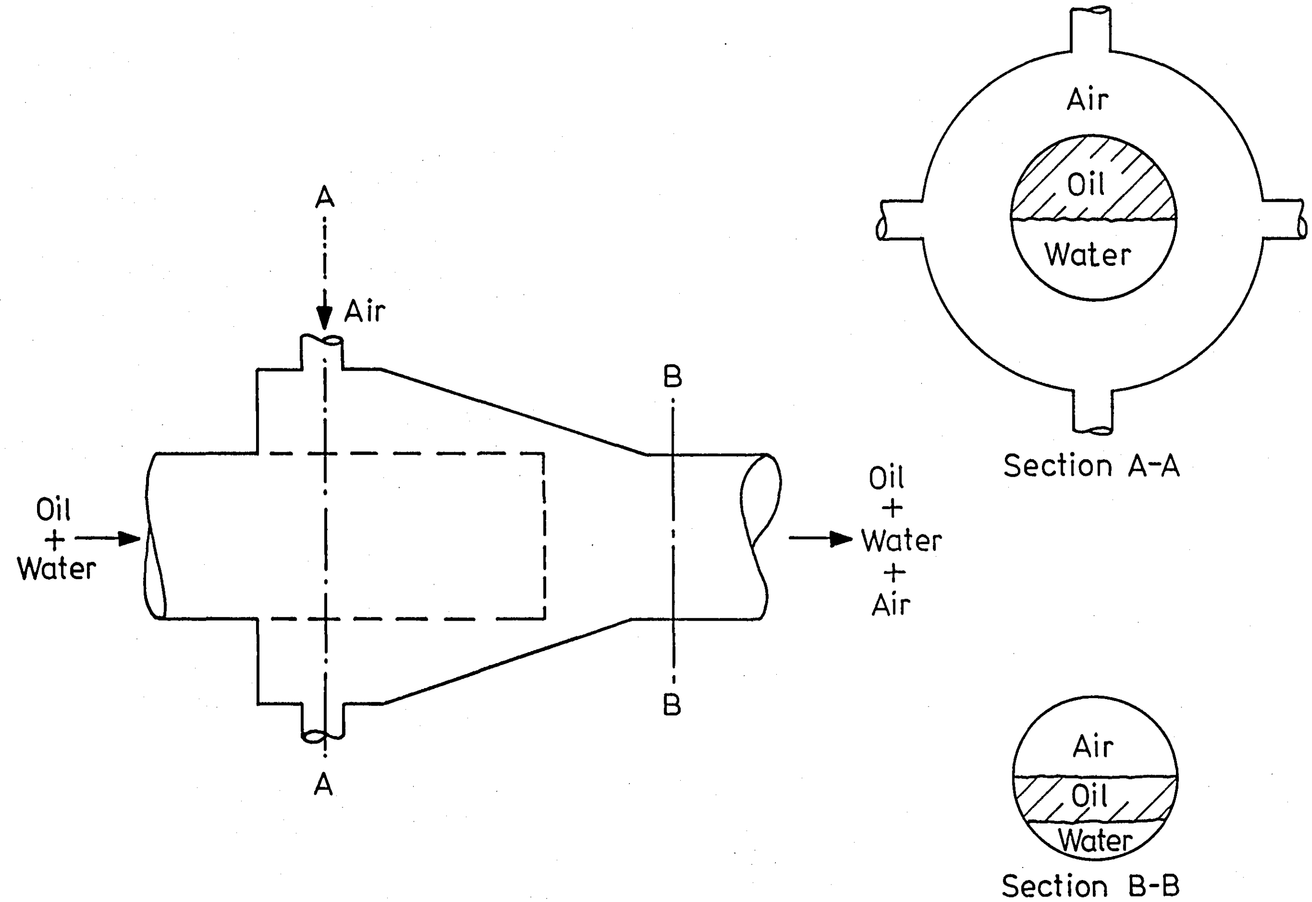


Fig 4.9 Gas/Liquids Mixer

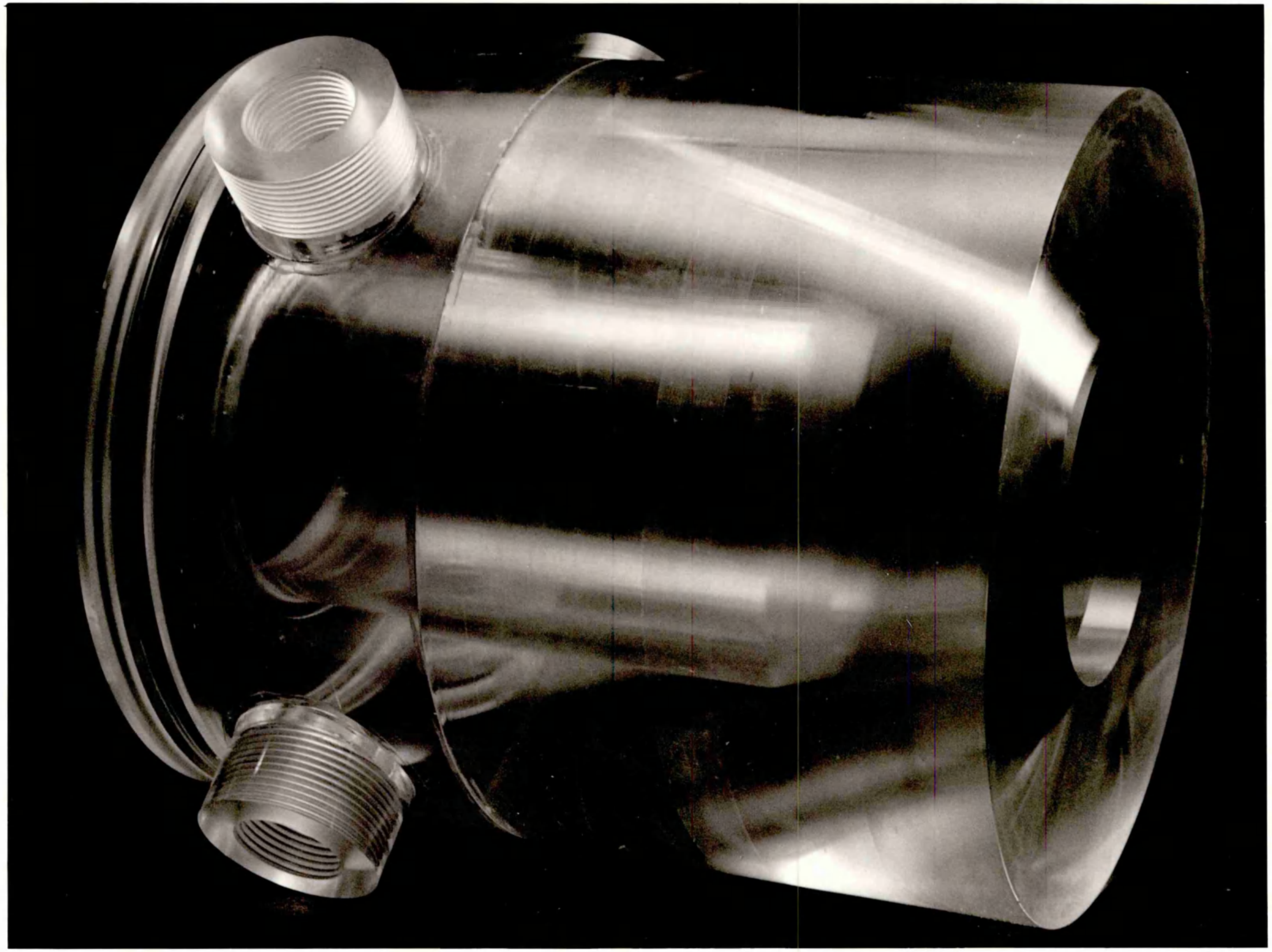


Fig 4.10 Gas/Liquids Mixer Before Installation

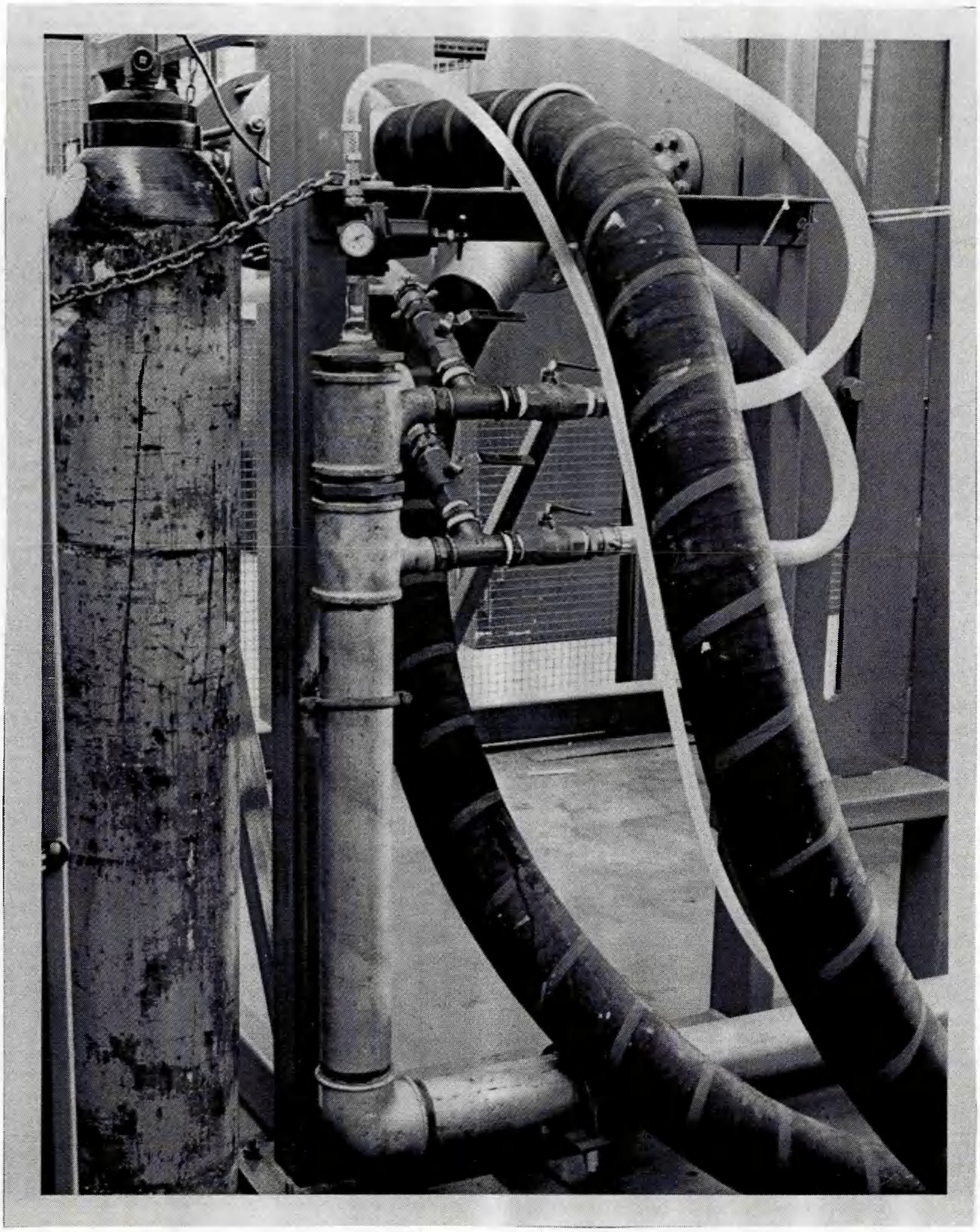


Fig 4.11 Gas Input Manifold

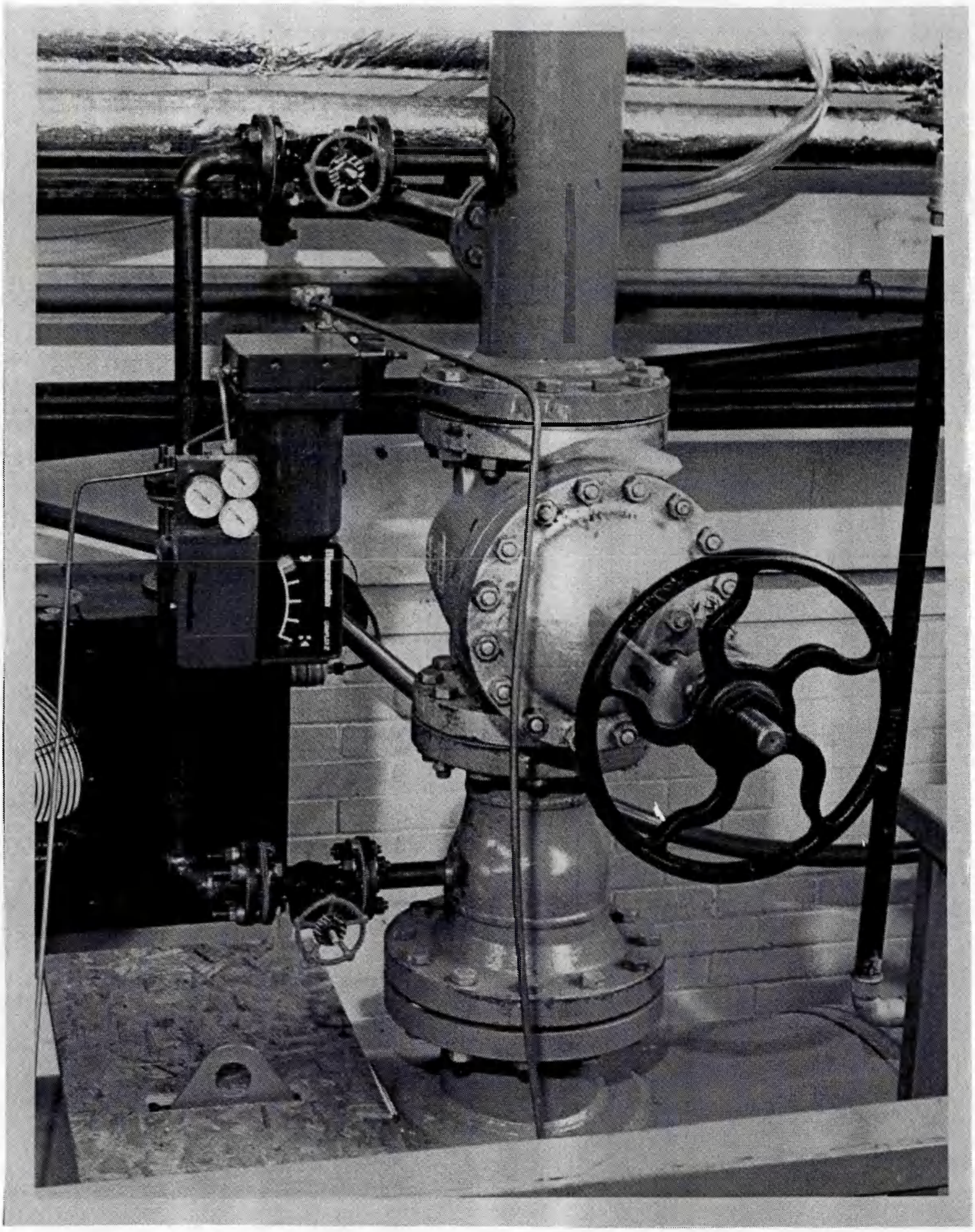


Fig 4.12 Separator Pressure Control Valve and Bypass

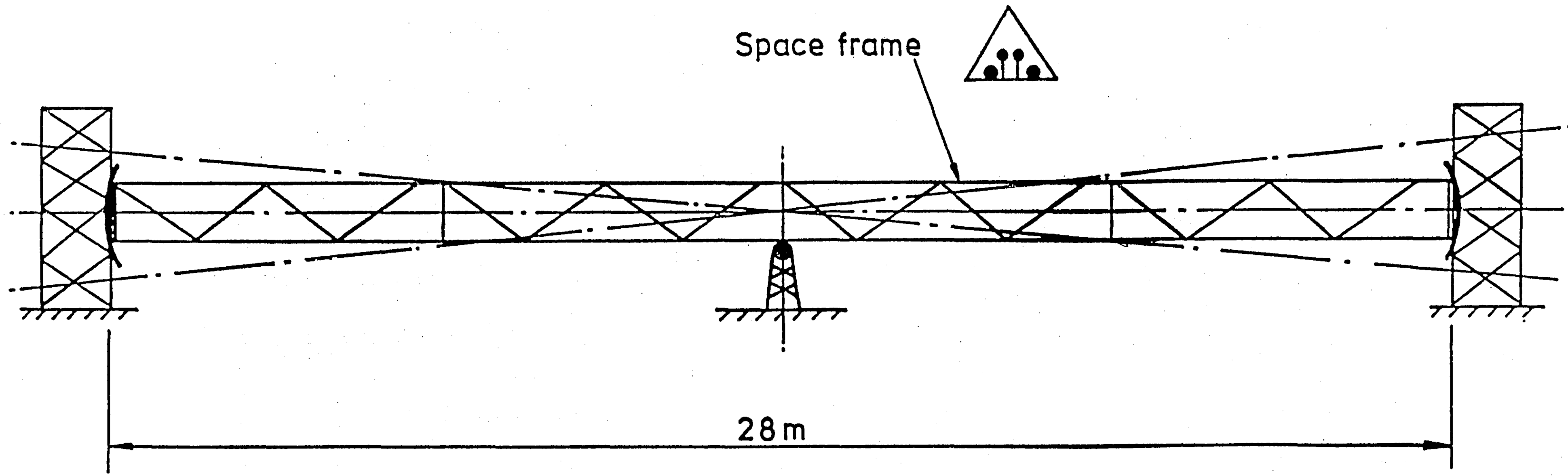


Fig 4.13 Pipe Bridge Structure

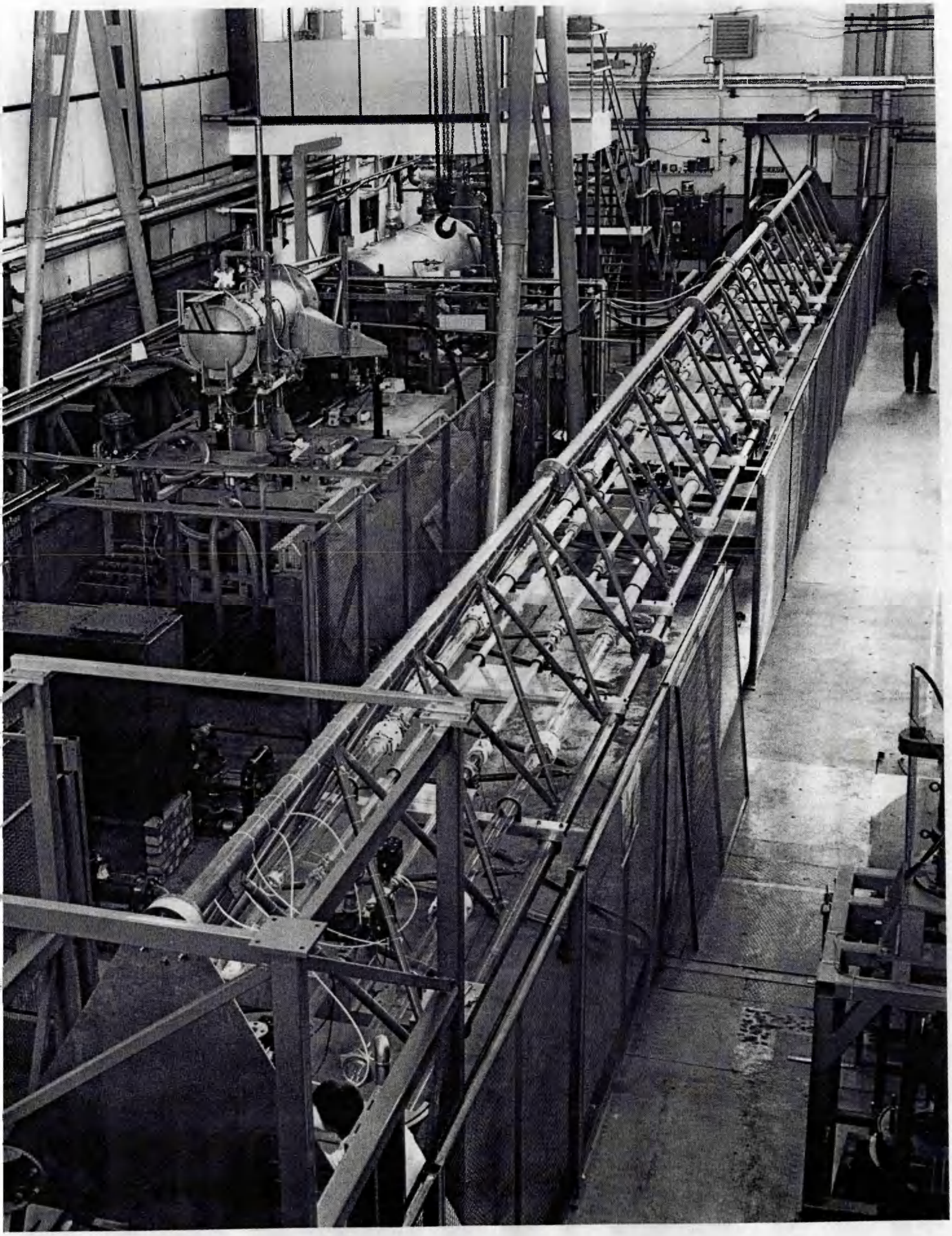


Fig 4.14 Pipe Bridge in Horizontal Position

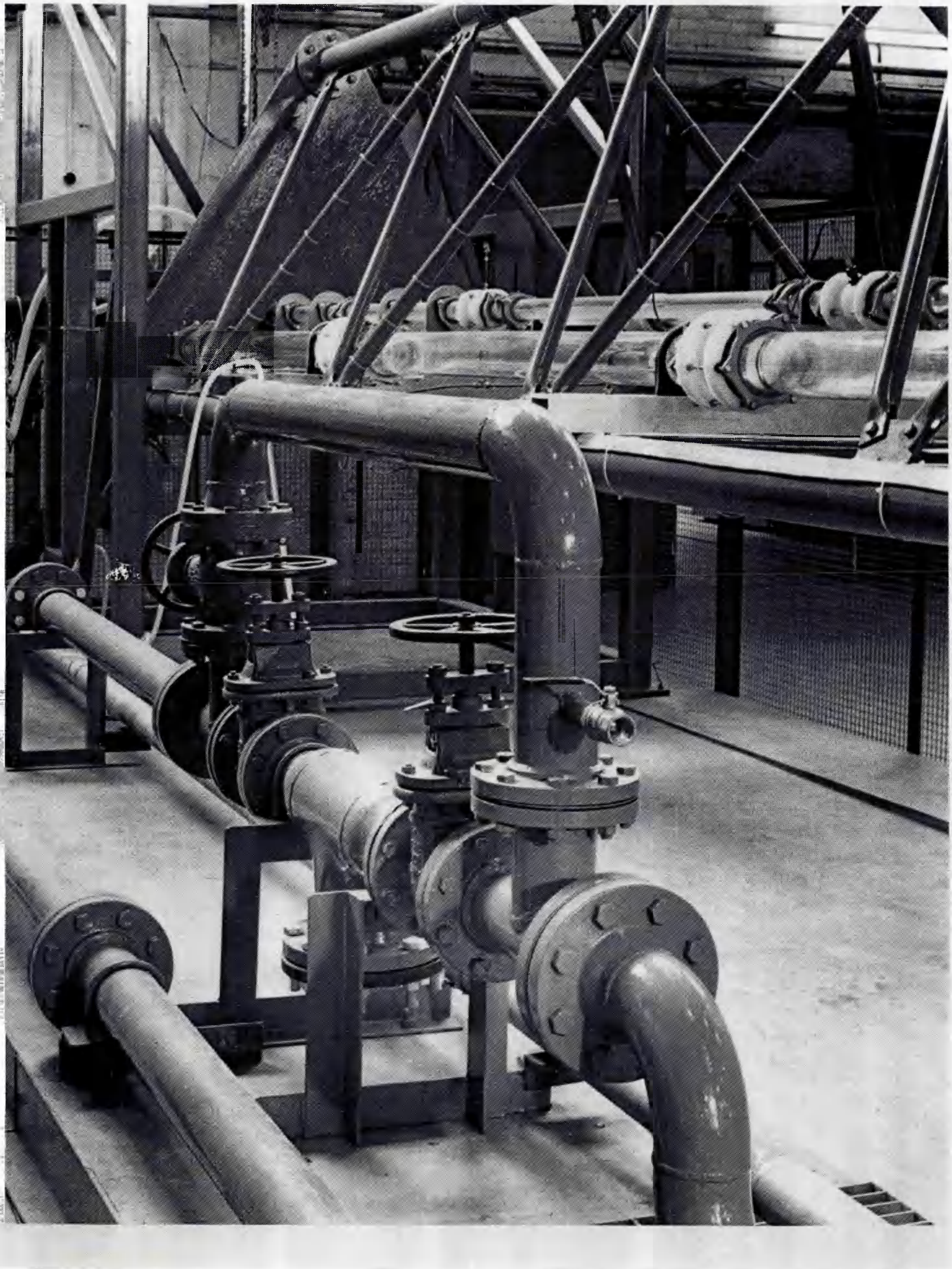


Fig 4.15 Pigging Bypass Section and Test Loops

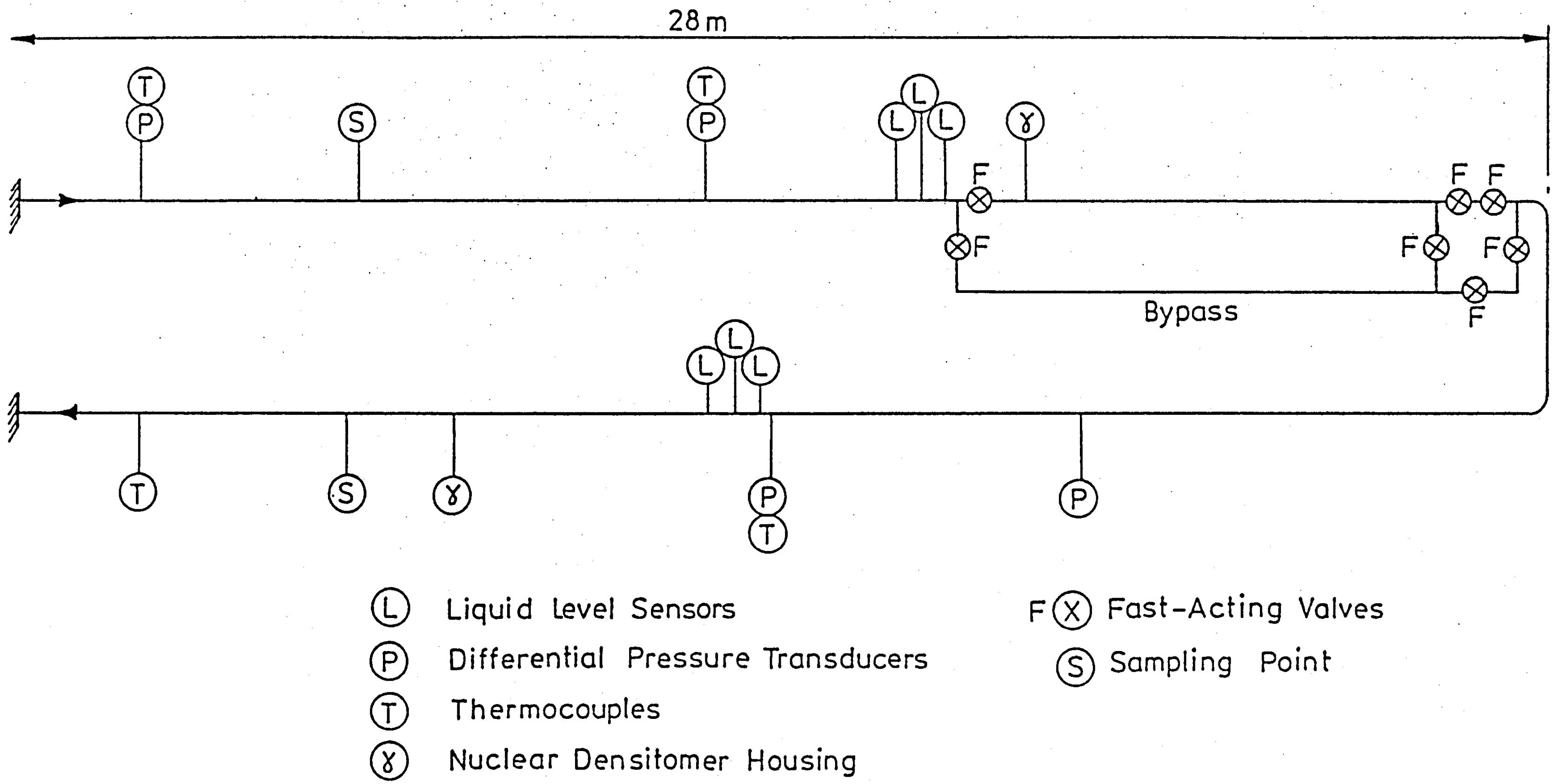


Fig 4.16 Location of Instrumentation on Test Loop

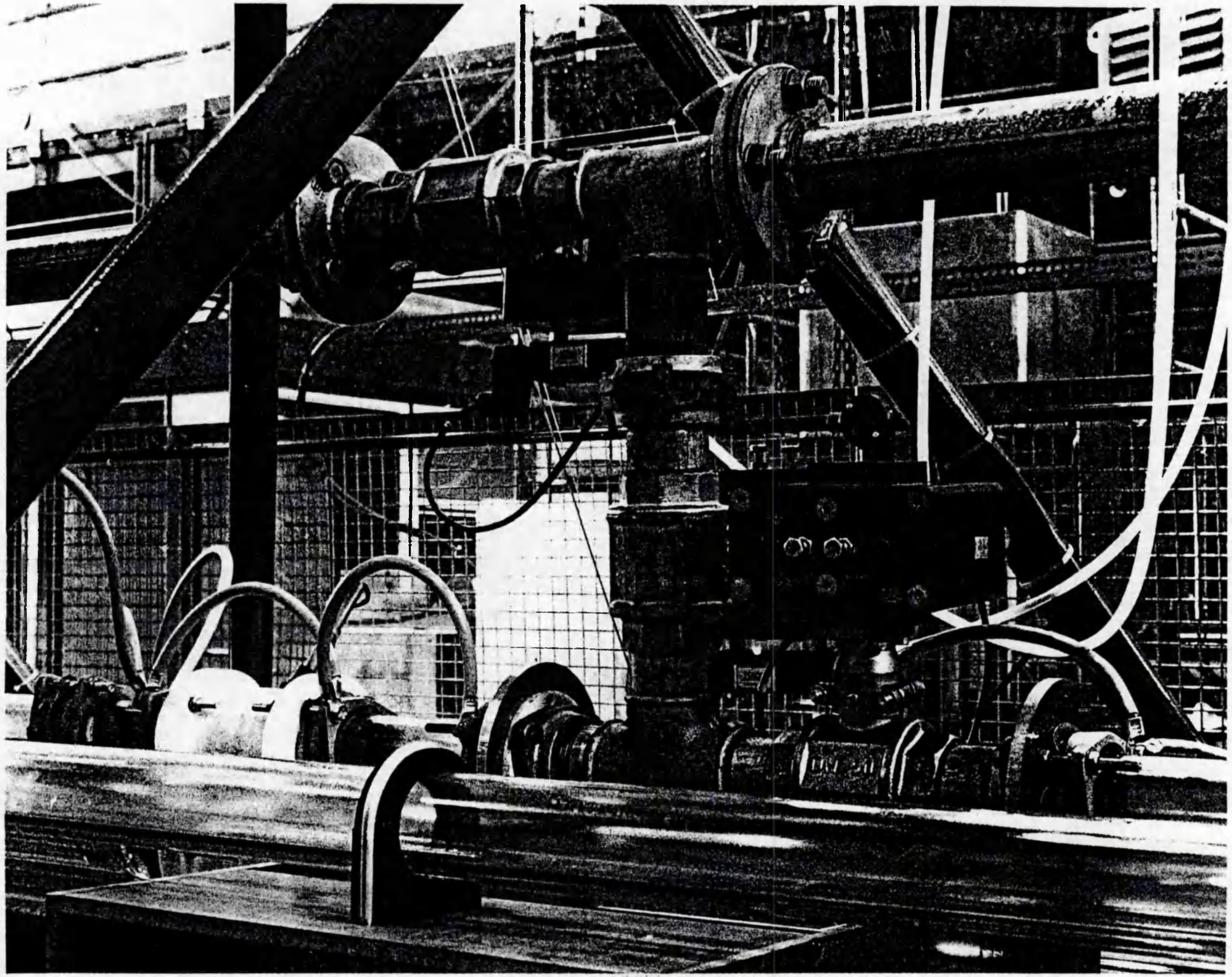


Fig 4.17 Tee Containing Pneumatic Fast–Acting Valves

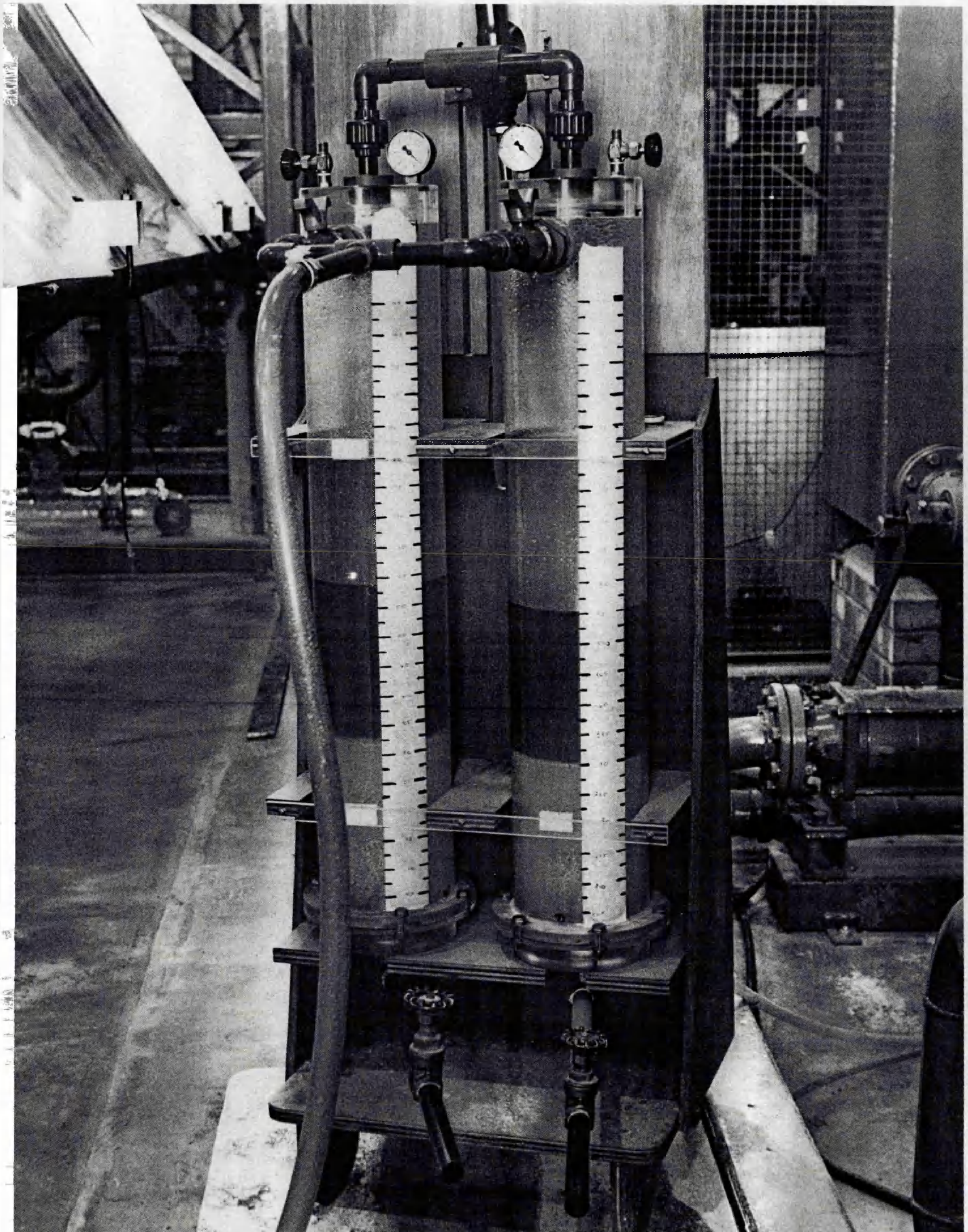


Fig 4.18 Liquids Holdup Drainage System

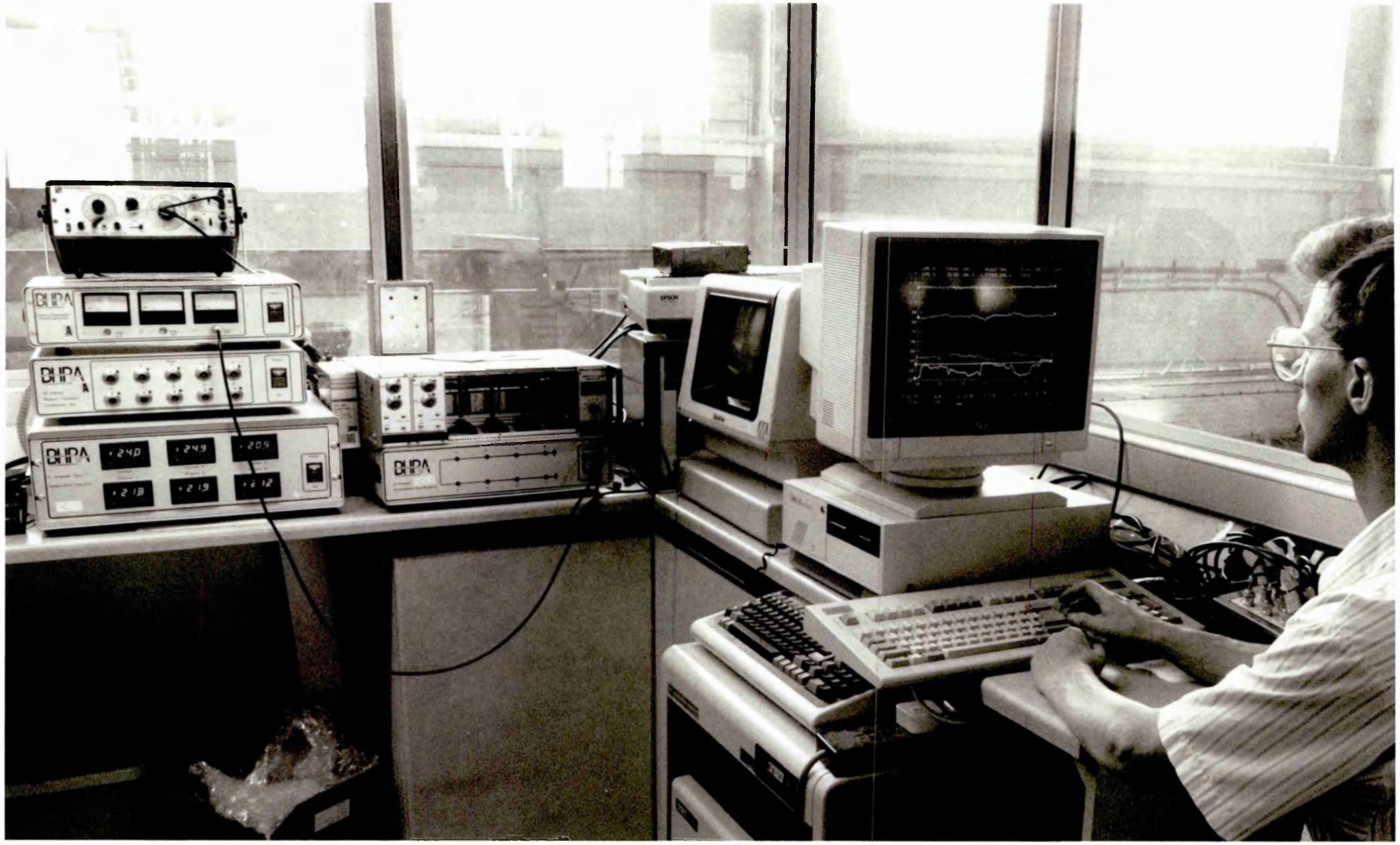


Fig 4.19 Facility Instrumentation and Control Cabin

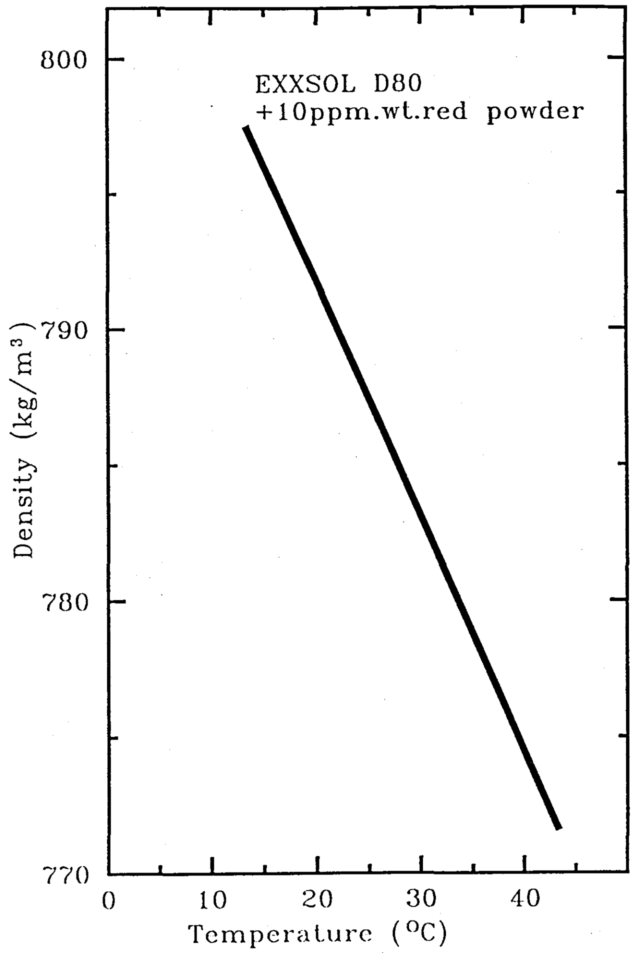


Fig 5.1 Oil No1 Density-Temperature Curve

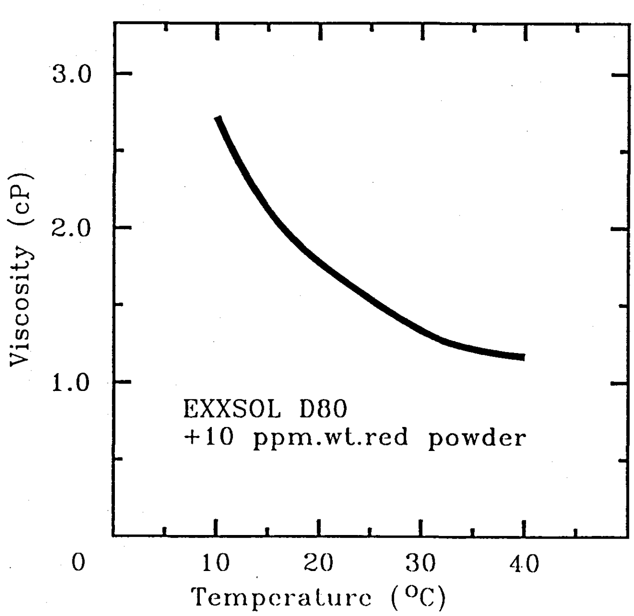


Fig 5.2 Oil No1 Viscosity-Temperature Curve

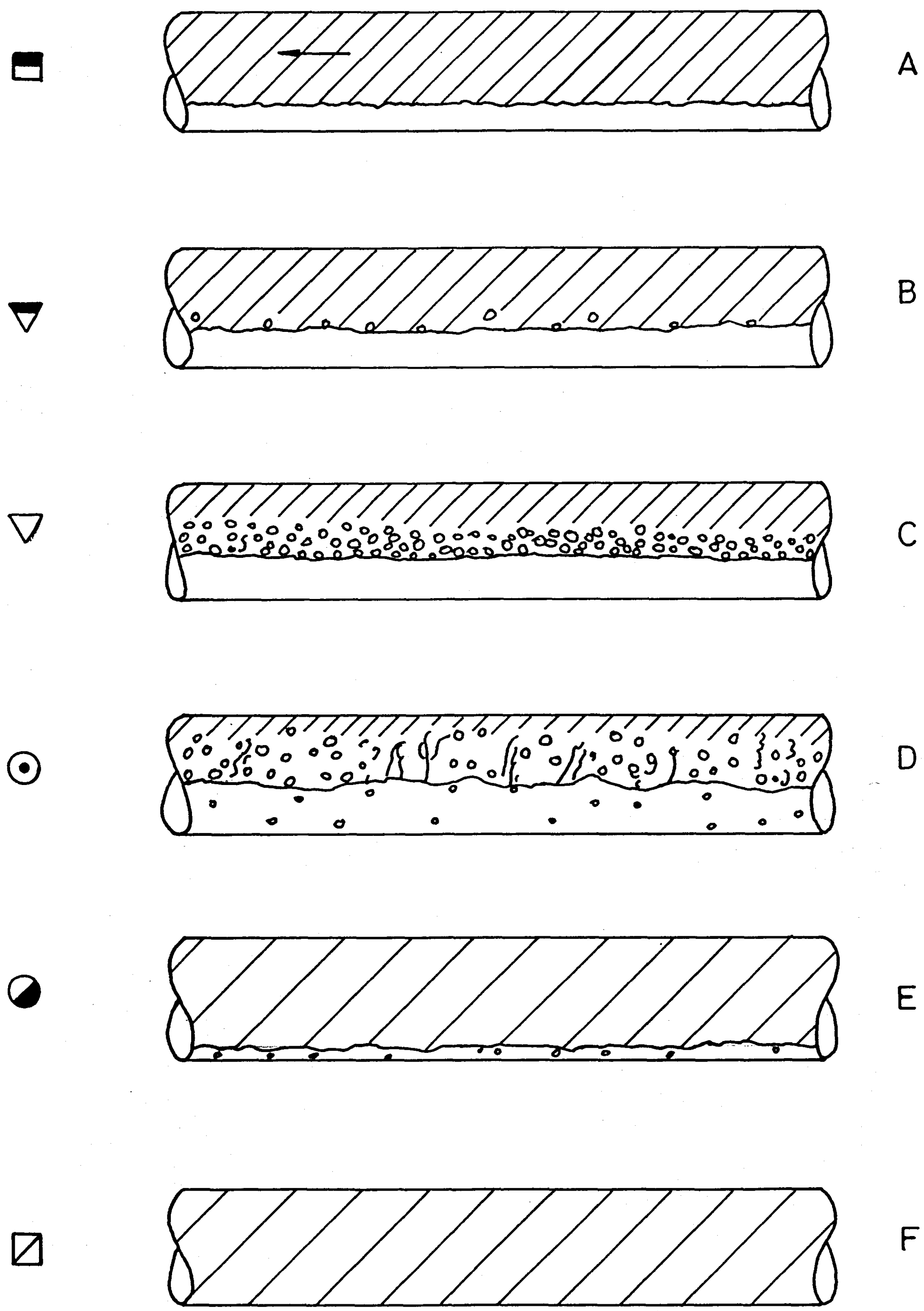


Fig 5.3 Oil-Water Flow Pattern Classifications

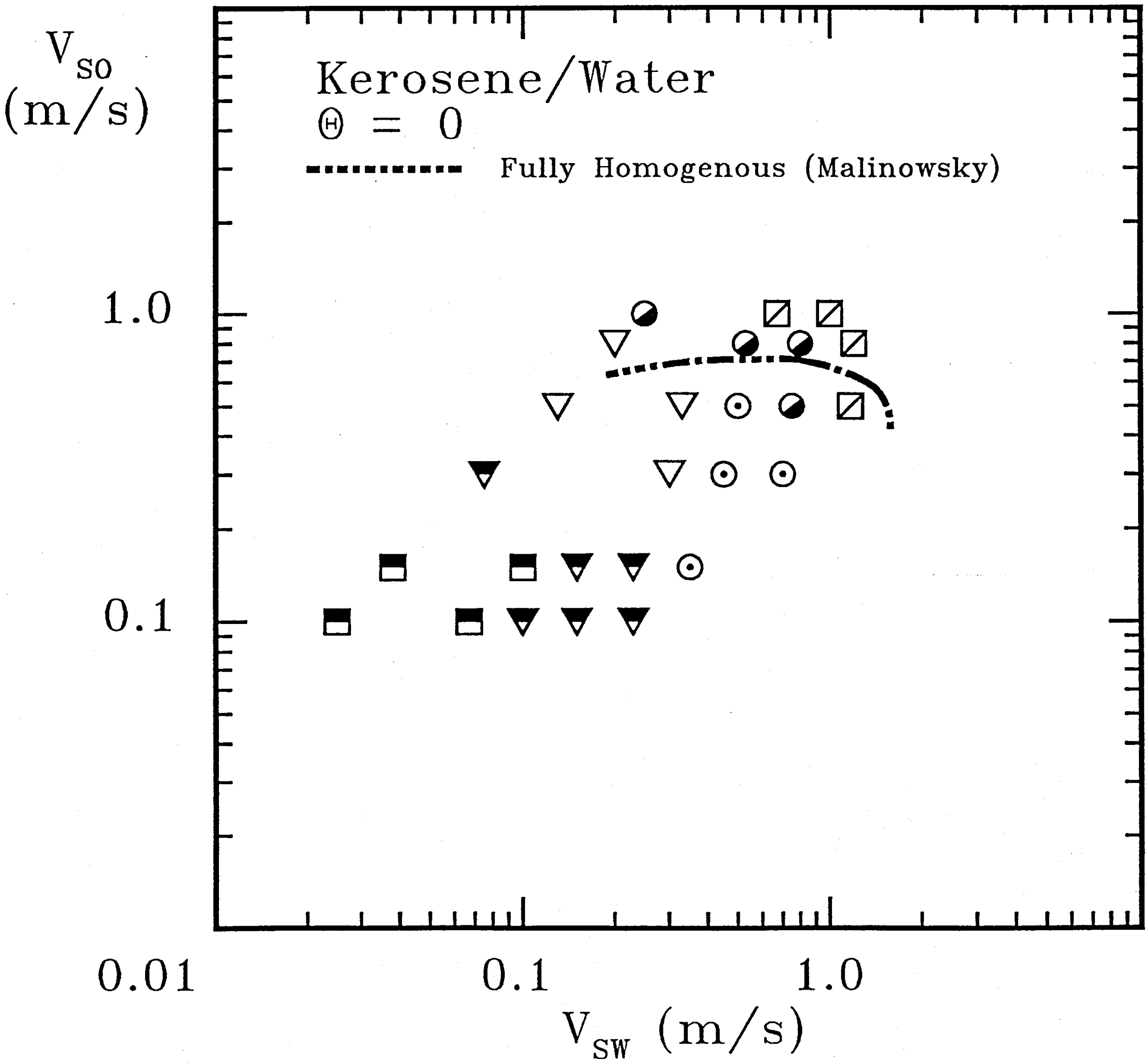


Fig 5.4 Oil No1 Oil–Water Flow Regime Map

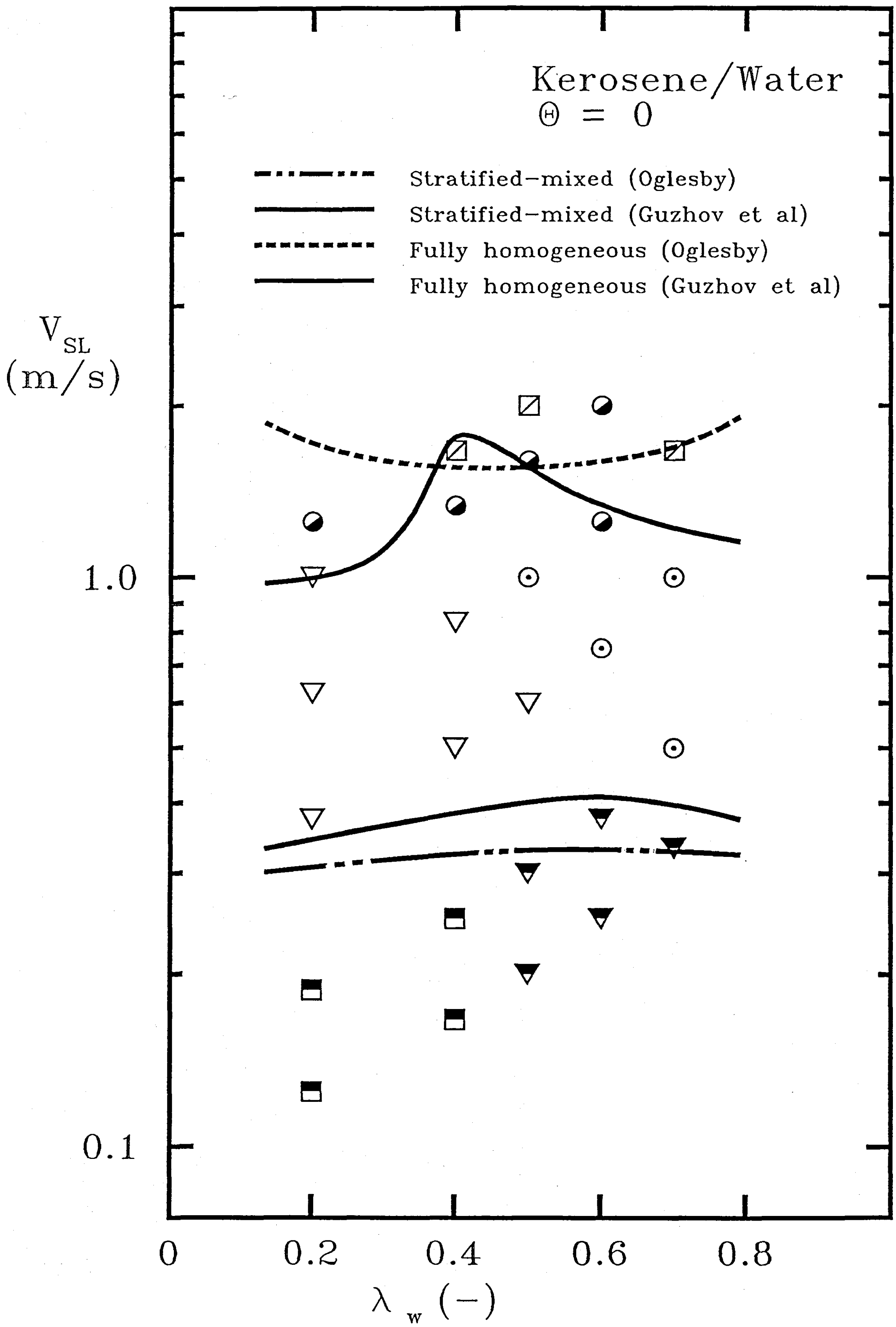


Fig 5.5 Oil No1 Oil-Water Flow Regime Map

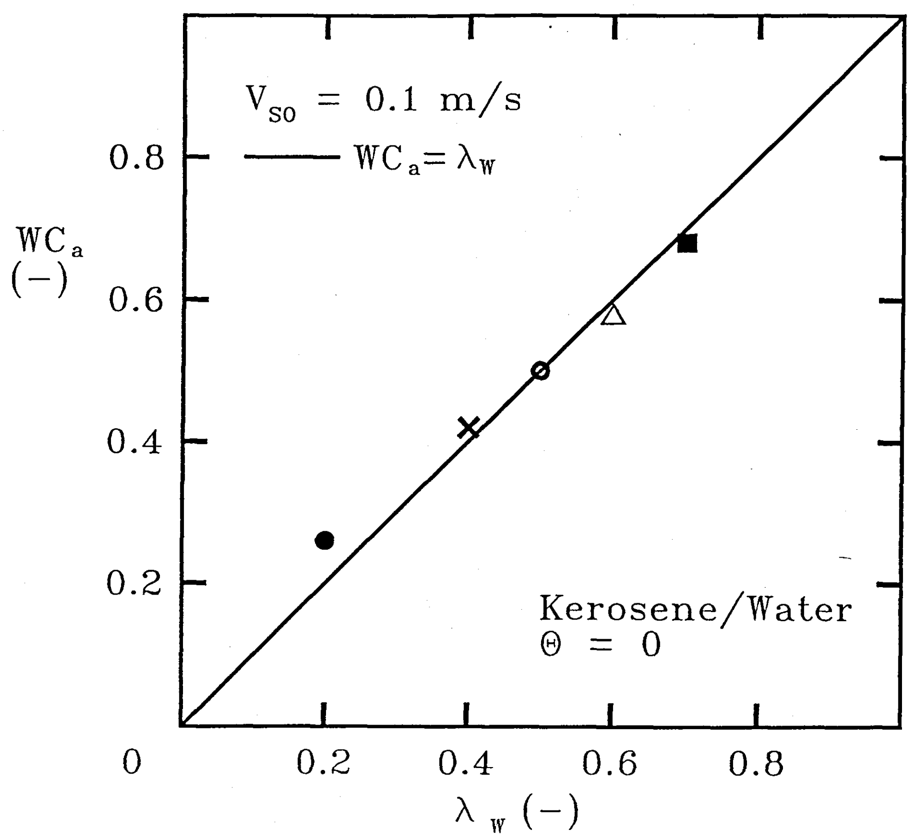


Fig 5.6 Oil−Water In−situ Water Fractions,Oil No1

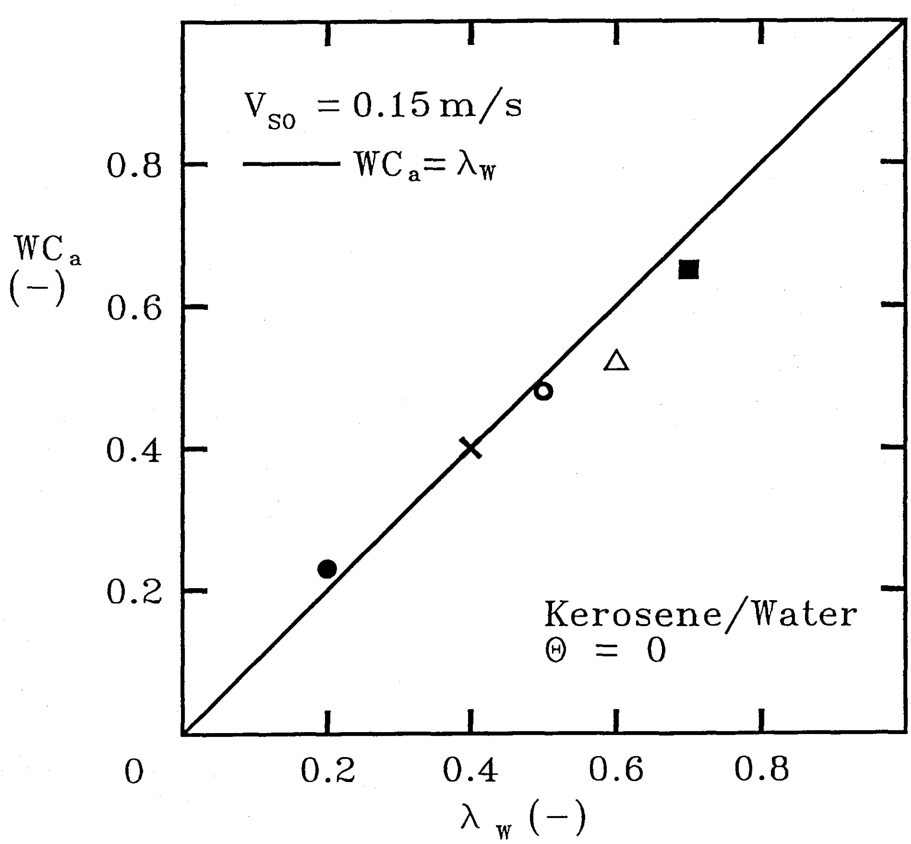


Fig 5.7 Oil−Water In−situ Water Fractions,Oil No1

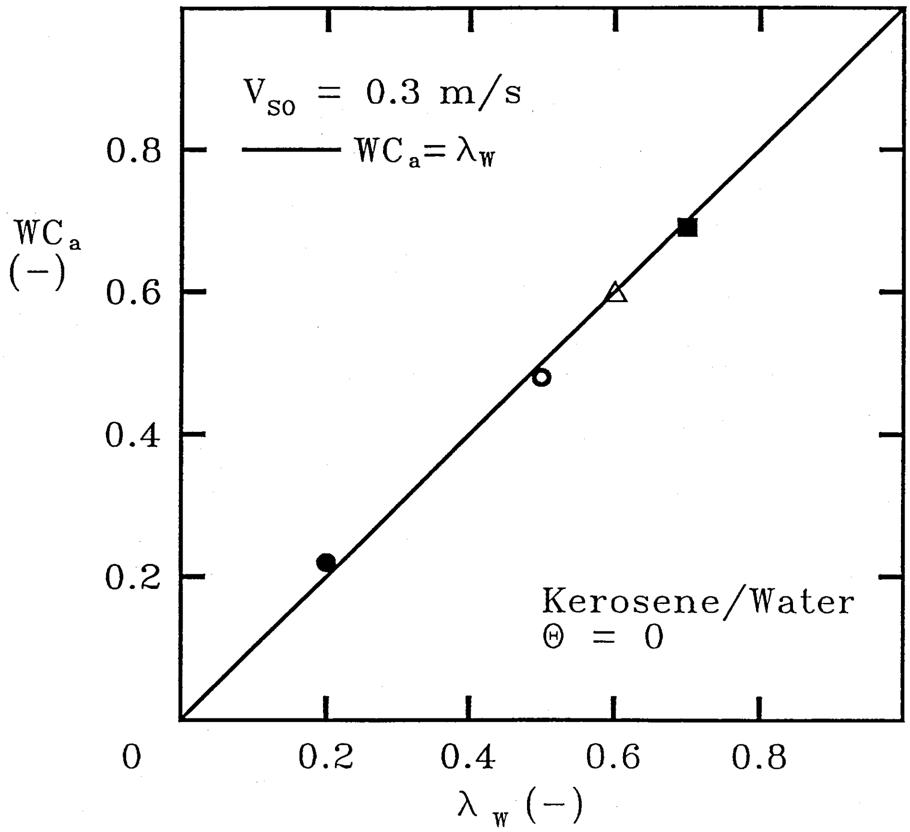


Fig 5.8 Oil–Water In–situ Water Fractions,Oil No1

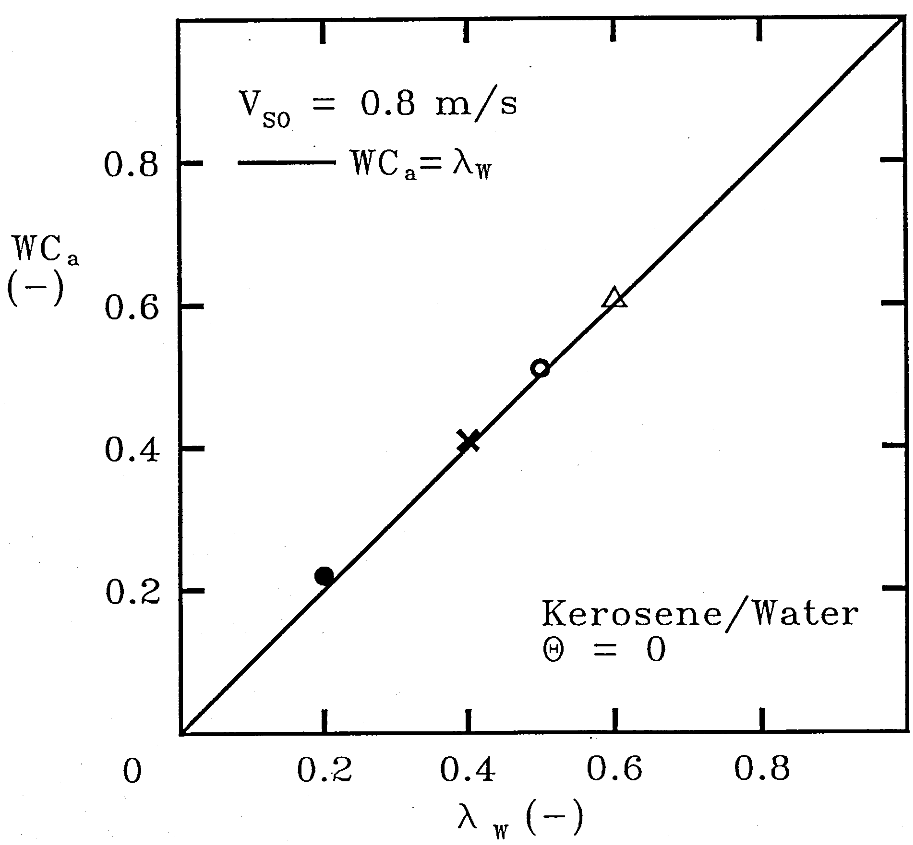


Fig 5.9 Oil–Water In–situ Water Fractions,Oil No1

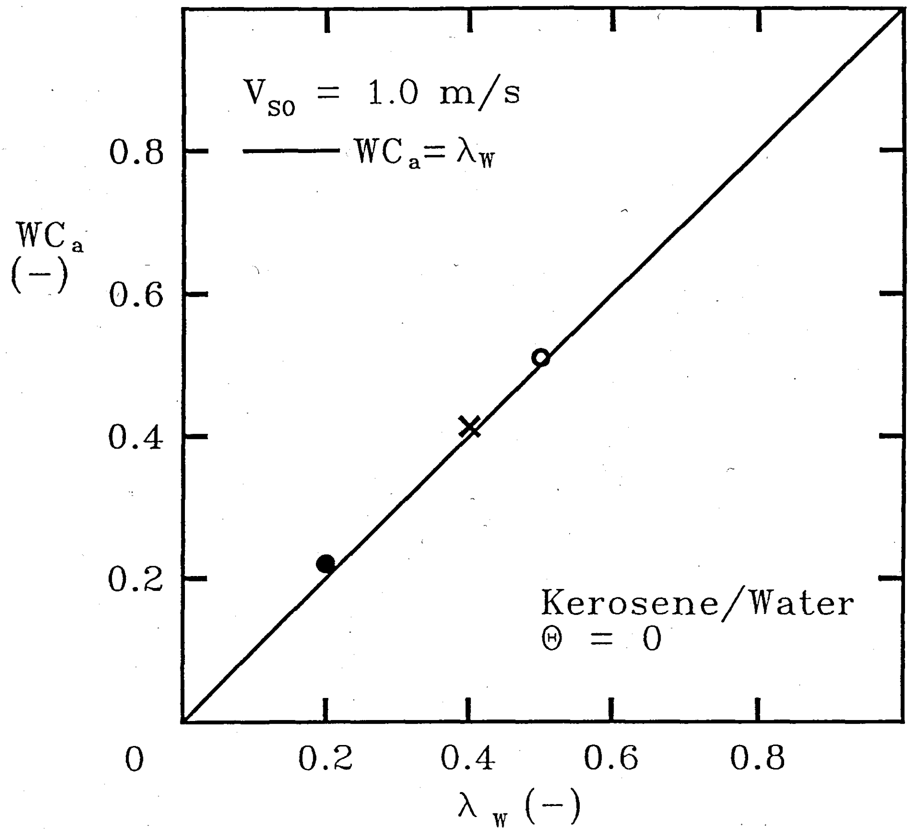


Fig 5.10 Oil−Water In−situ Water Fractions,Oil No1

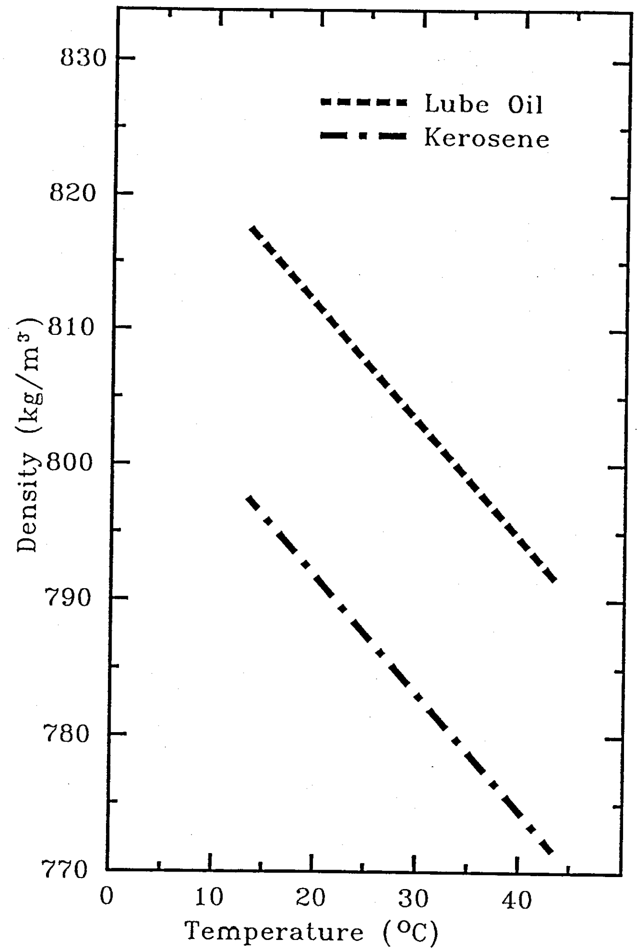


Fig 5.11 Oil No2 Density–Temperature Curve

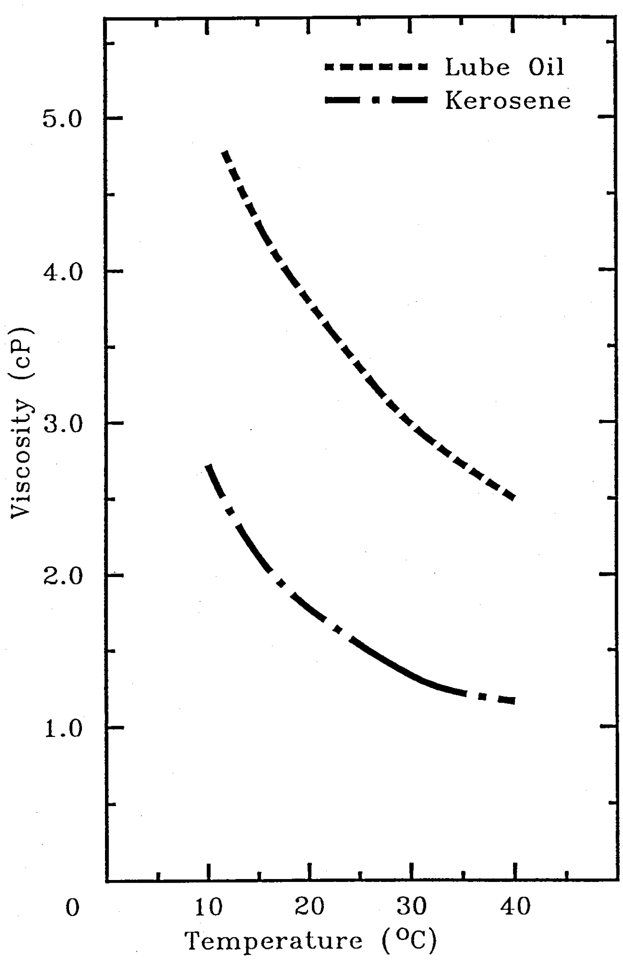


Fig 5.12 Oil No2 Viscosity–Temperature Curve

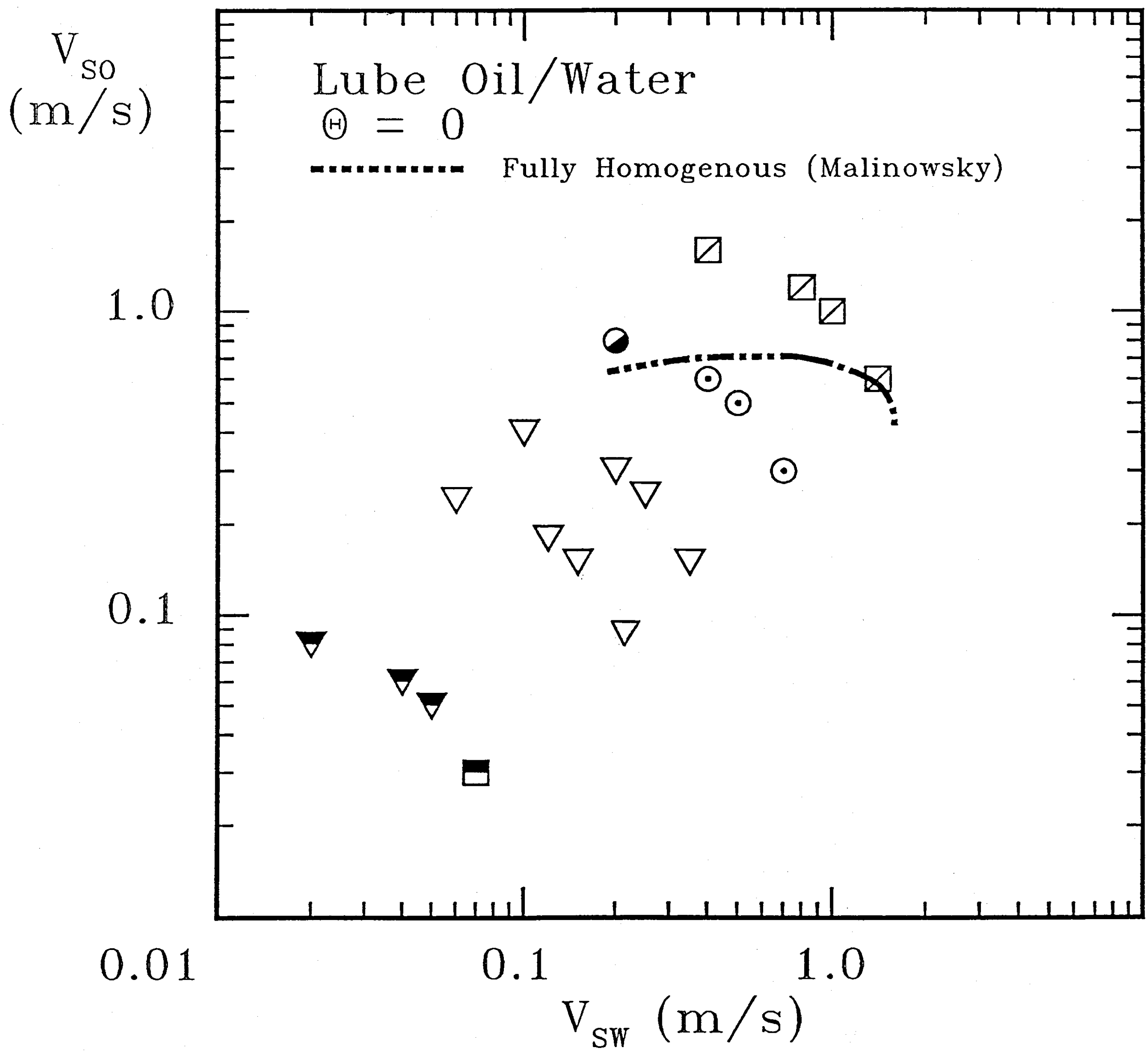


Fig 5.13 Oil No2 Oil–Water Flow Regime Map

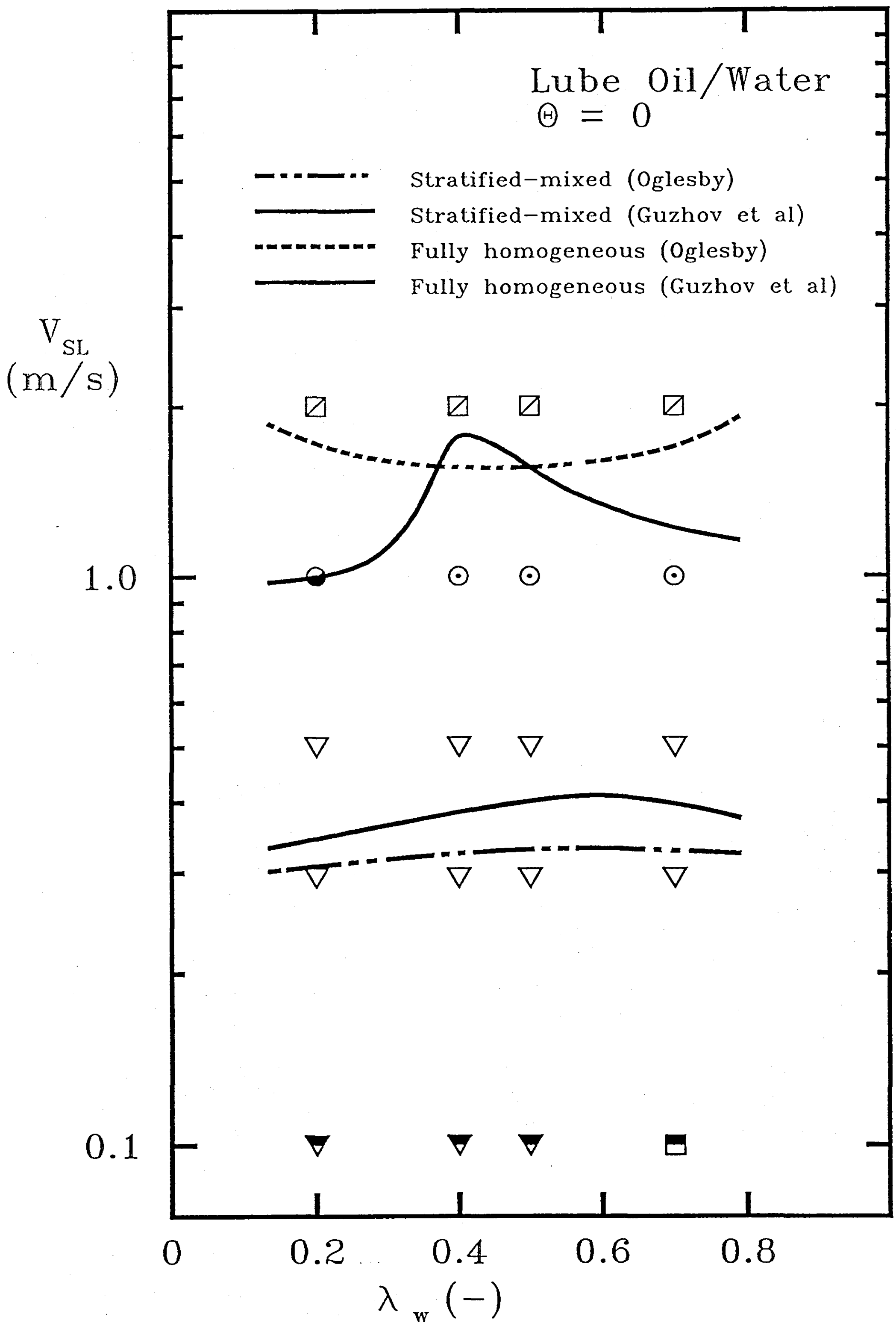
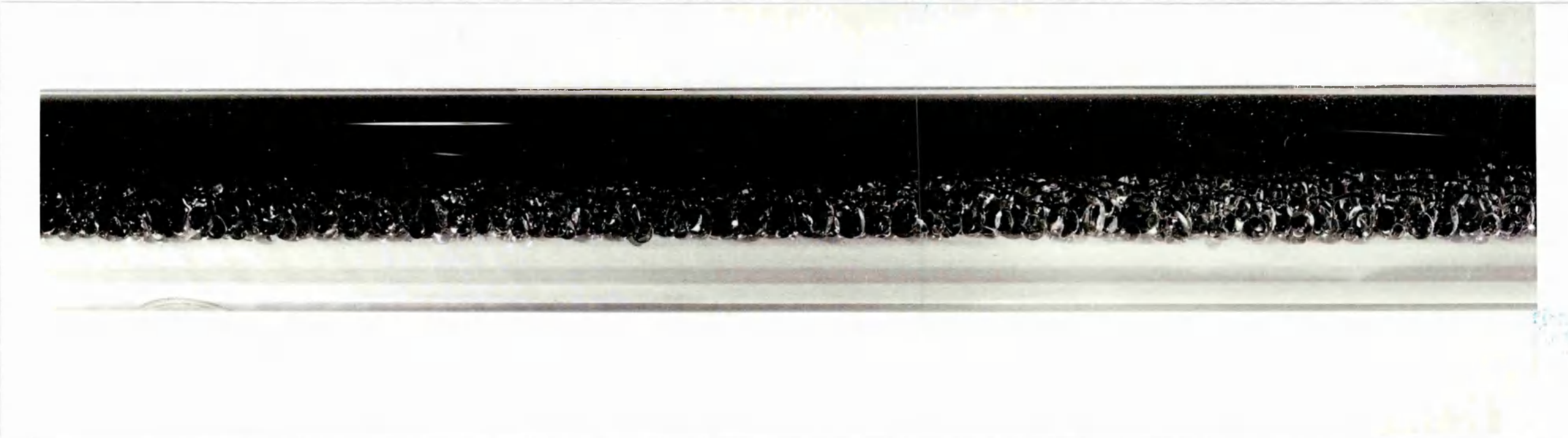
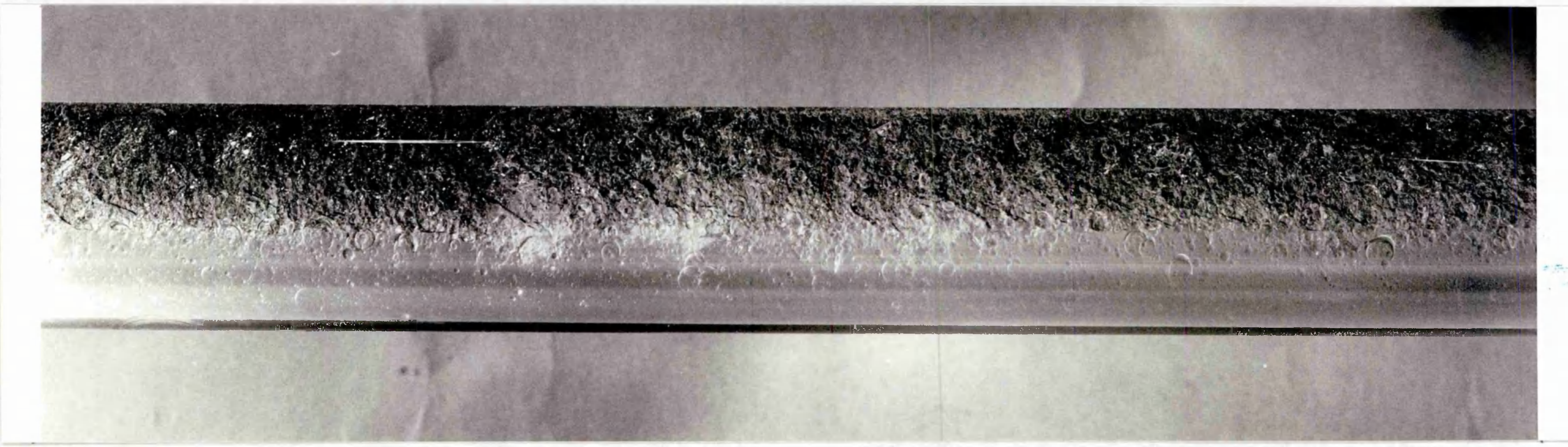


Fig 5.14 Oil No2 Oil–Water Flow Regime Map

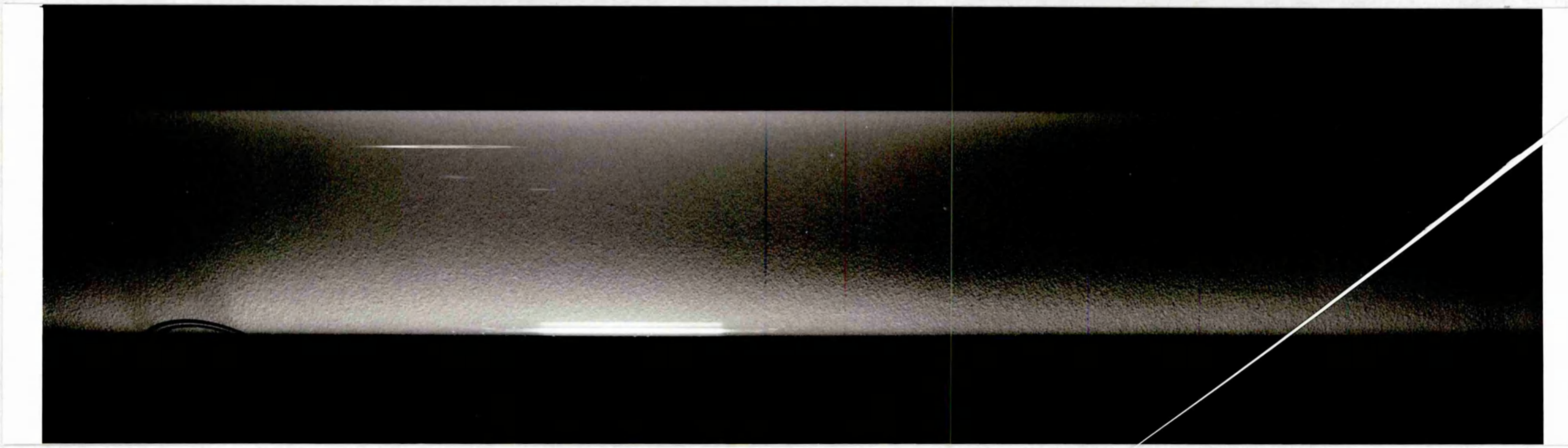


(a) $V_{SL} = 0.3$ m/s, $\lambda_W = 0.4$ (Flow R to L)



(b) $V_{SL} = 1.0$ m/s, $\lambda_W = 0.7$ (Flow R to L)

Fig 5.15 Oil–Water Flow Pattern Photographs at L = 1000d ,Oil No2



(c) $V_{SL} = 2.0$ m/s, $\lambda_w = 0.4$ (Flow R to L)

Fig 5.15 Oil-Water Flow Pattern Photographs at L = 1000d ,Oil No2

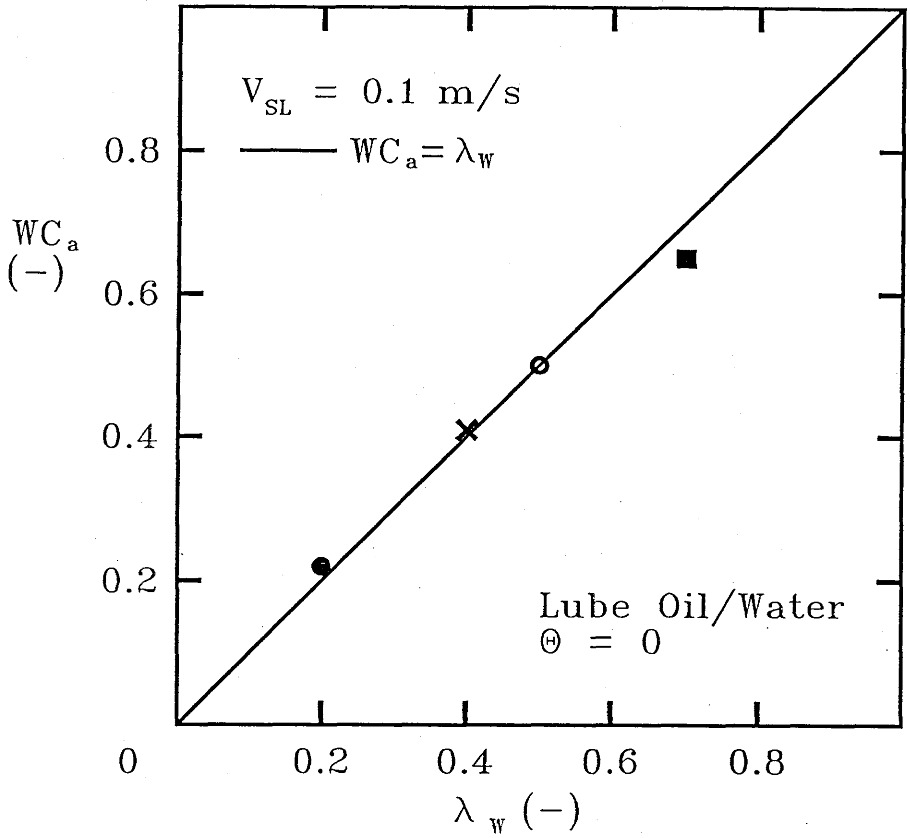


Fig 5.16 Oil−Water In−situ Water Fractions,Oil No2

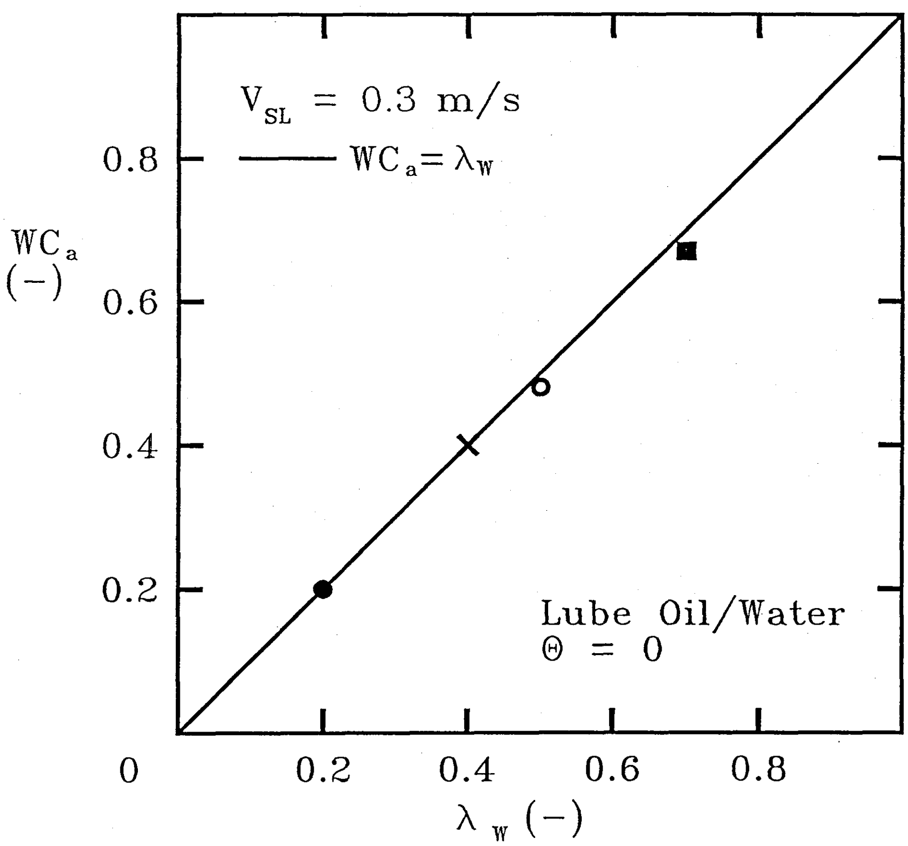


Fig 5.17 Oil−Water In−situ Water Fractions,Oil No2

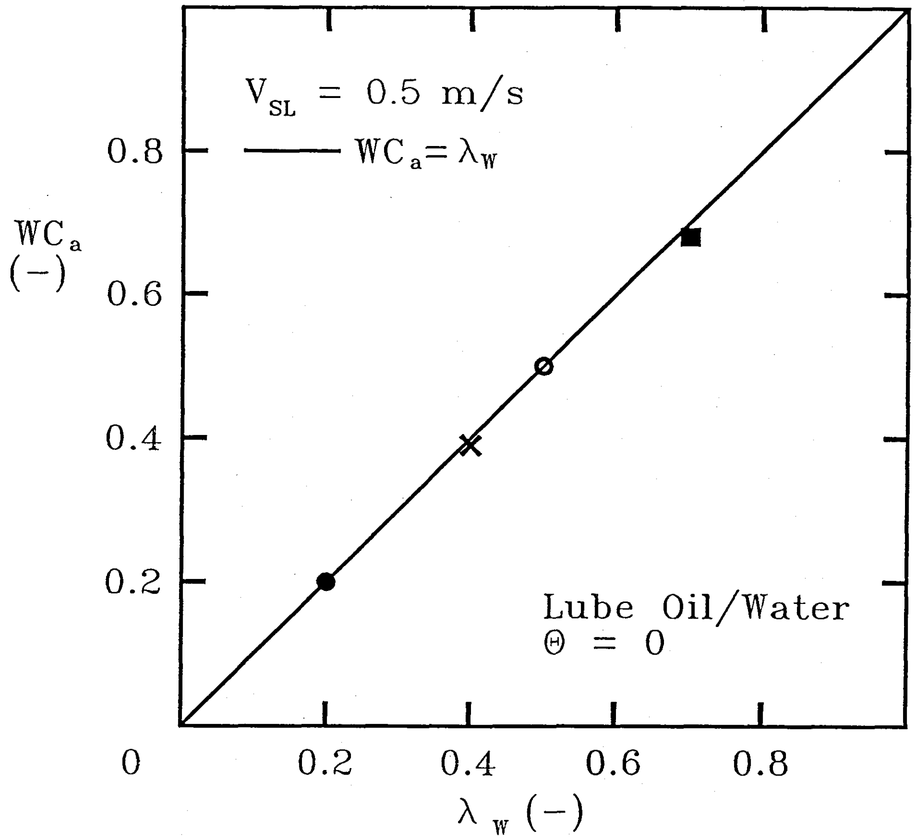


Fig 5.18 Oil−Water In−situ Water Fractions,Oil No2

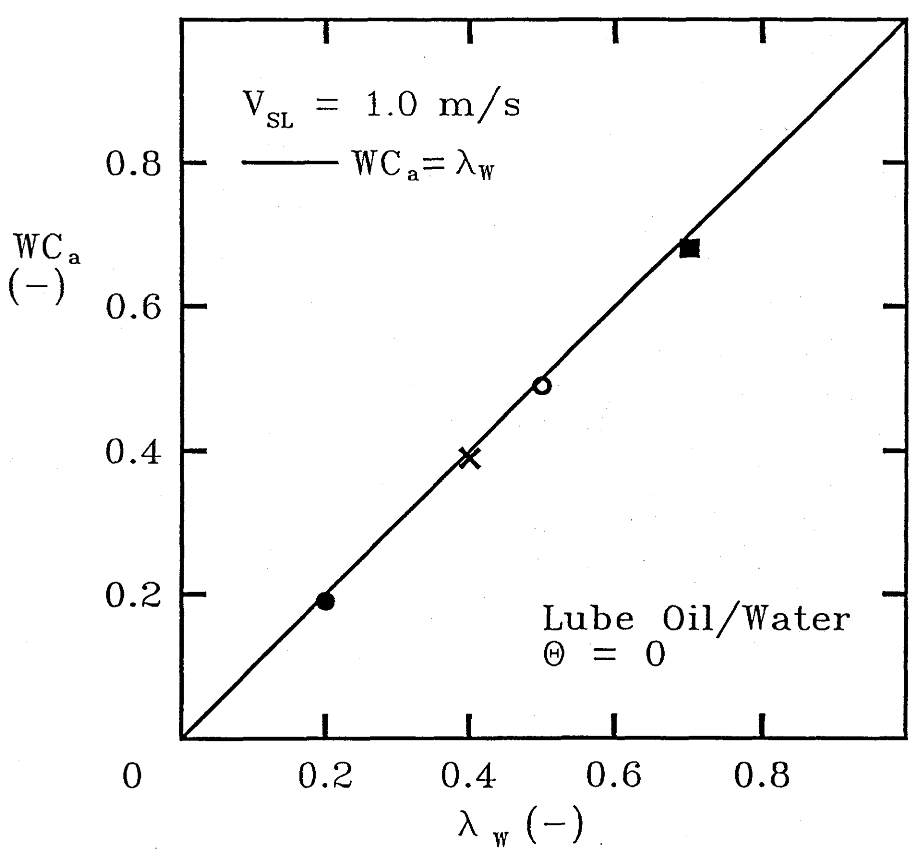


Fig 5.19 Oil−Water In−situ Water Fractions,Oil No2

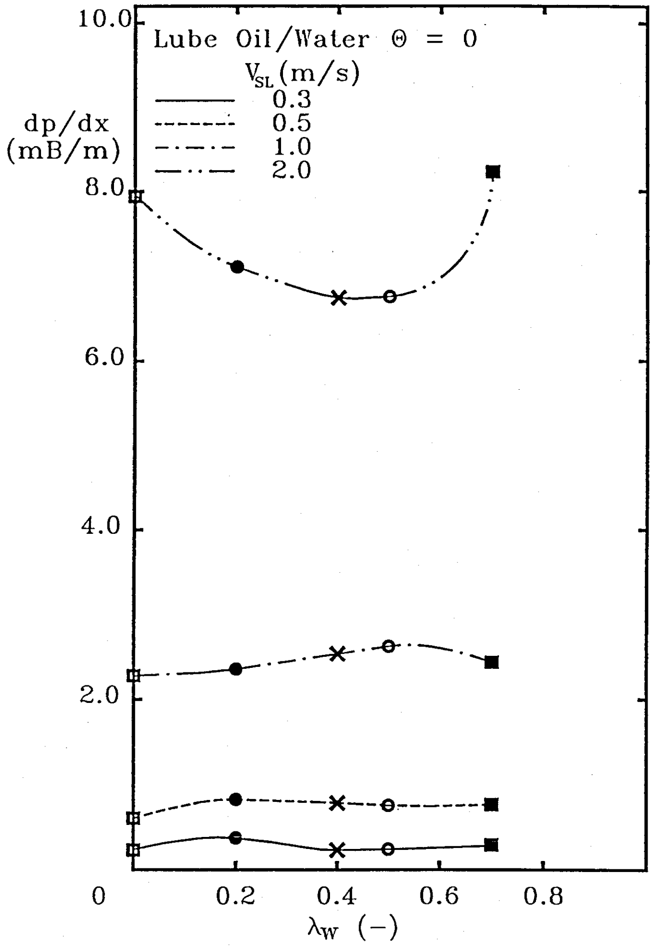


Fig 5.20 Oil−Water Pressure Loss,Oil No2

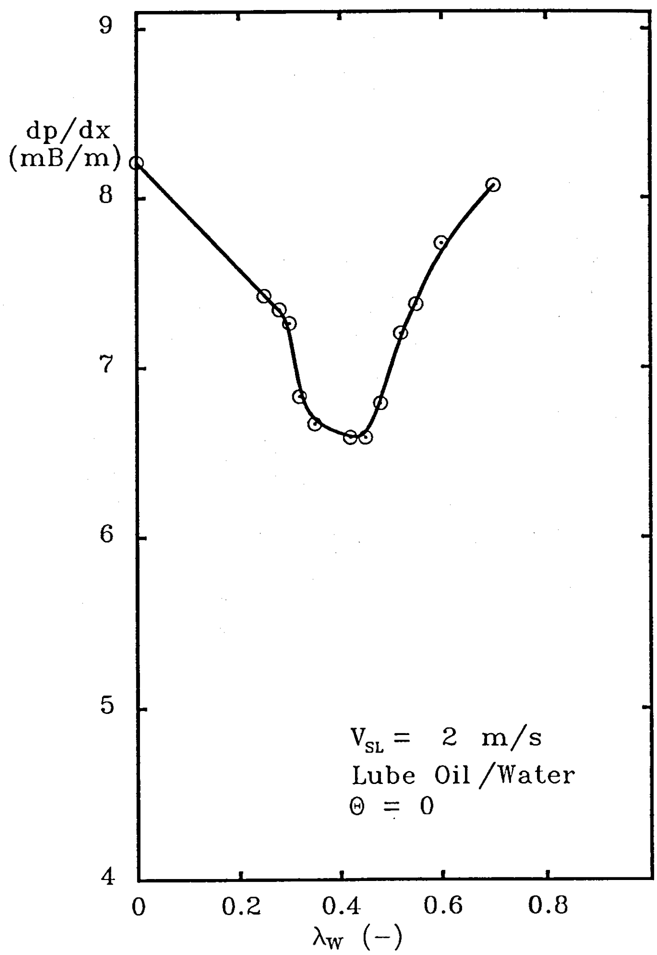


Fig 5.21 Oil−Water Pressure Loss,Oil No2

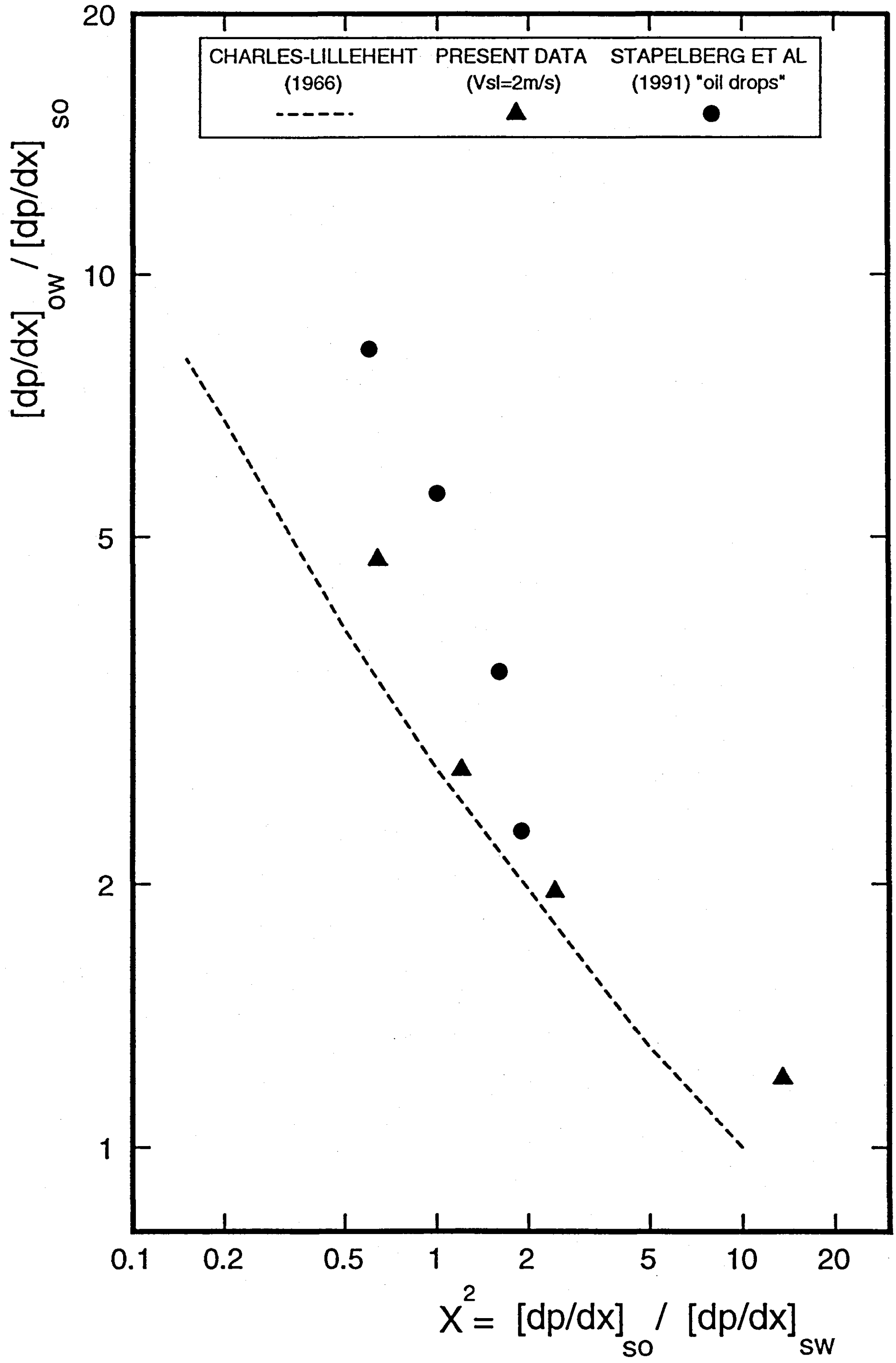


Fig 5.22 Oil–Water Pressure Loss Compared to Charles–Lilleheht (1966),Oil No2

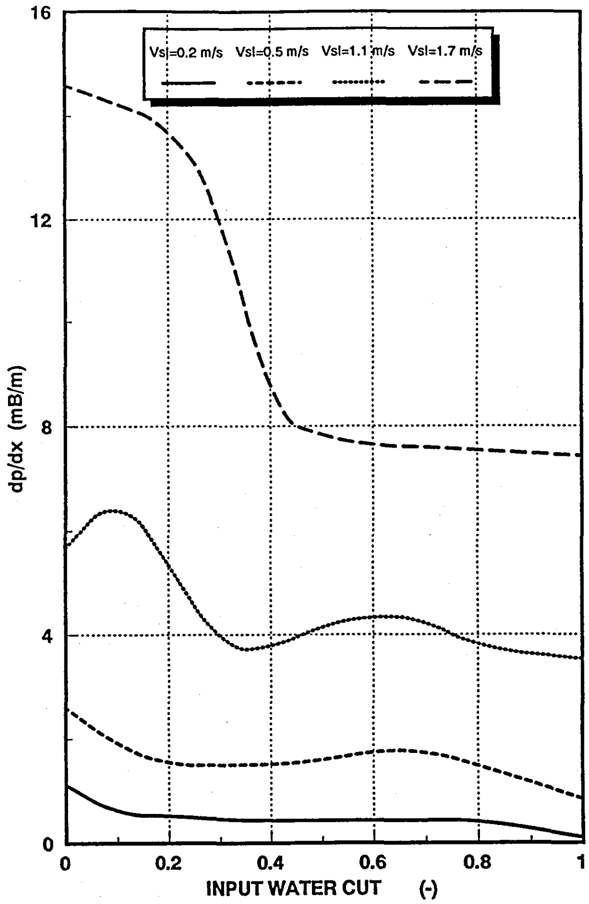


Fig 5.23 Oil–Water Pressure Loss Data of Guzhov et al (1973)

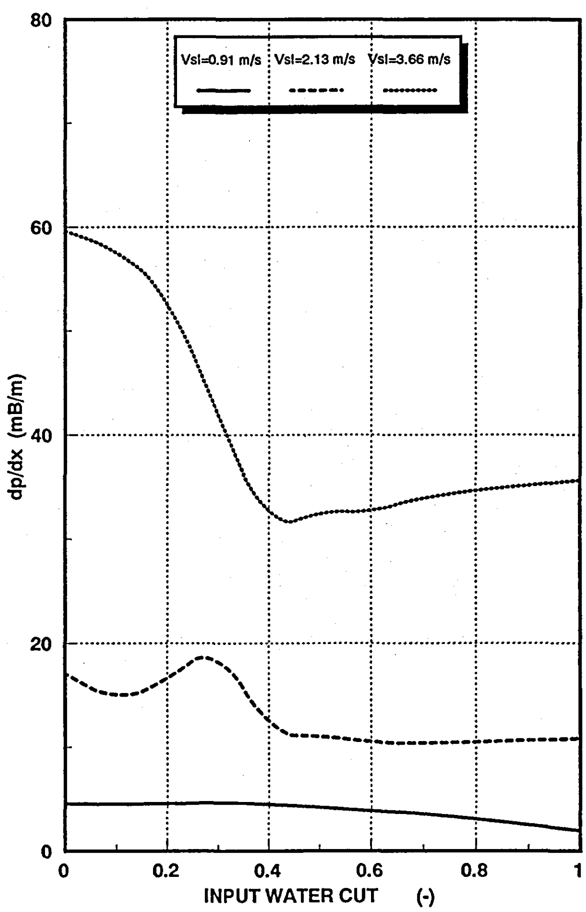


Fig 5.24 Oil–Water Pressure Loss Data of Oglesby (1979)

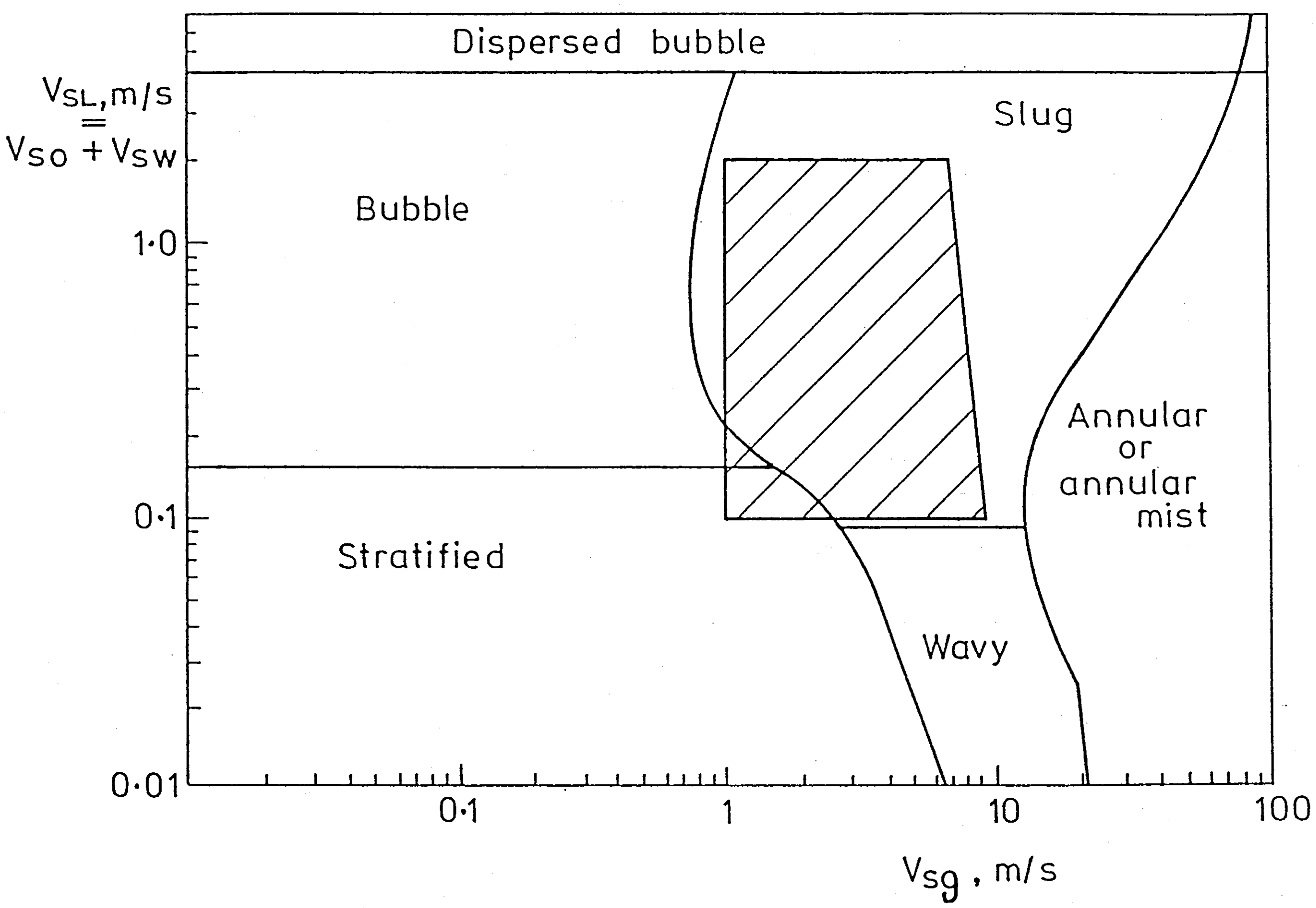


Fig 6.1 Experimental Matrix on Flow Regime Map of Mandhane et al(1974)

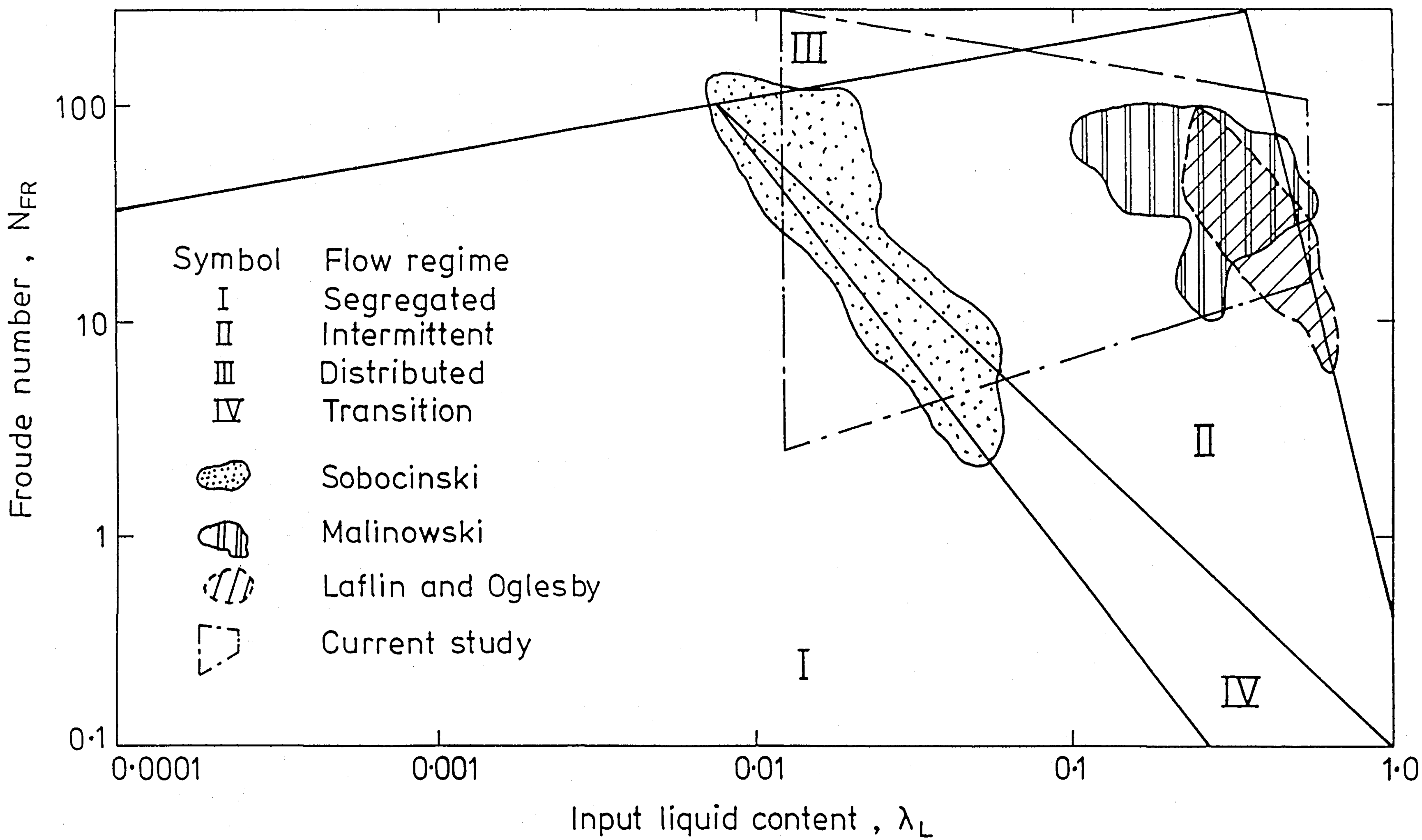


Fig 6.2 Experimental Matrix on Flow Regime Map of Beggs and Brill(1973)

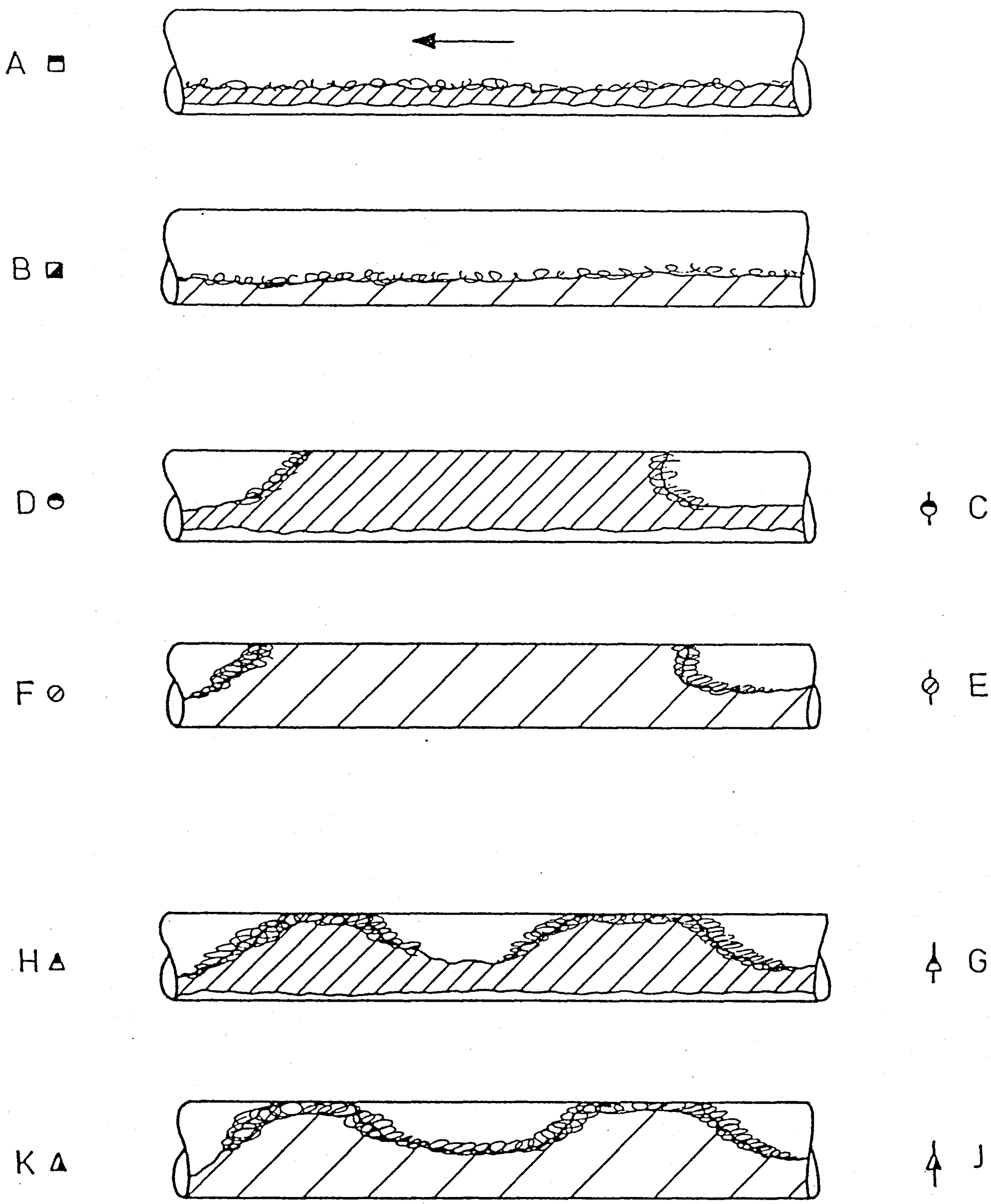


Fig 6.3 Oil/Water/Gas Flow Regime Classifications Used on Figs 6.4-6.13

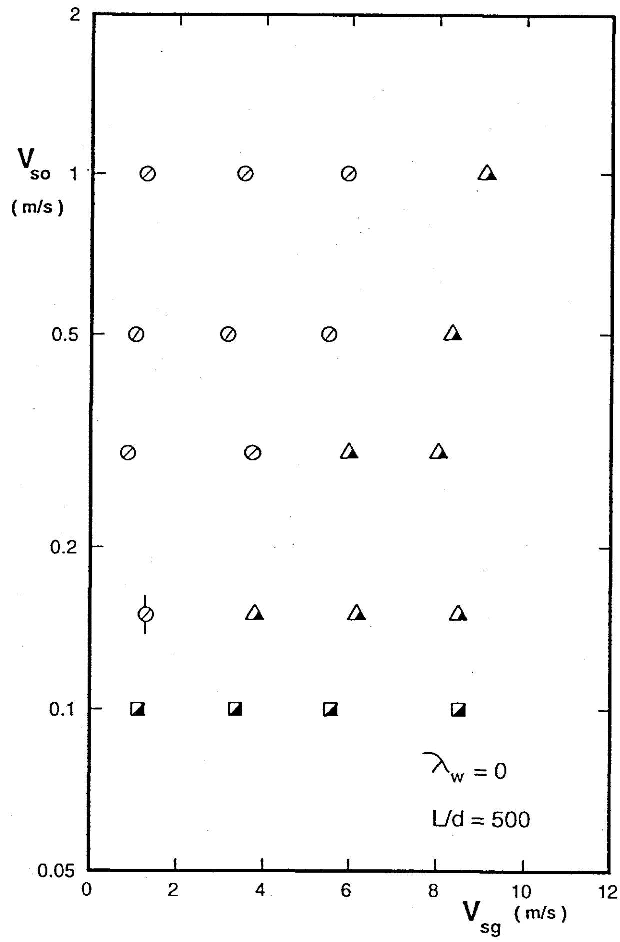


Fig 6.4 Oil/Water/Gas Flow Pattern at L=500d,Oil No1

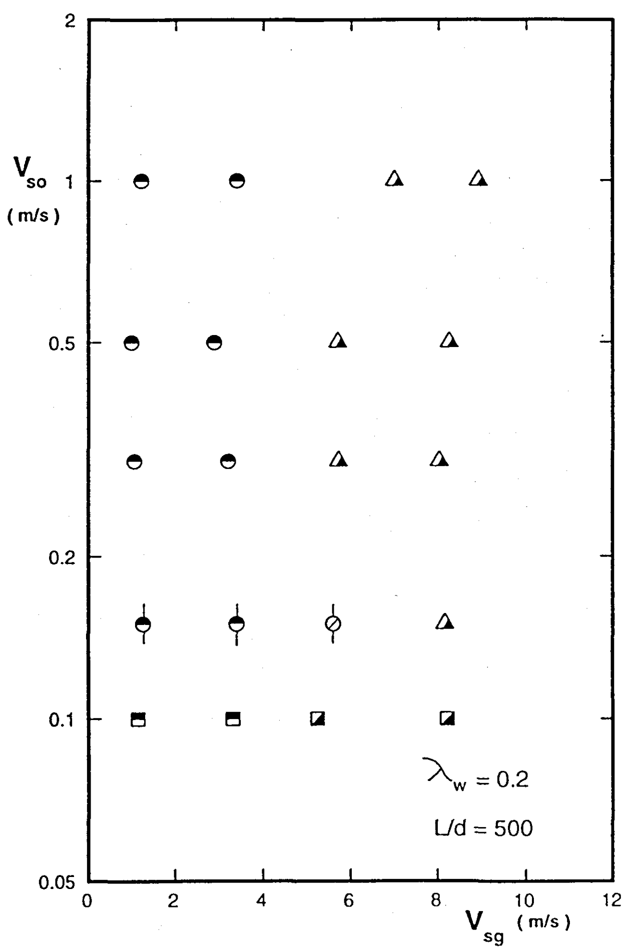


Fig 6.5 Oil/Water/Gas Flow Pattern at L=500d,Oil No1

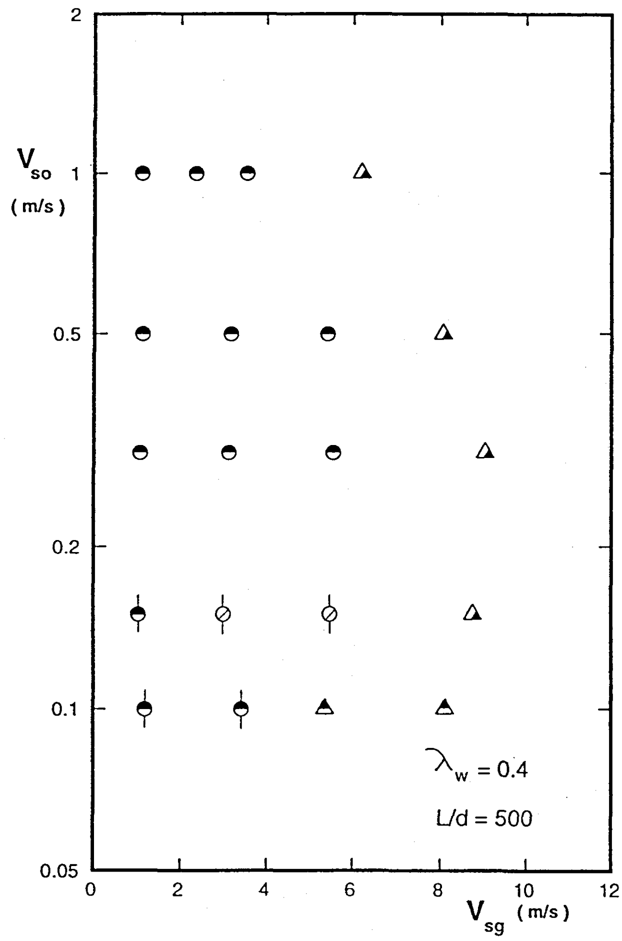


Fig 6.6 Oil/Water/Gas Flow Pattern at L=500d,Oil No1

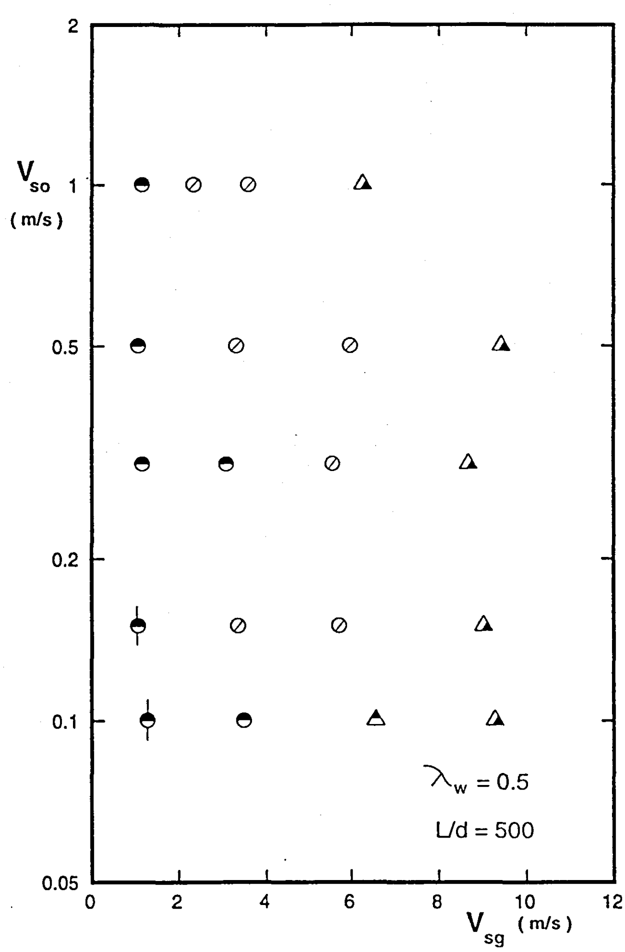


Fig 6.7 Oil/Water/Gas Flow Pattern at L=500d,Oil No1

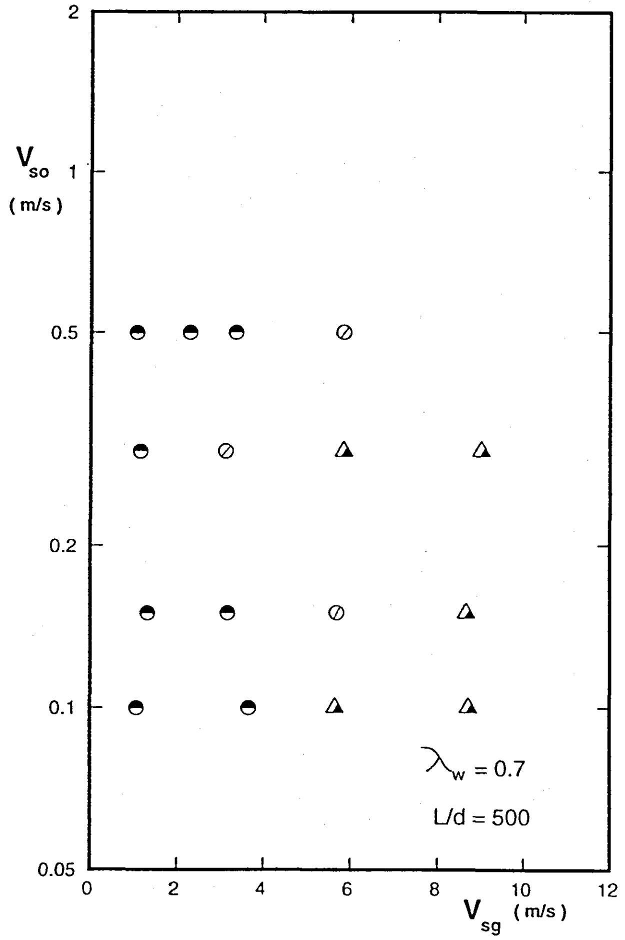


Fig 6.8 Oil/Water/Gas Flow Pattern at L=500d,Oil No1

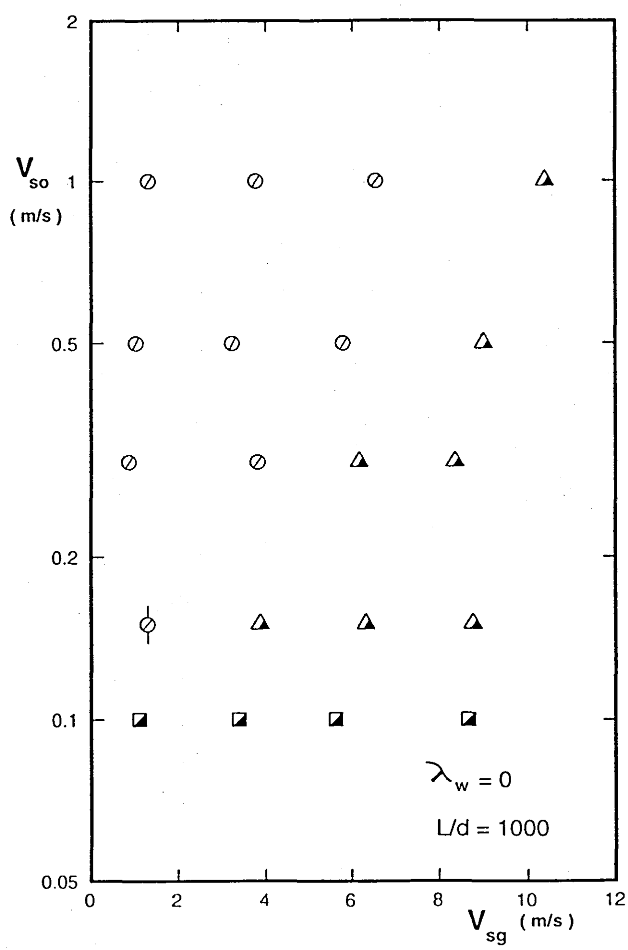


Fig 6.9 Oil/Water/Gas Flow Pattern at L=1000d,Oil No1

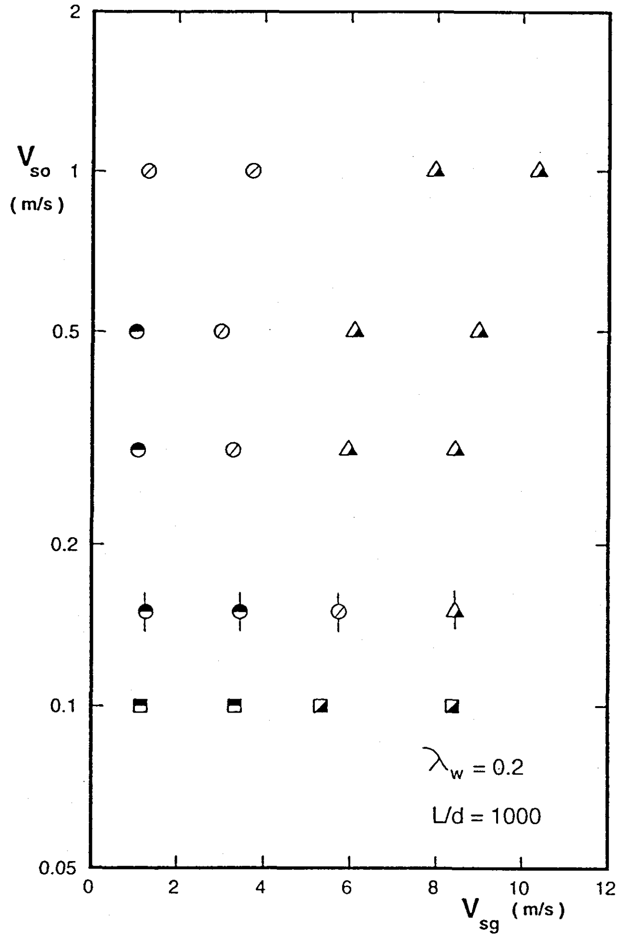


Fig 6.10 Oil/Water/Gas Flow Pattern at L=1000d,Oil No1

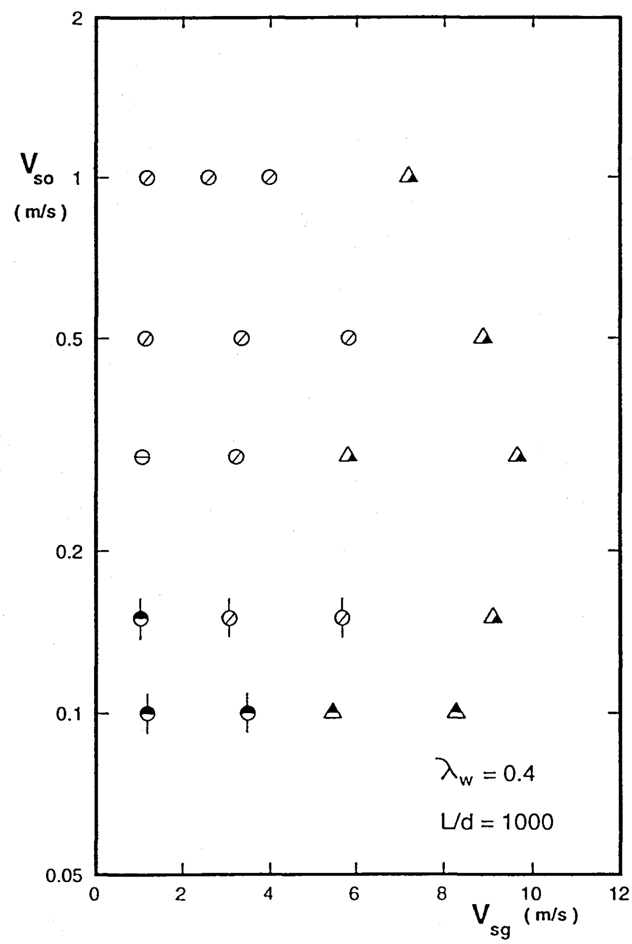


Fig 6.11 Oil/Water/Gas Flow Pattern at L=1000d,Oil No1

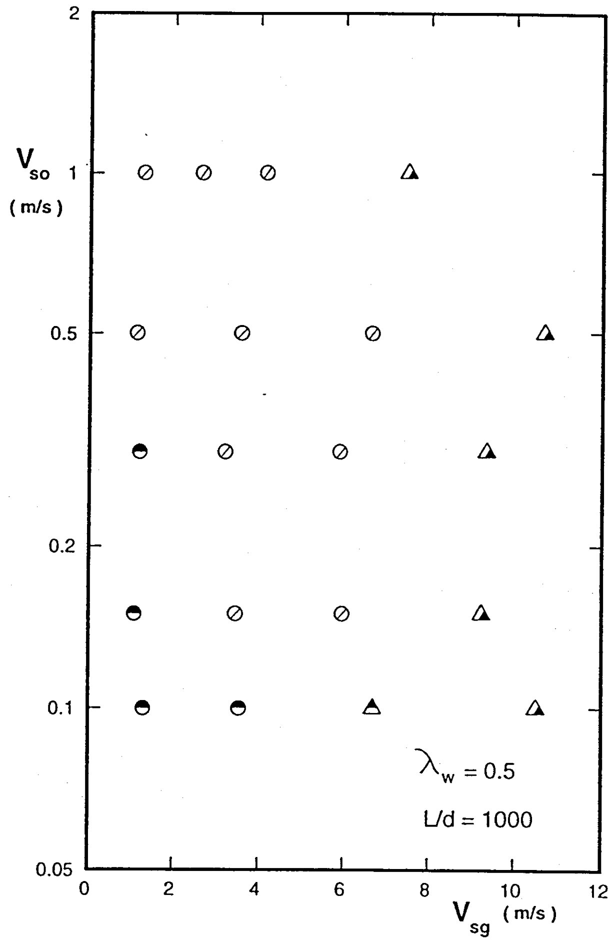


Fig 6.12 Oil/Water/Gas Flow Pattern at L=1000d,Oil No1

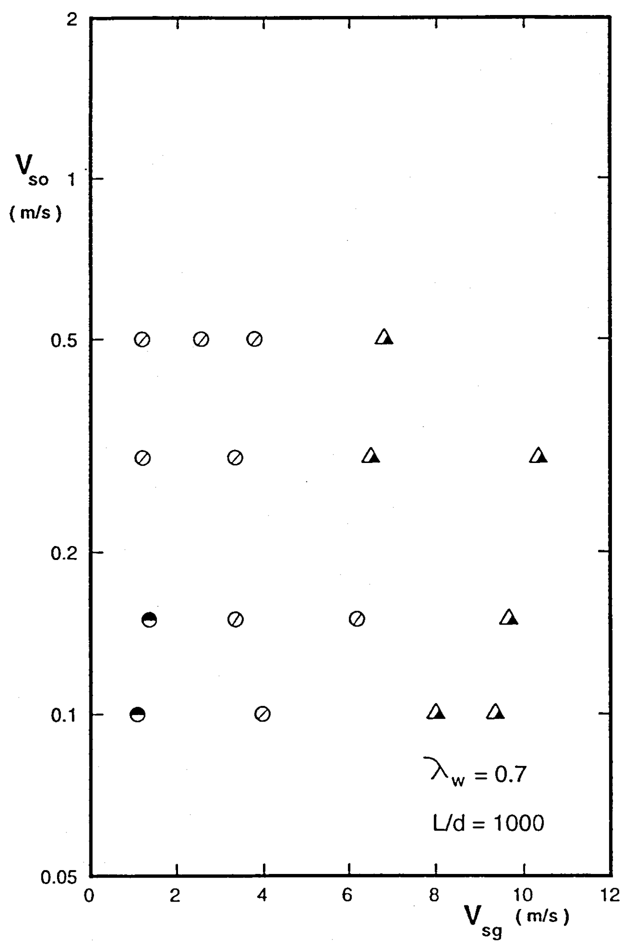
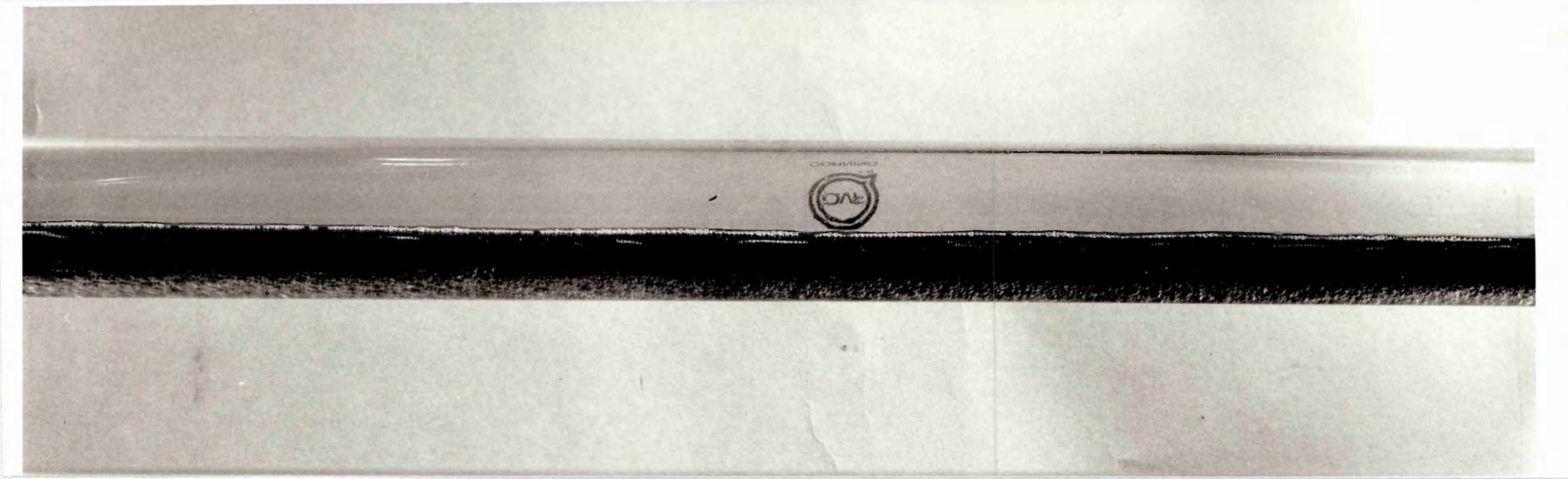
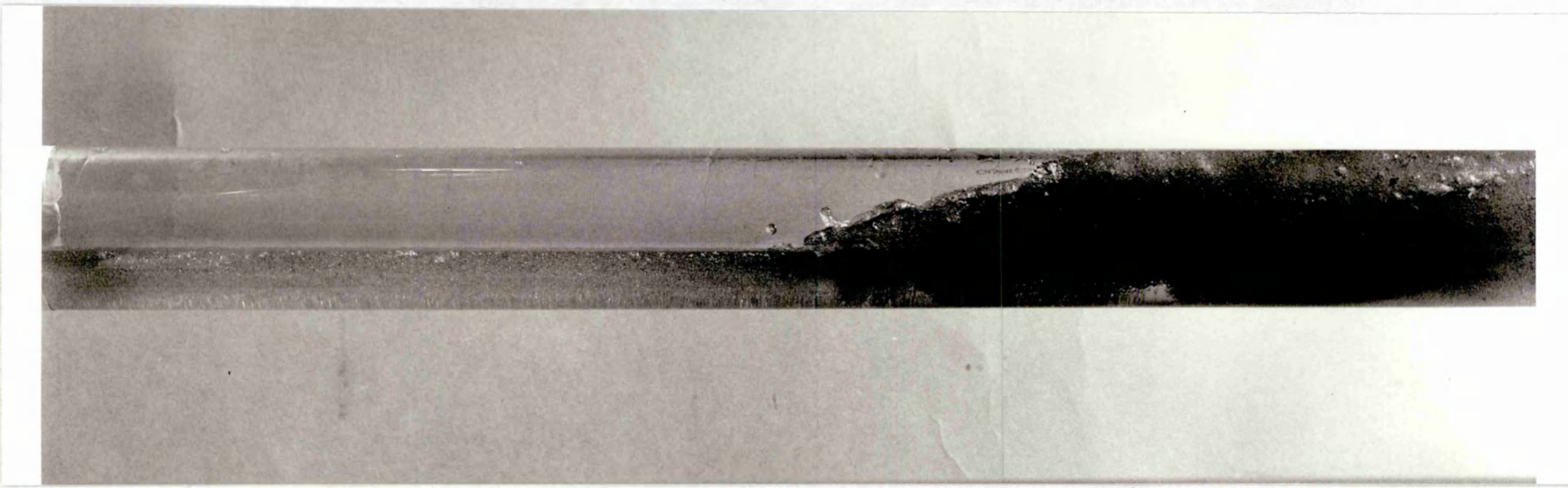


Fig 6.13 Oil/Water/Gas Flow Pattern at L=1000d,Oil No1

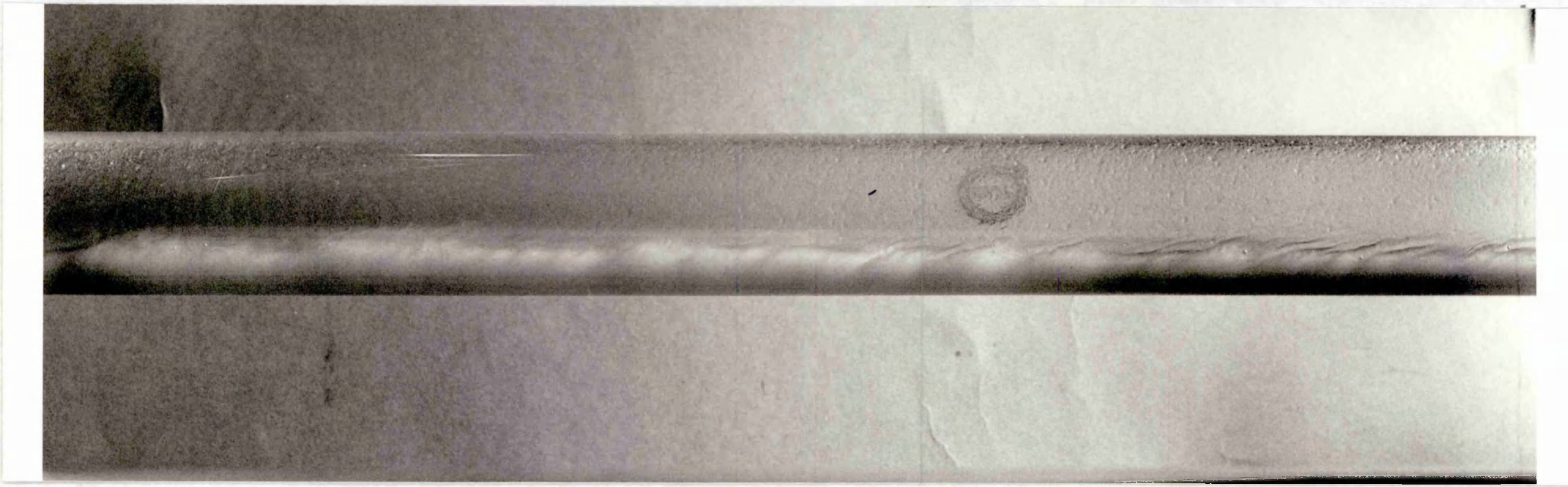


(a) $V_{so} = 0.1 m/s, \lambda_w = 0.2, V_{sg} = 1$ m/s (Flow R to L)



(b) $V_{so} = 0.1 m/s, \lambda_w = 0.7, V_{sg} = 1$ m/s (Flow R to L)

Fig 6.14 Oil/Water/Gas Flow Pattern Photographs at L=1000d,Oil No1



(c) $V_{so} = 0.3 m/s, \lambda_w = 0.7, V_{sg} = 5.5$ m/s (Flow R to L)

Fig 6.14 Oil/Water/Gas Flow Pattern Photographs at L=1000d,Oil No1

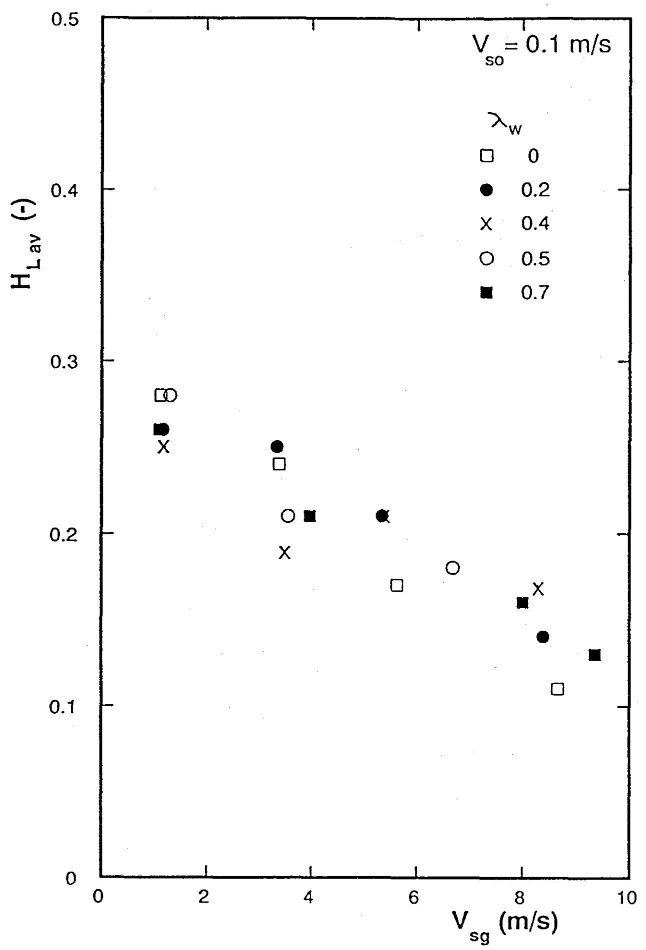


Fig 6.15 Oil/Water/Gas Total Liquids Holdup,Oil No1

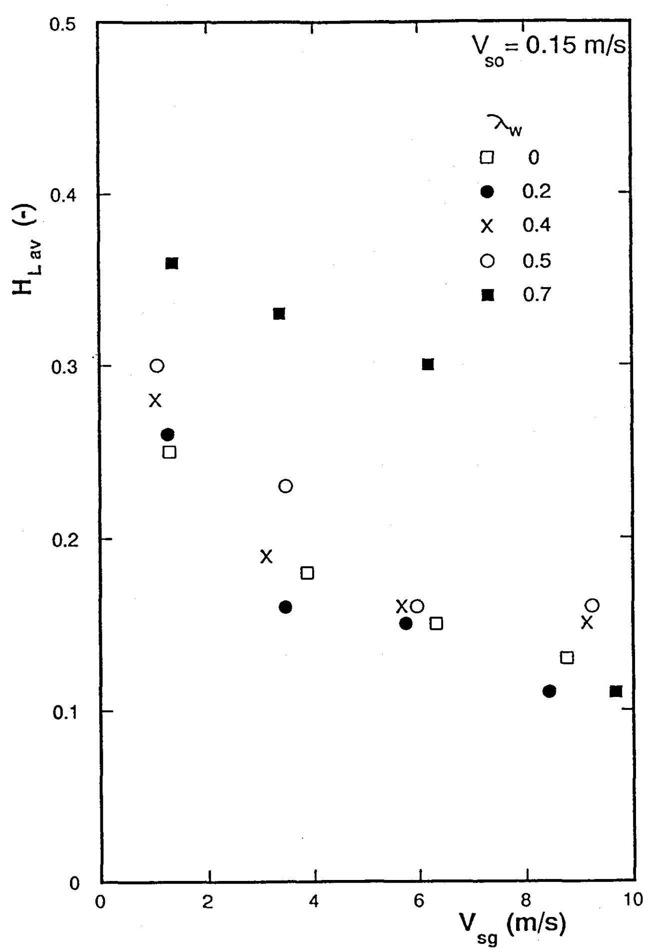


Fig 6.16 Oil/Water/Gas Total Liquids Holdup,Oil No1

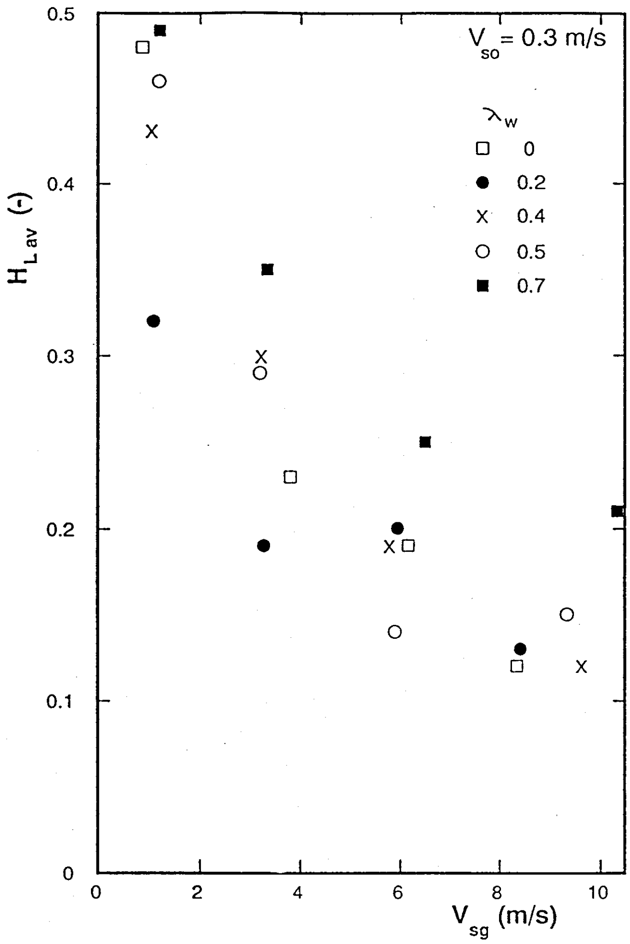


Fig 6.17 Oil/Water/Gas Total Liquids Holdup,Oil No1

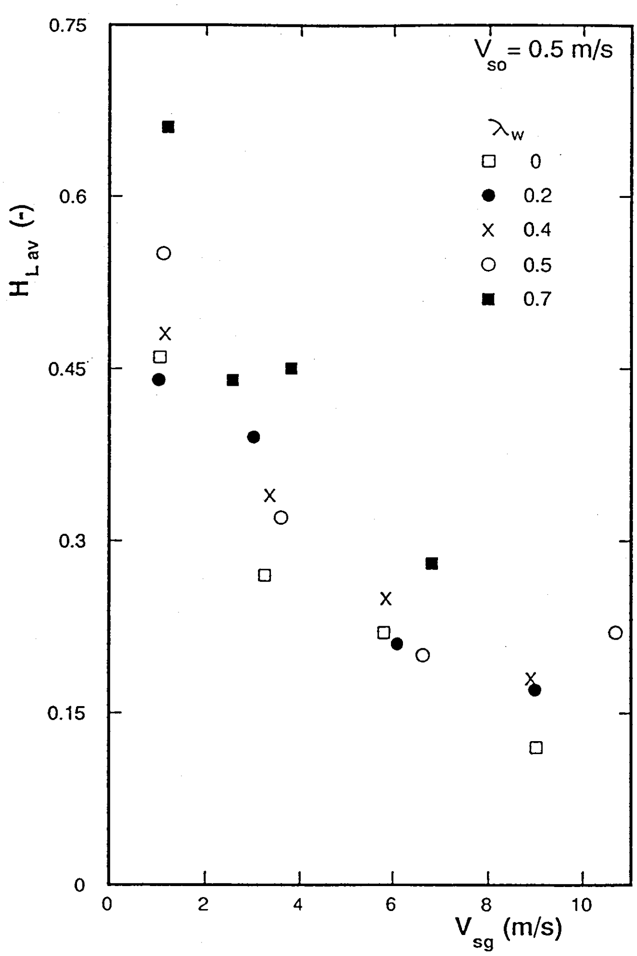


Fig 6.18 Oil/Water/Gas Total Liquids Holdup,Oil No1

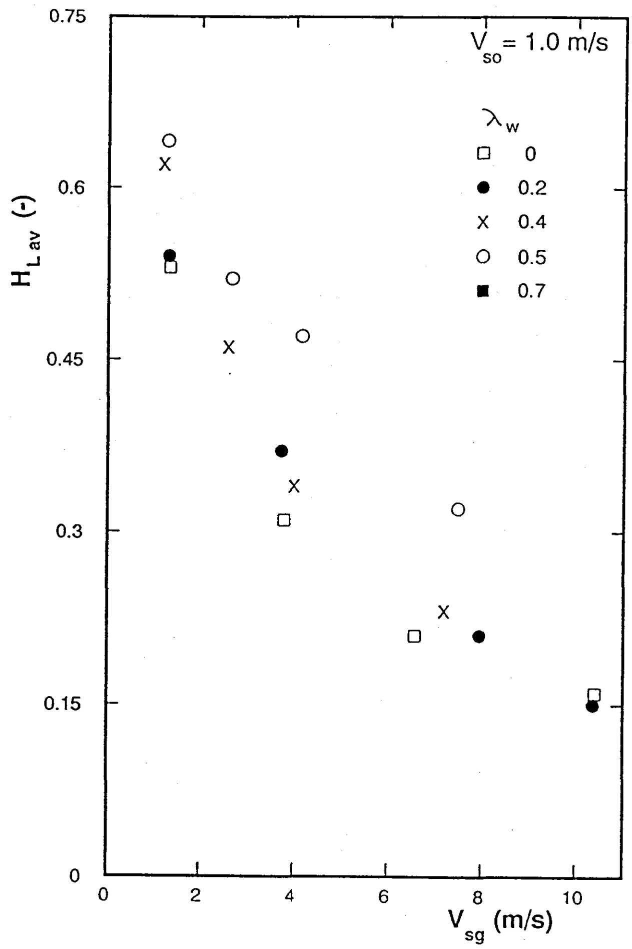


Fig 6.19 Oil/Water/Gas Total Liquids Holdup,Oil No1

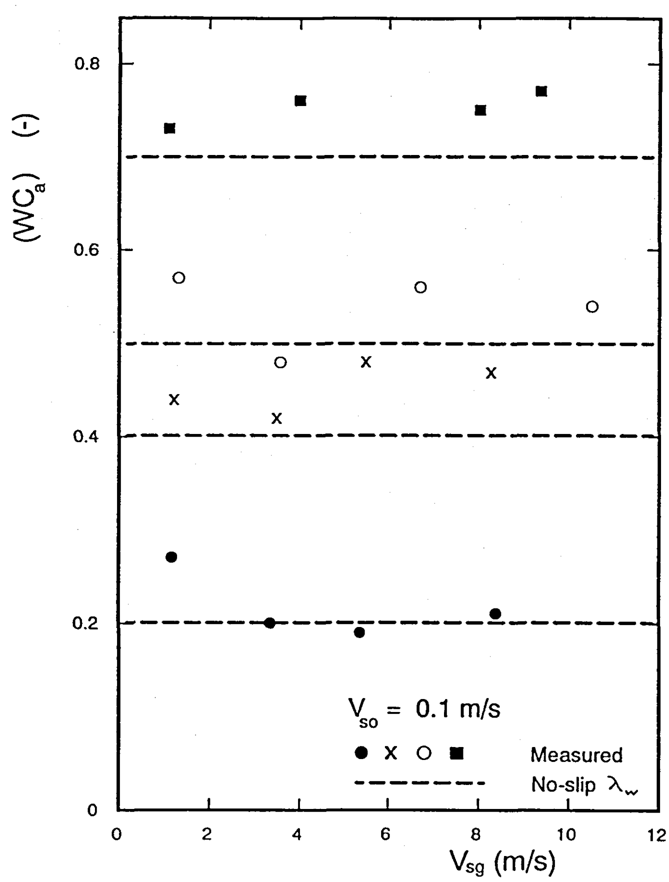


Fig 6.20 Oil/Water/Gas In-situ Water Fractions,Oil No1

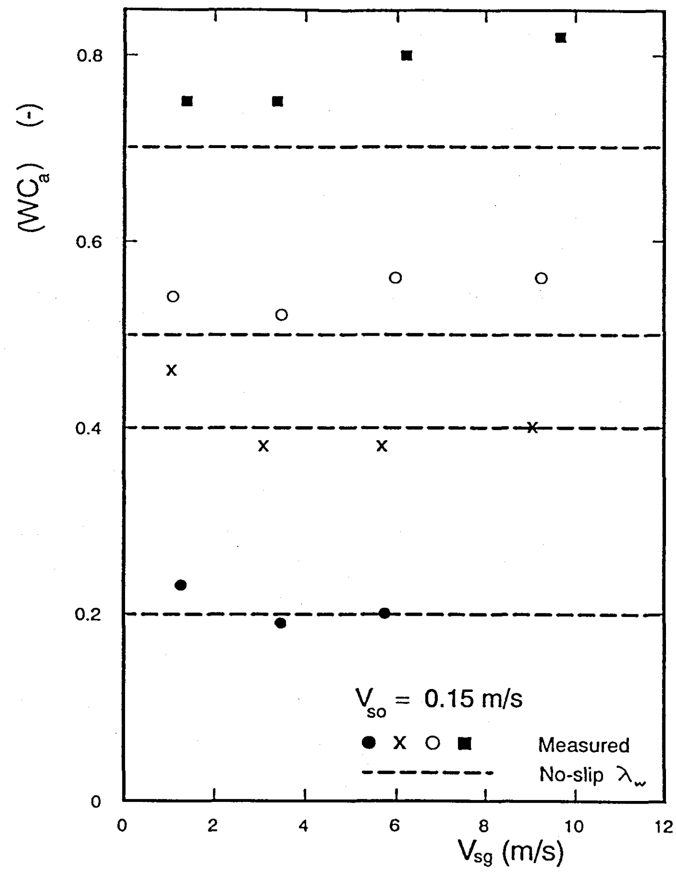


Fig 6.21 Oil/Water/Gas In-situ Water Fractions,Oil No1

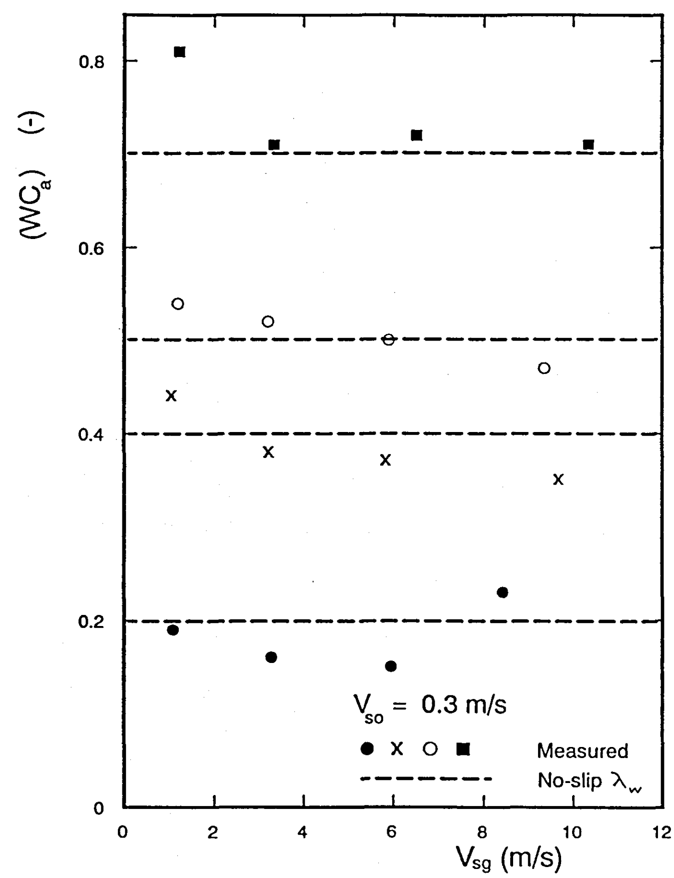


Fig 6.22 Oil/Water/Gas In-situ Water Fractions,Oil No1

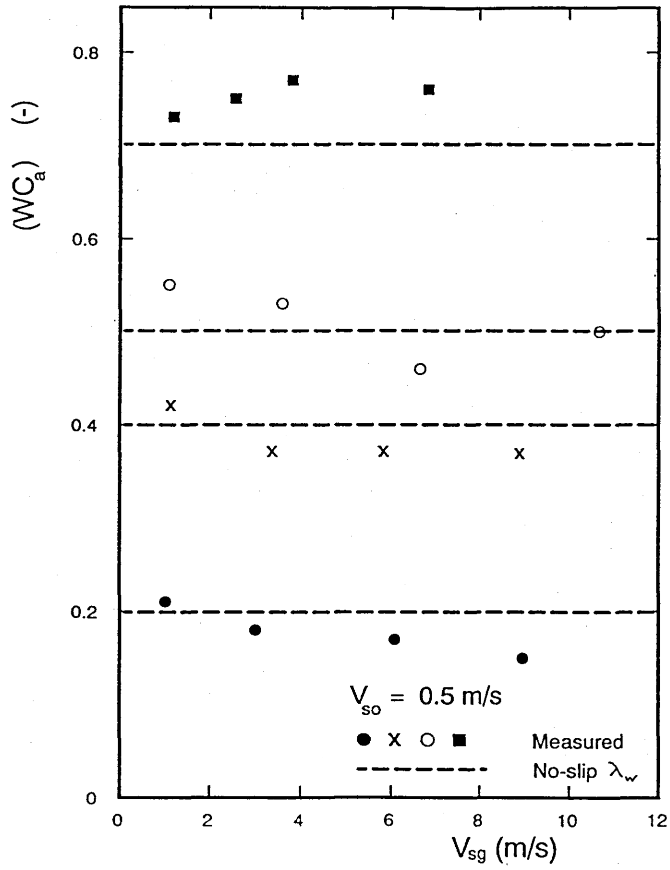


Fig 6.23 Oil/Water/Gas In-situ Water Fractions,Oil No1

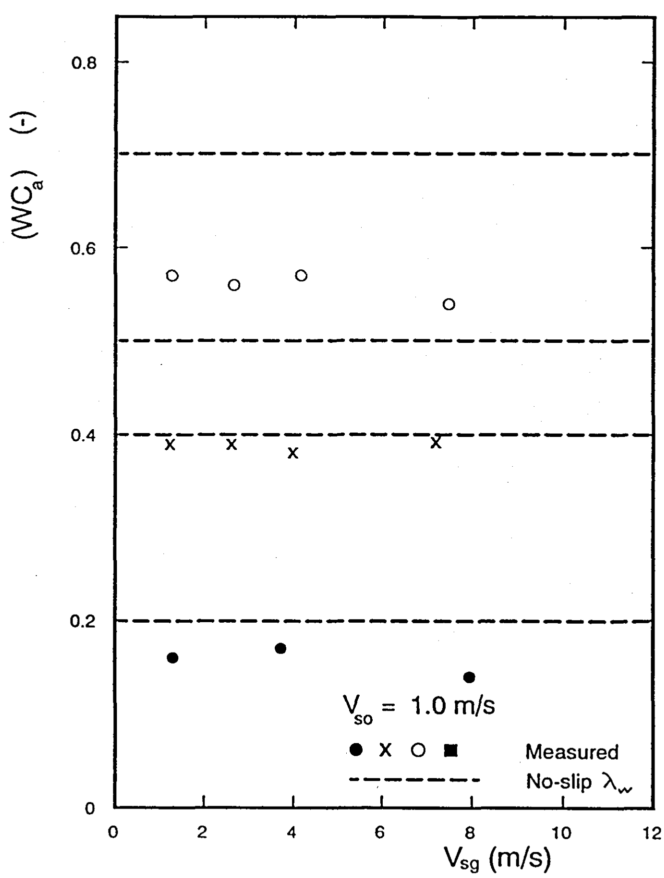


Fig 6.24 Oil/Water/Gas In-situ Water Fractions,Oil No1

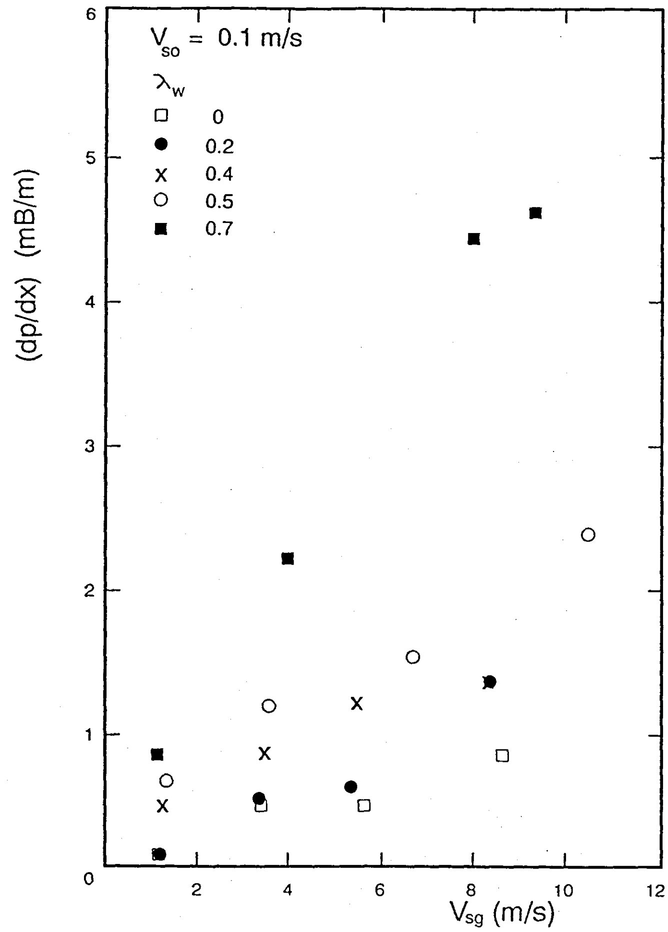


Fig 6.25 Oil/Water/Gas Pressure Drop,Oil No1

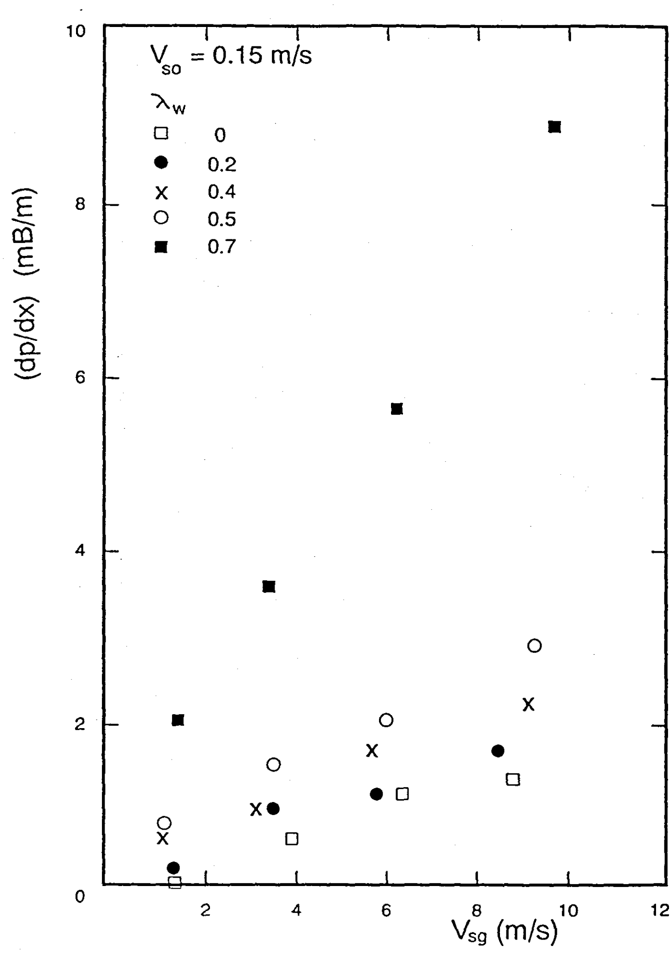


Fig 6.26 Oil/Water/Gas Pressure Drop,Oil No1

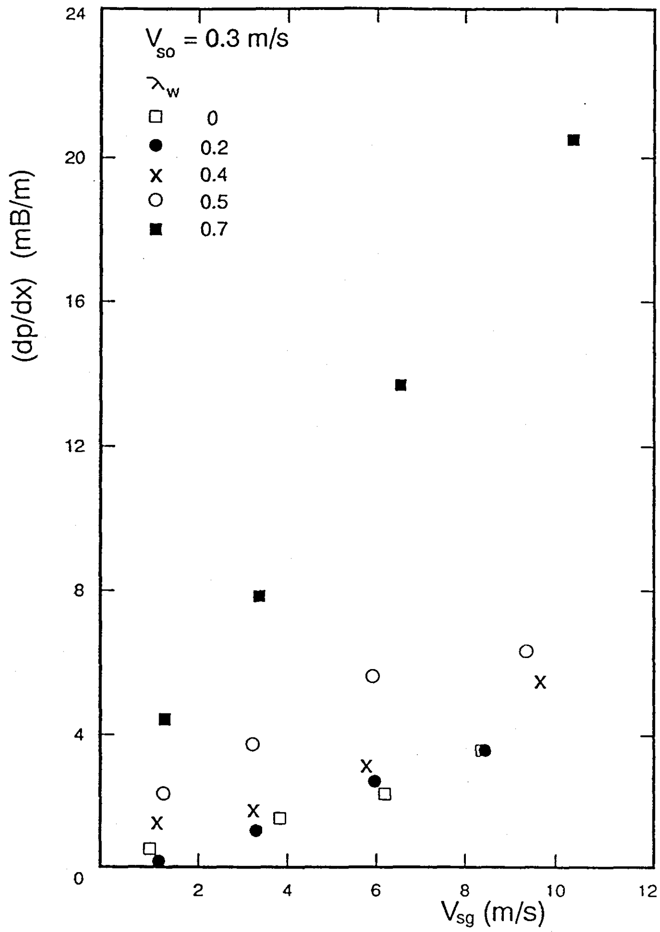


Fig 6.27 Oil/Water/Gas Pressure Drop,Oil No1

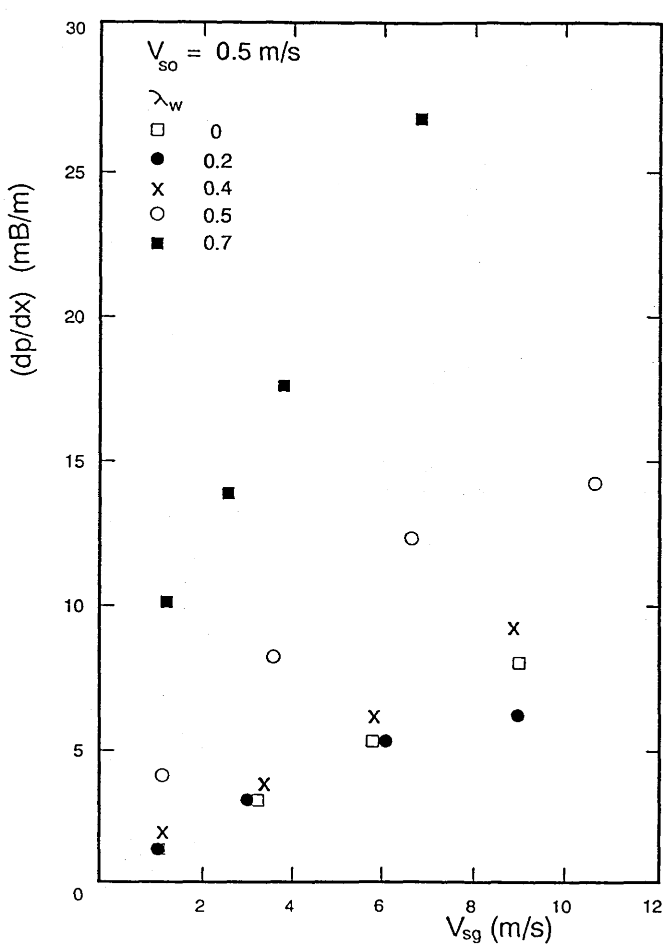


Fig 6.28 Oil/Water/Gas Pressure Drop,Oil No1

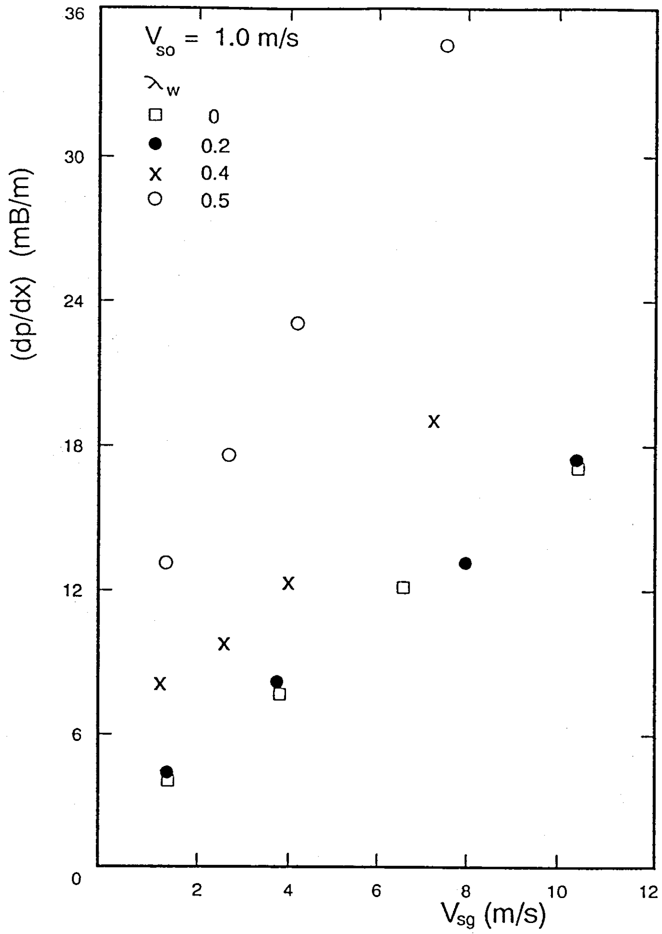


Fig 6.29 Oil/Water/Gas Pressure Drop,Oil No1

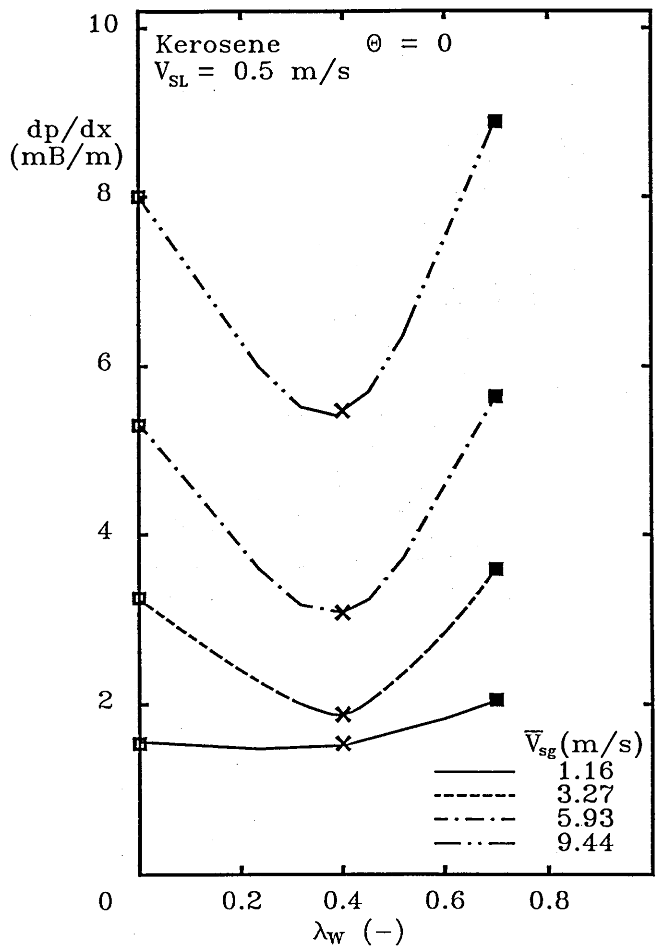


Fig 6.30 Oil/Water/Gas Pressure Drop,V_{sL} = 0.5 m/s,Oil No 1

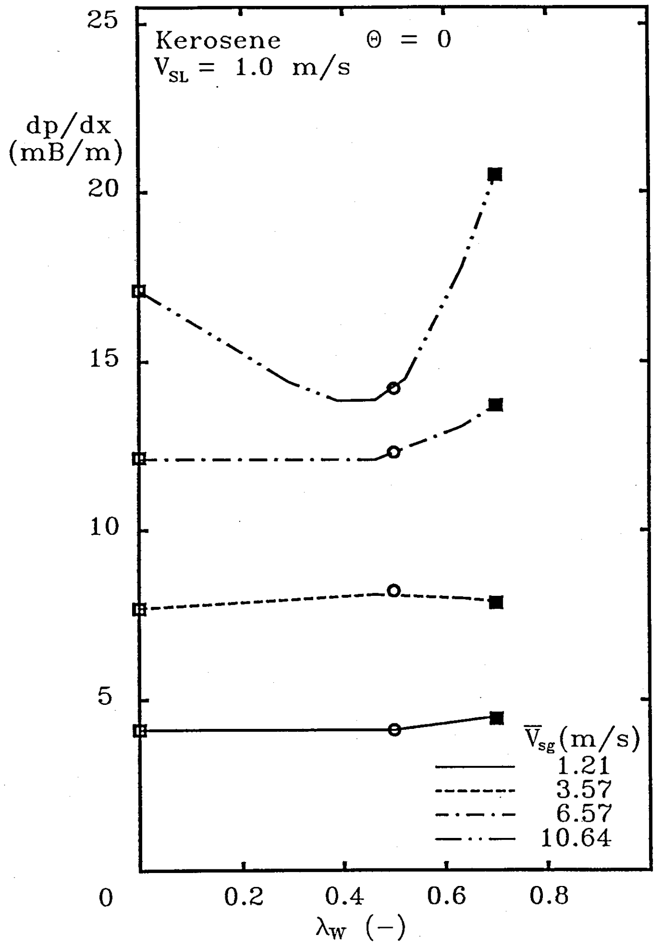


Fig 6.31 Oil/Water/Gas Pressure Drop, V_{sL} = 1.0 m/s, Oil No 1

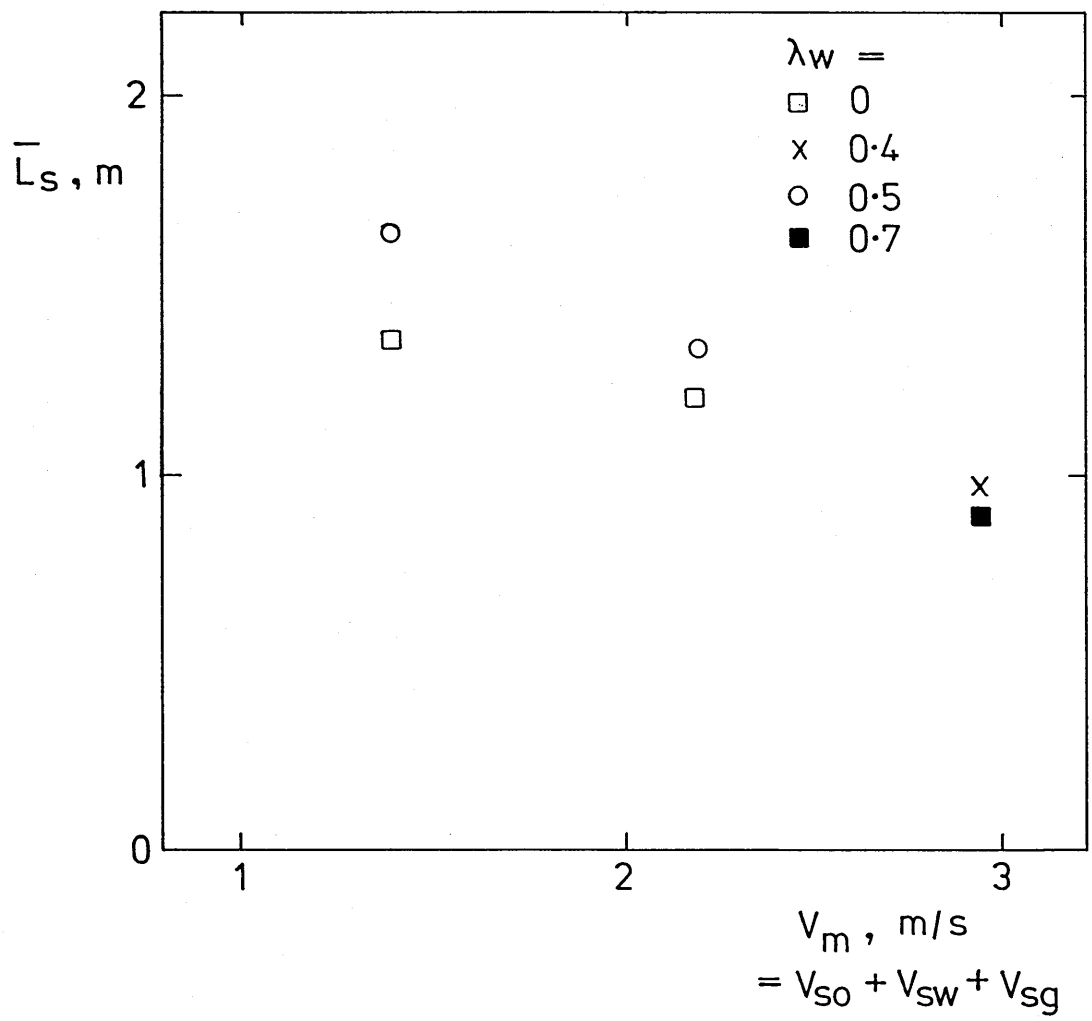


Fig 6.32 Oil/Water/Gas Average Slug Length at L=1000d, Oil No1

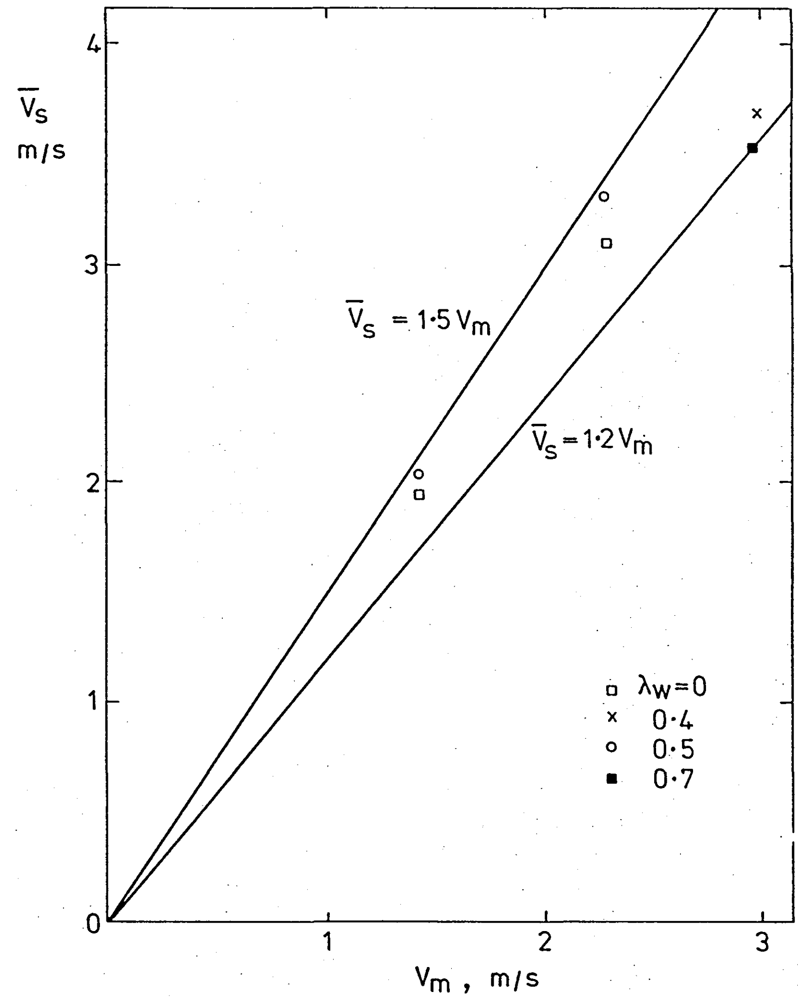


Fig 6.33 Oil/Water/Gas Average Slug Front Velocity at L=1000d,Oil No1

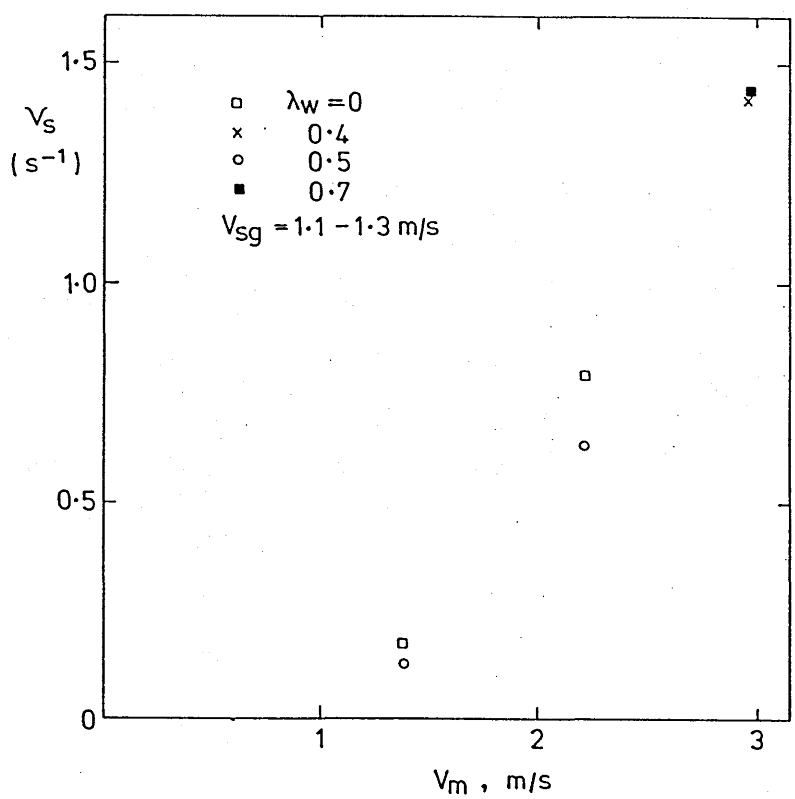


Fig 6.34 Oil/Water/Gas Average Slug Frequency at L=1000d,Oil No1

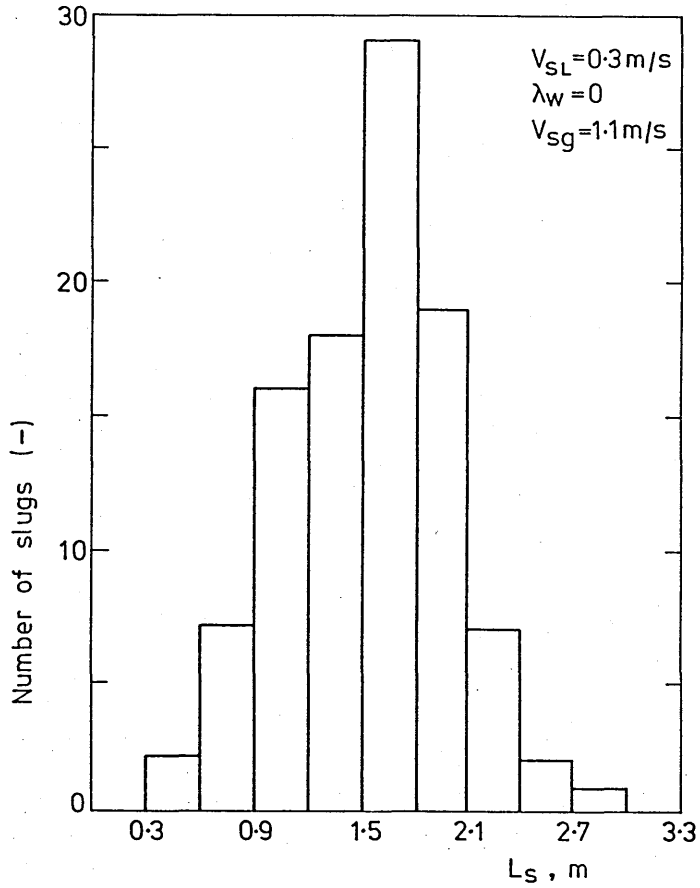


Fig 6.35 Oil/Water/Gas Average Slug Length Distribution at L=1000d,Oil No1

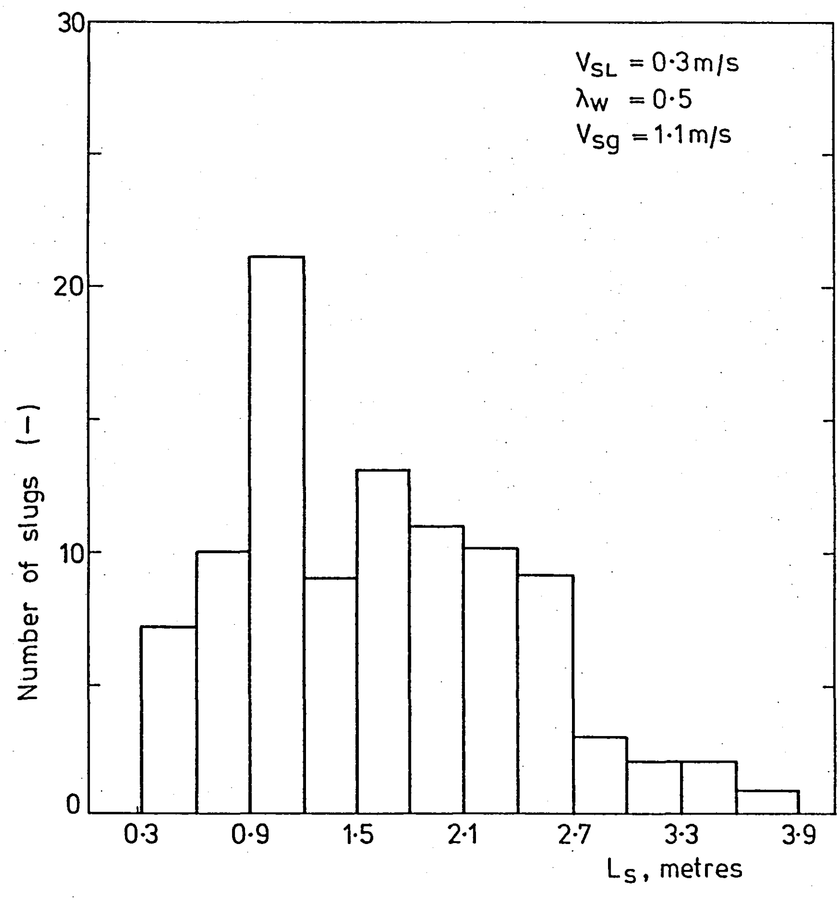


Fig 6.36 Oil/Water/Gas Average Slug Length Distribution at L=1000d,Oil No1

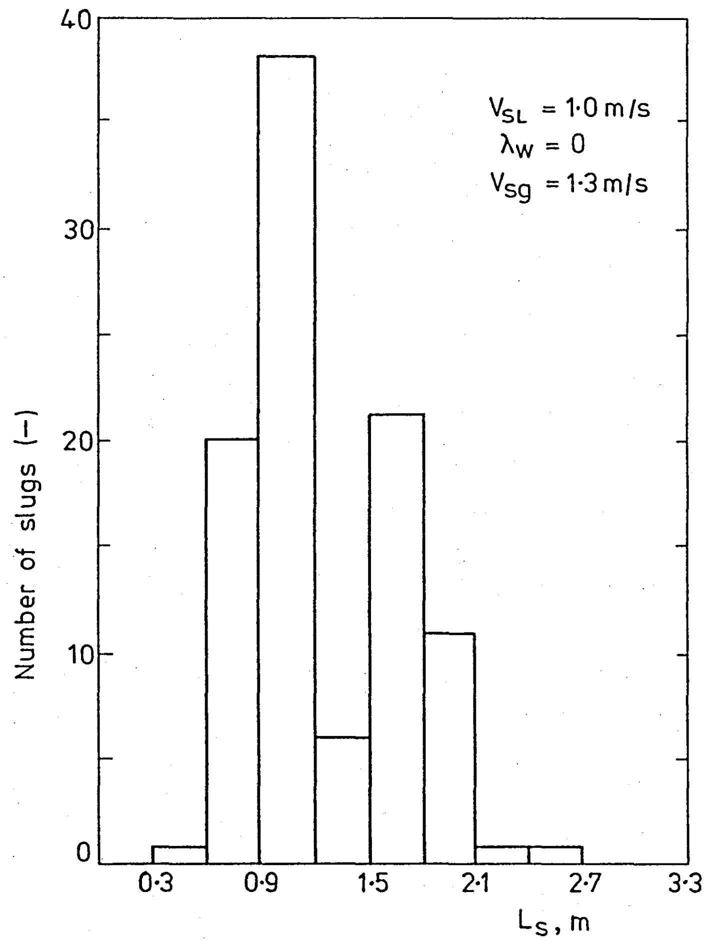


Fig 6.37 Oil/Water/Gas Average Slug Length Distribution at L=1000d,Oil No1

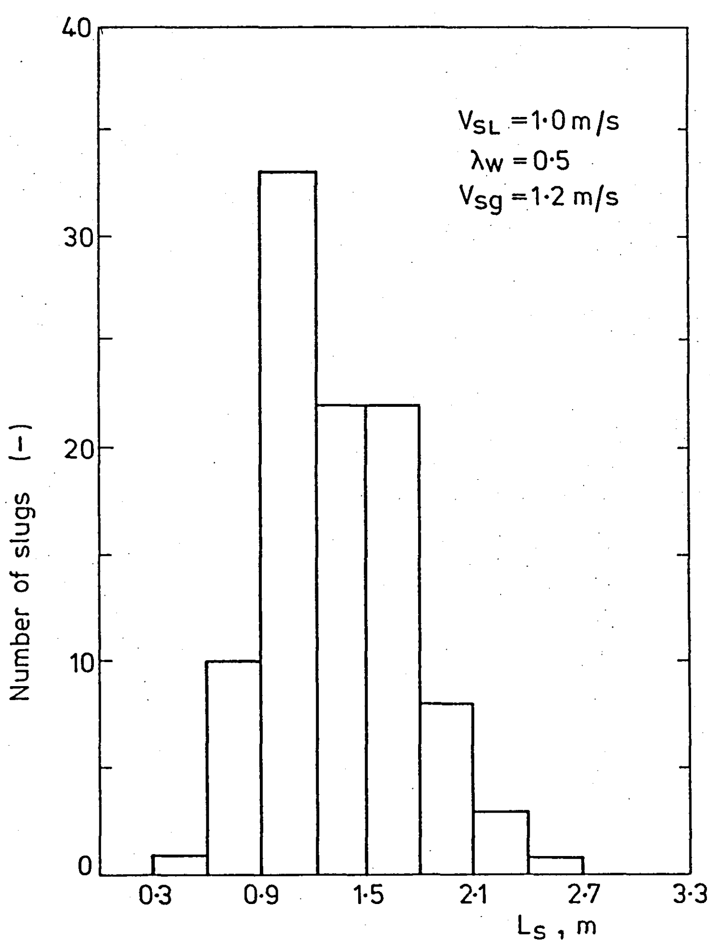


Fig 6.38 Oil/Water/Gas Average Slug Length Distribution at L=1000d,Oil No1

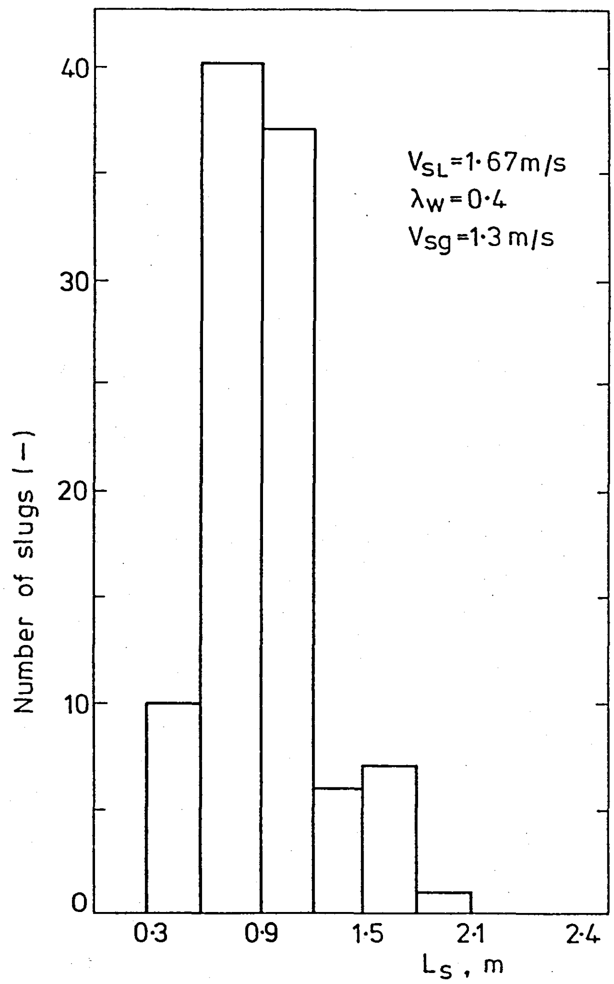


Fig 6.39 Oil/Water/Gas Average Slug Length Distribution at L=1000d,Oil No1

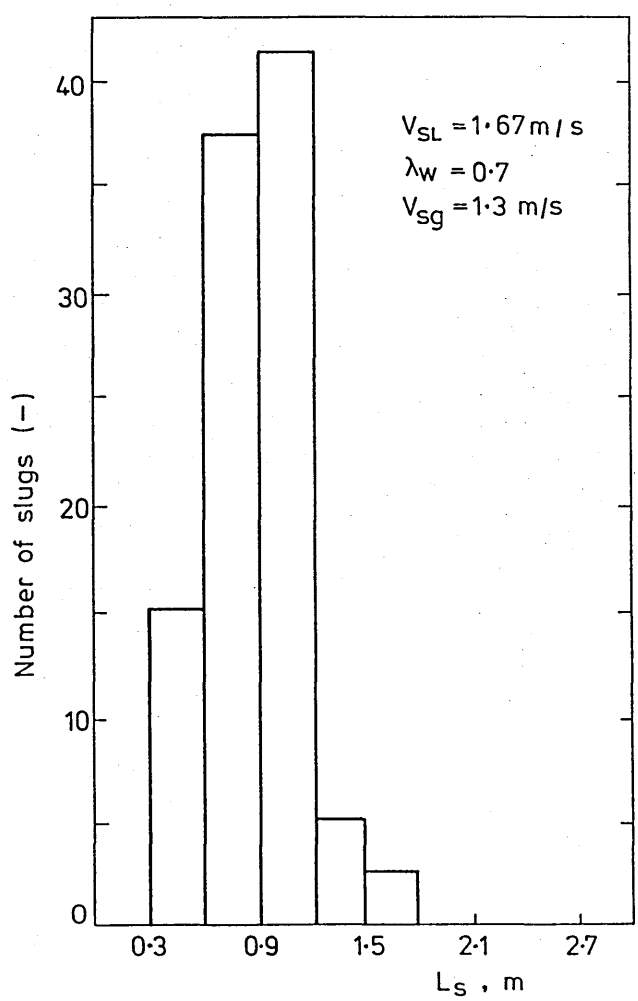


Fig 6.40 Oil/Water/Gas Average Slug Length Distribution at L=1000d,Oil No1

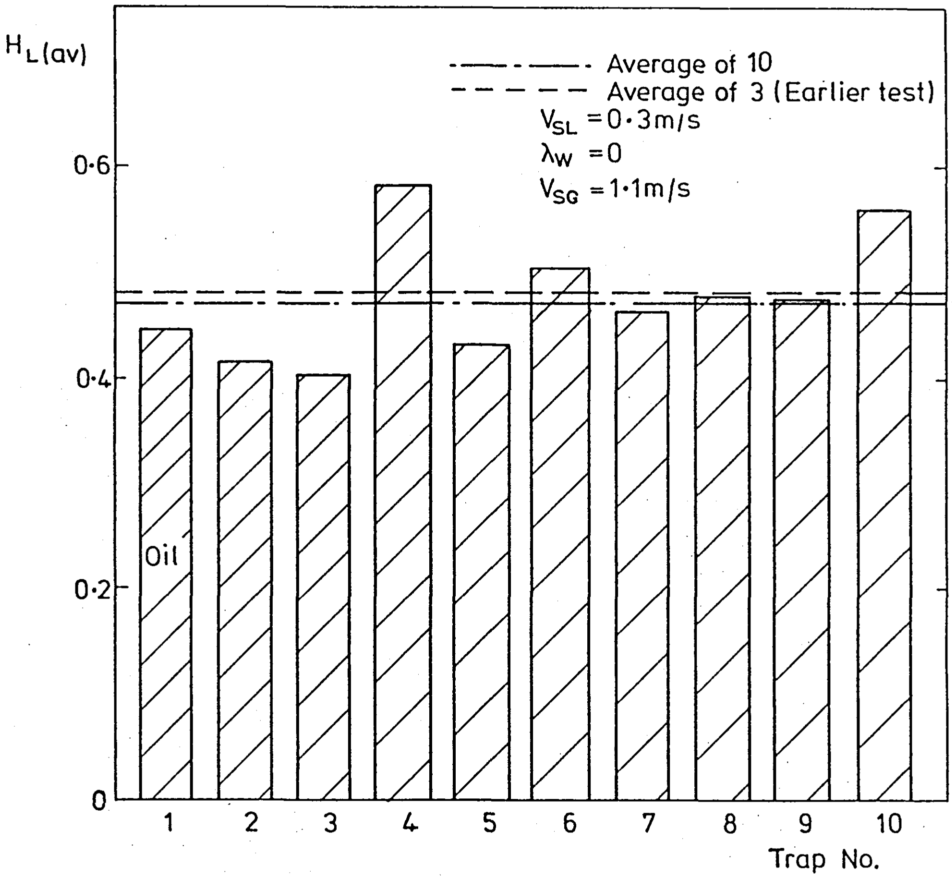


Fig 6.41 Oil/Water/Gas Average Holdup Data,Oil No1

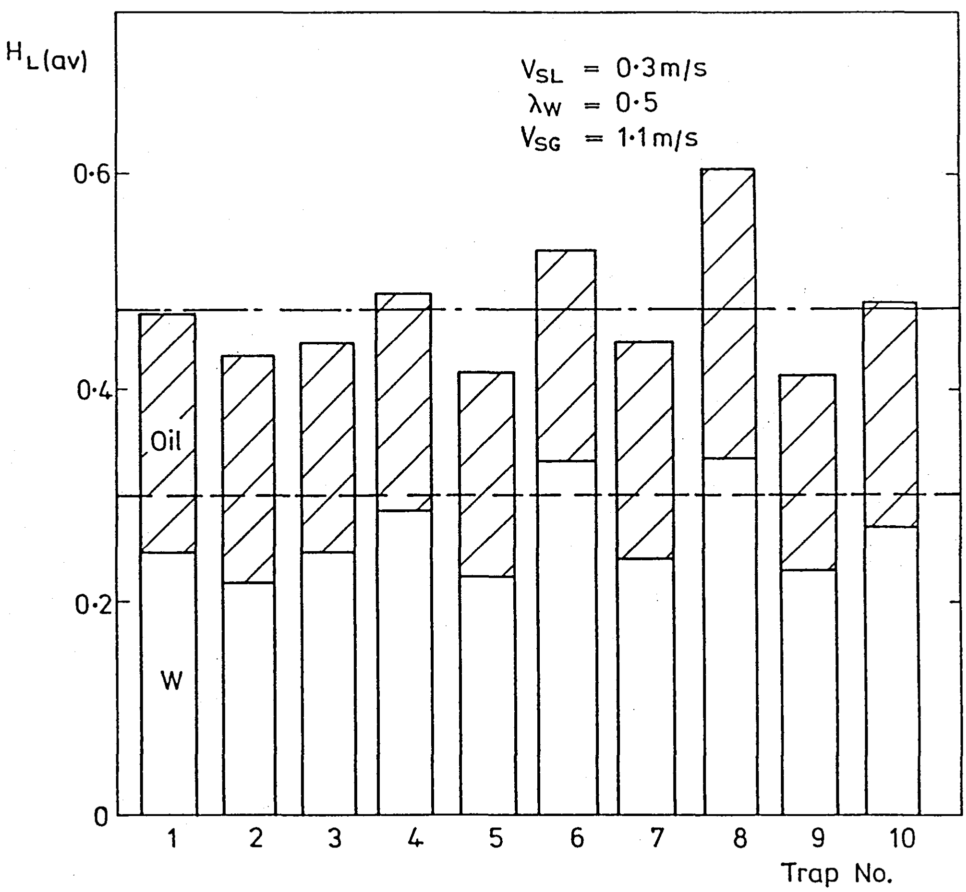


Fig 6.42 Oil/Water/Gas Average Holdup Data,Oil No1

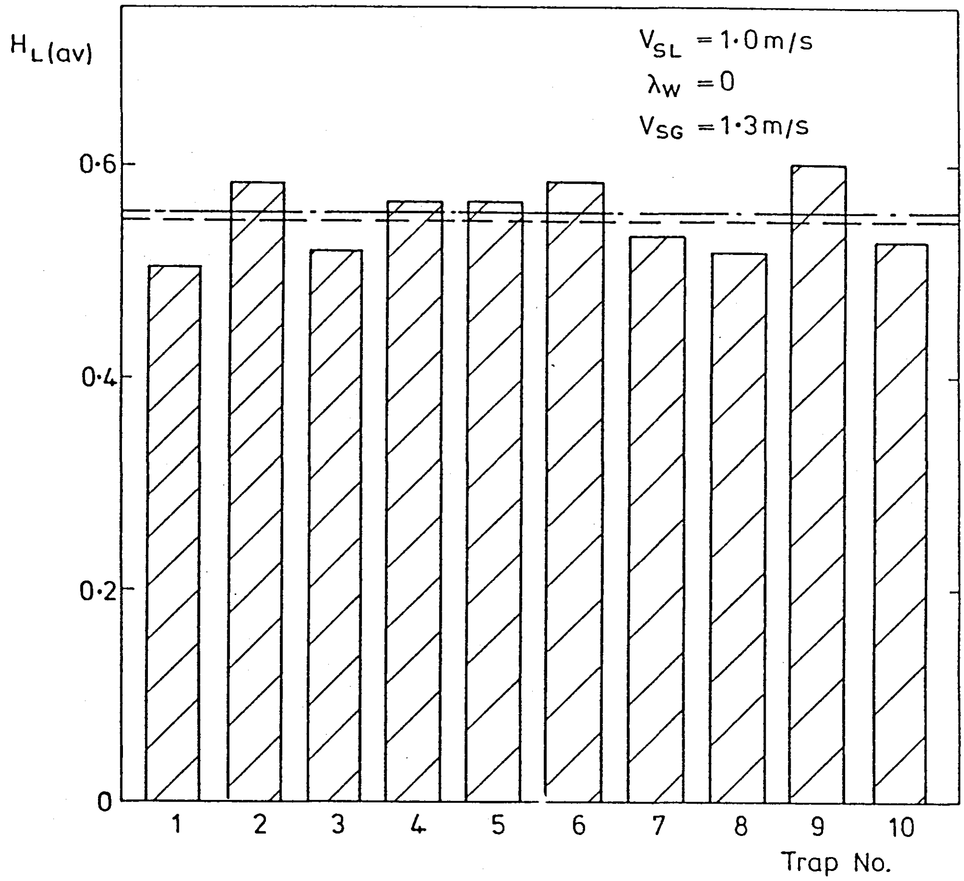


Fig 6.43 Oil/Water/Gas Average Holdup Data,Oil No1

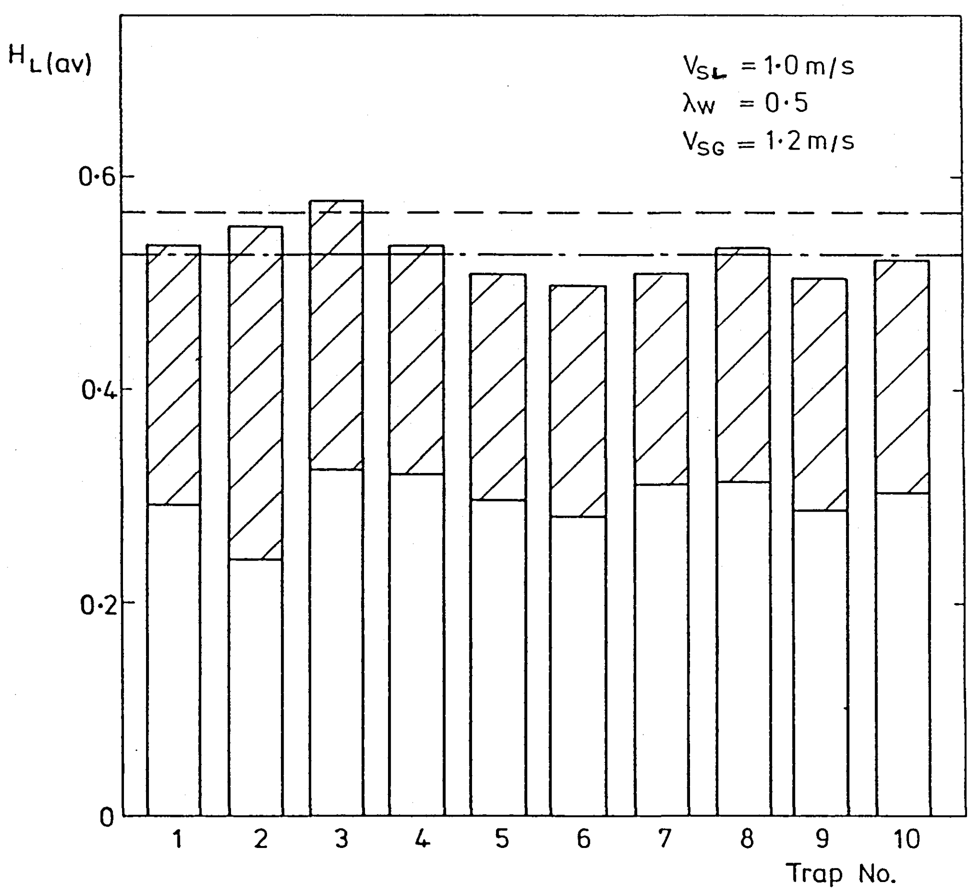


Fig 6.44 Oil/Water/Gas Average Holdup Data,Oil No1

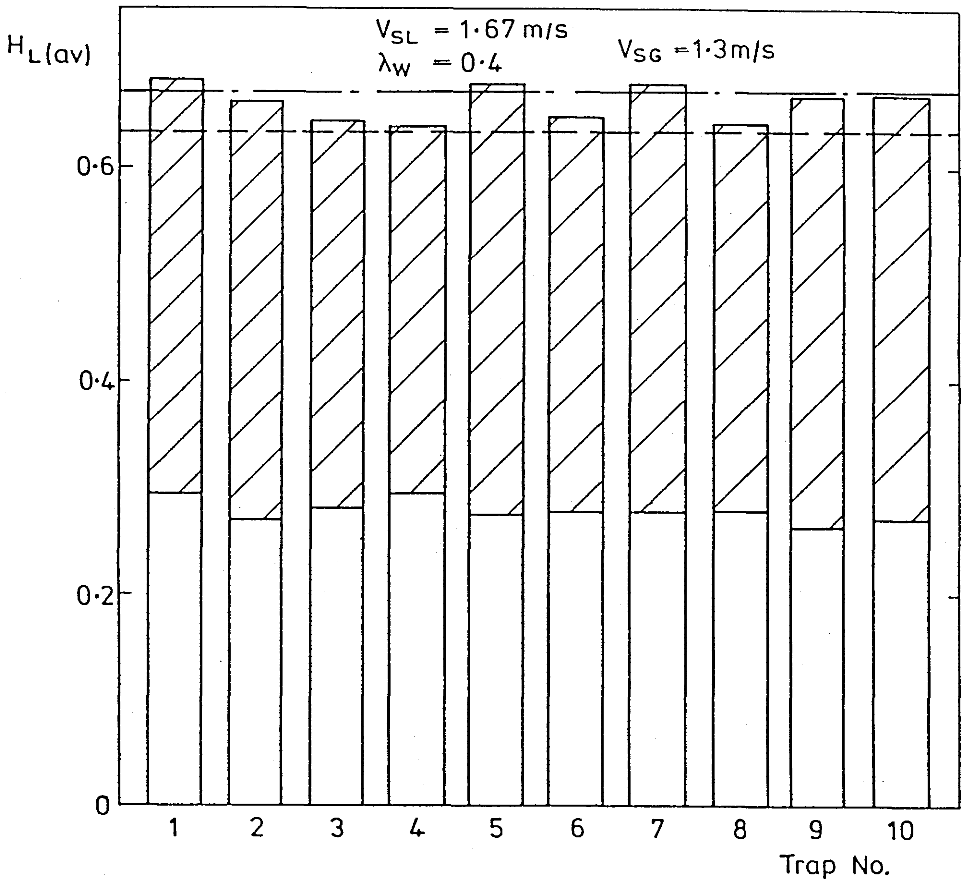


Fig 6.45 Oil/Water/Gas Average Holdup Data,Oil No1

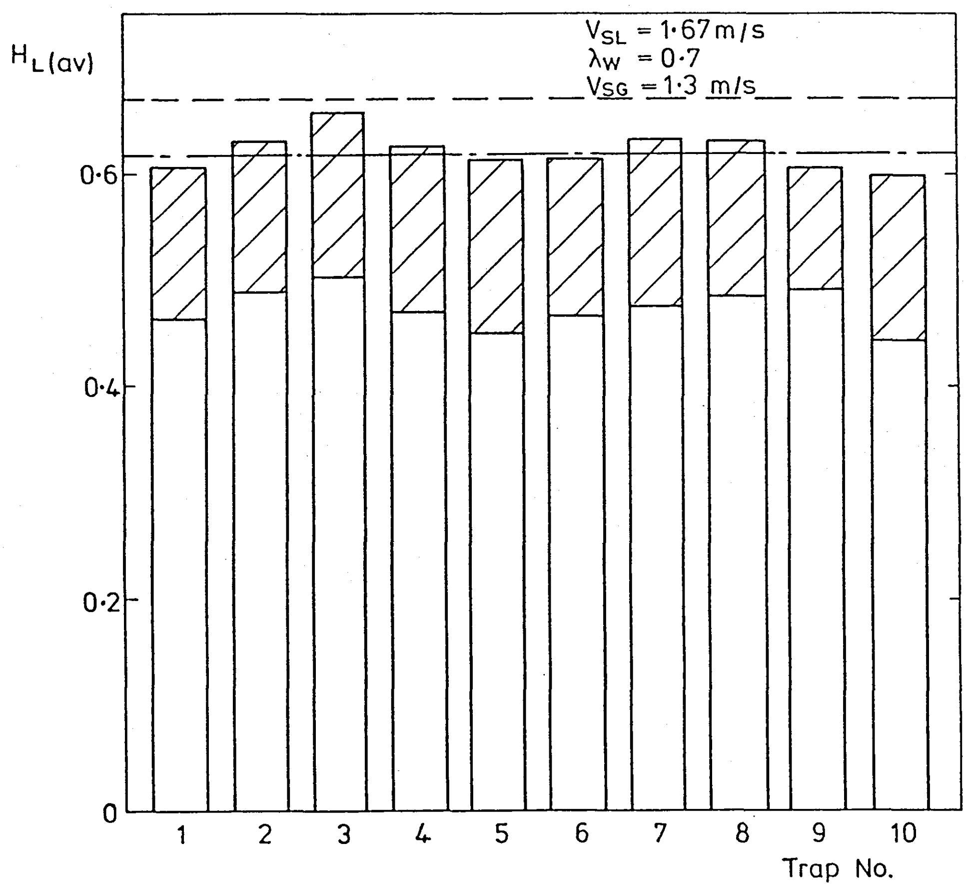


Fig 6.46 Oil/Water/Gas Average Holdup Data,Oil No1

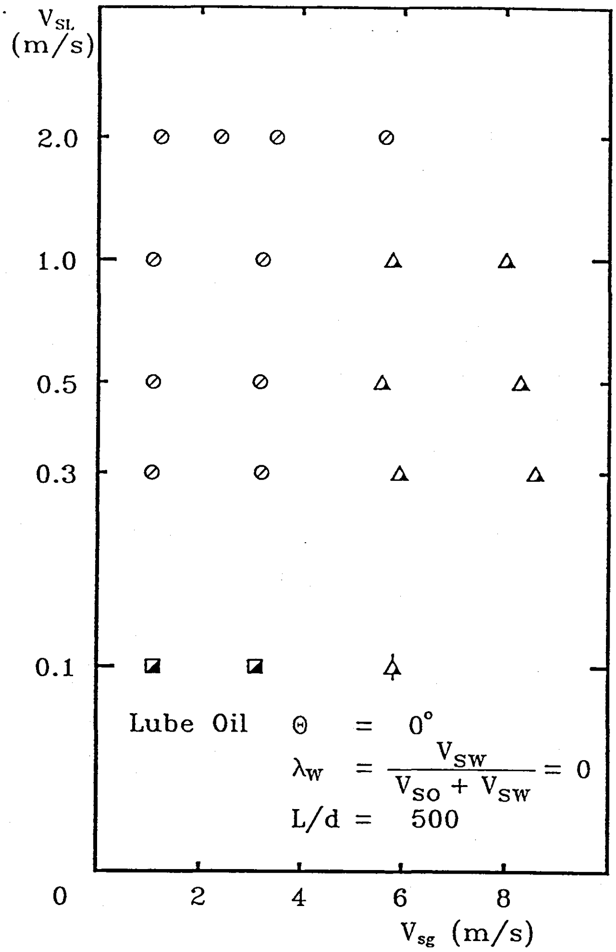


Fig 6.47 Oil/Water/Gas Flow Regime,L=500d,Oil No2

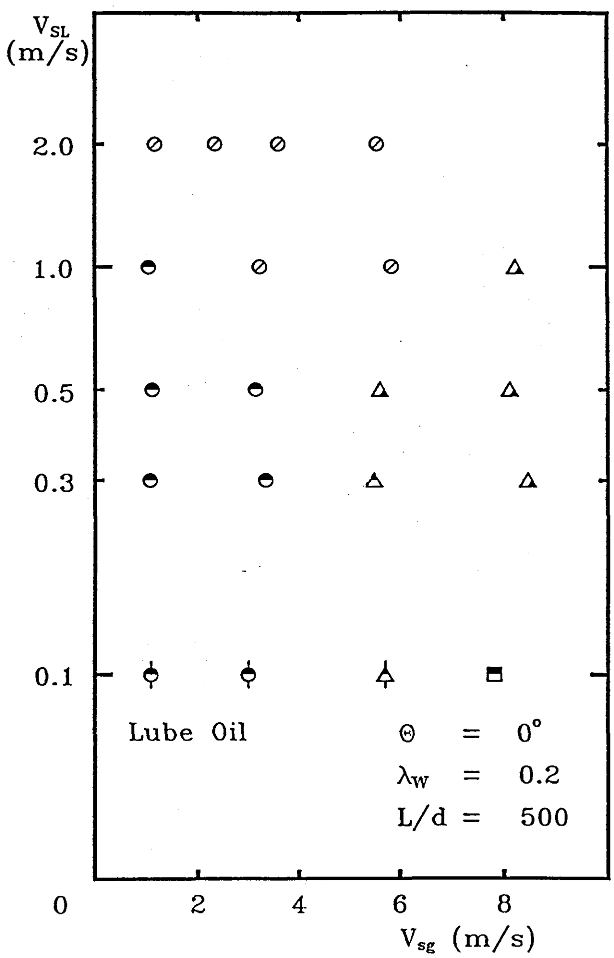


Fig 6.48 Oil/Water/Gas Flow Regime,L=500d,Oil No2

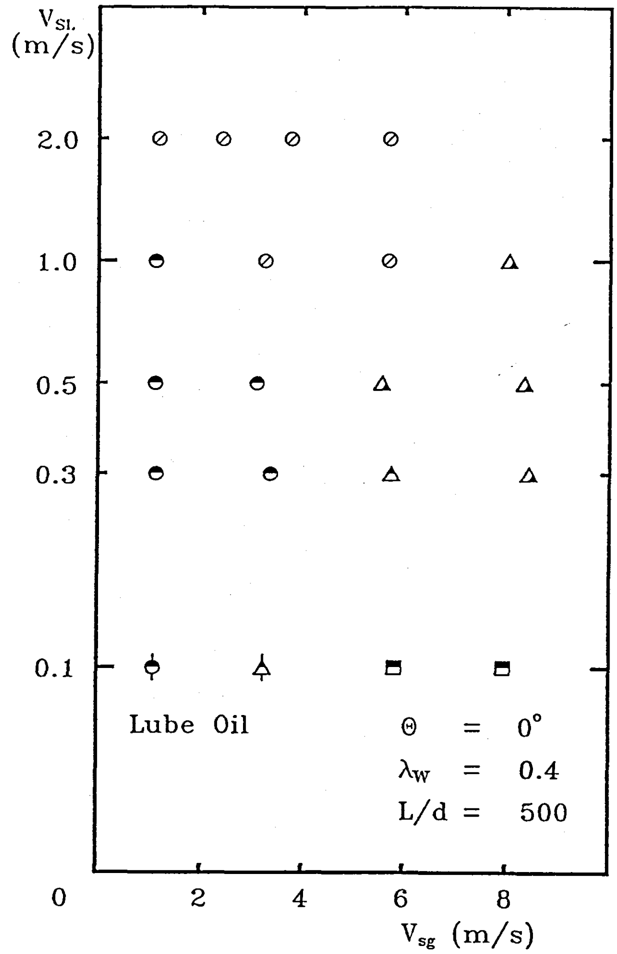


Fig 6.49 Oil/Water/Gas Flow Regime,L=500d,Oil No2

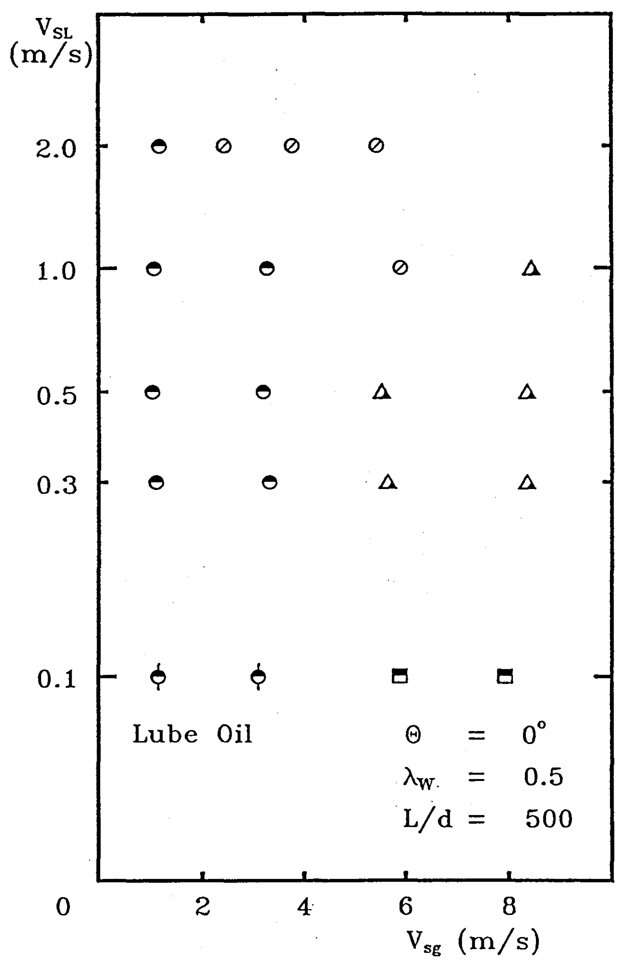


Fig 6.50 Oil/Water/Gas Flow Regime,L=500d,Oil No2

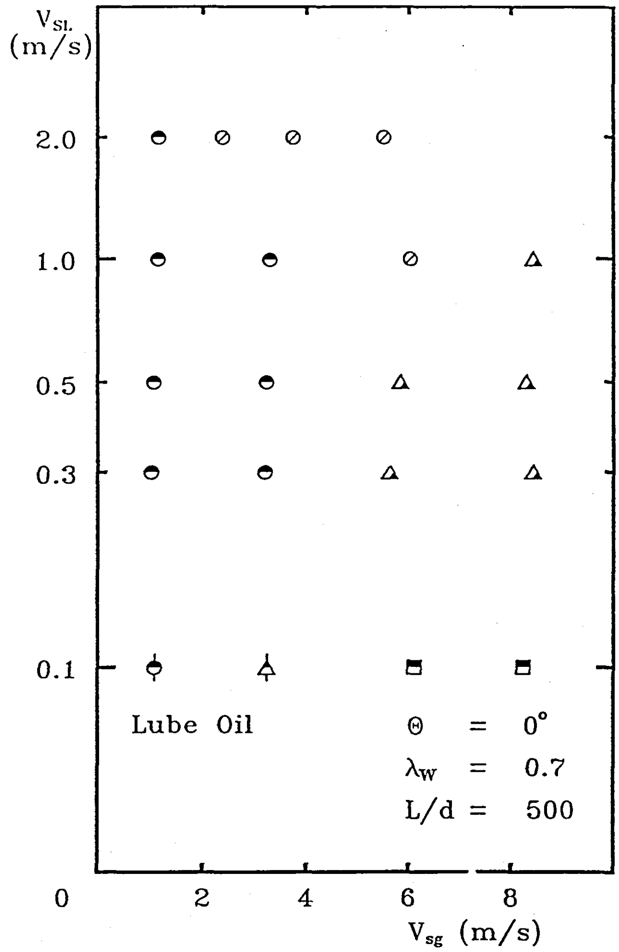


Fig 6.51 Oil/Water/Gas Flow Regime,L=500d,Oil No2

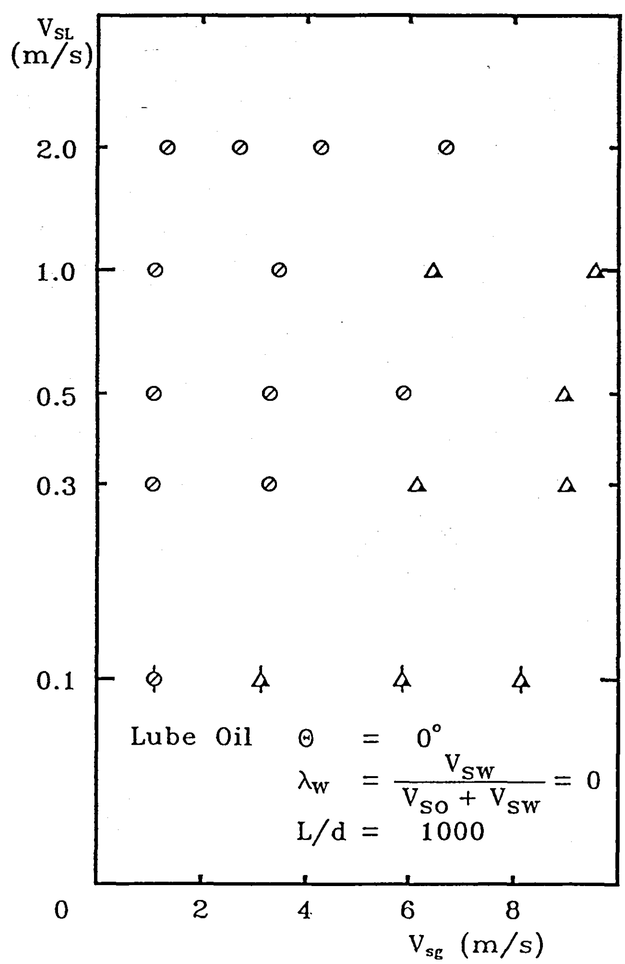


Fig 6.52 Oil/Water/Gas Flow Regime,L=1000d,Oil No2

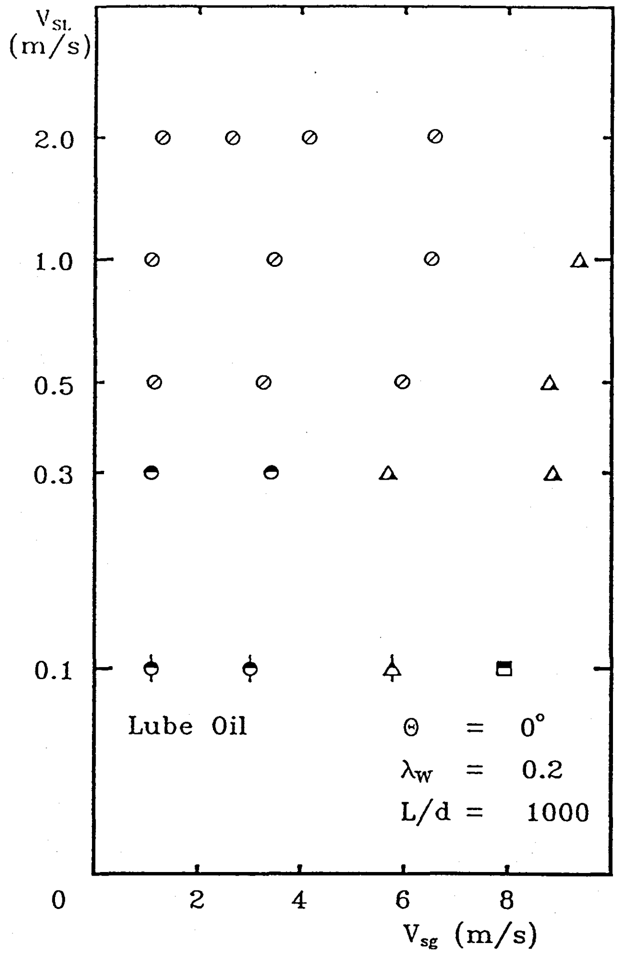


Fig 6.53 Oil/Water/Gas Flow Regime,L=1000d,Oil No2

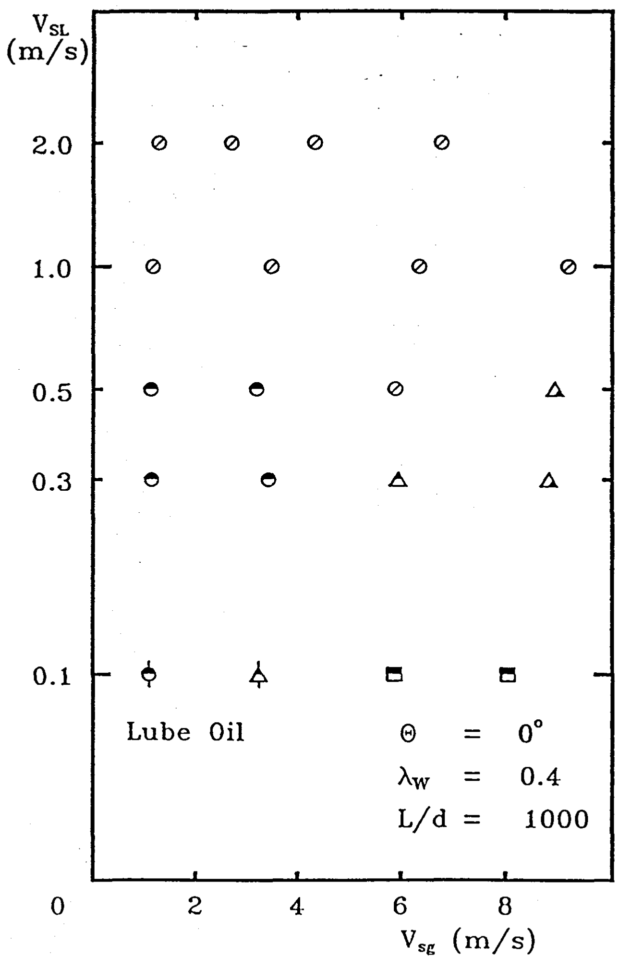


Fig 6.54 Oil/Water/Gas Flow Regime,L=1000d,Oil No2

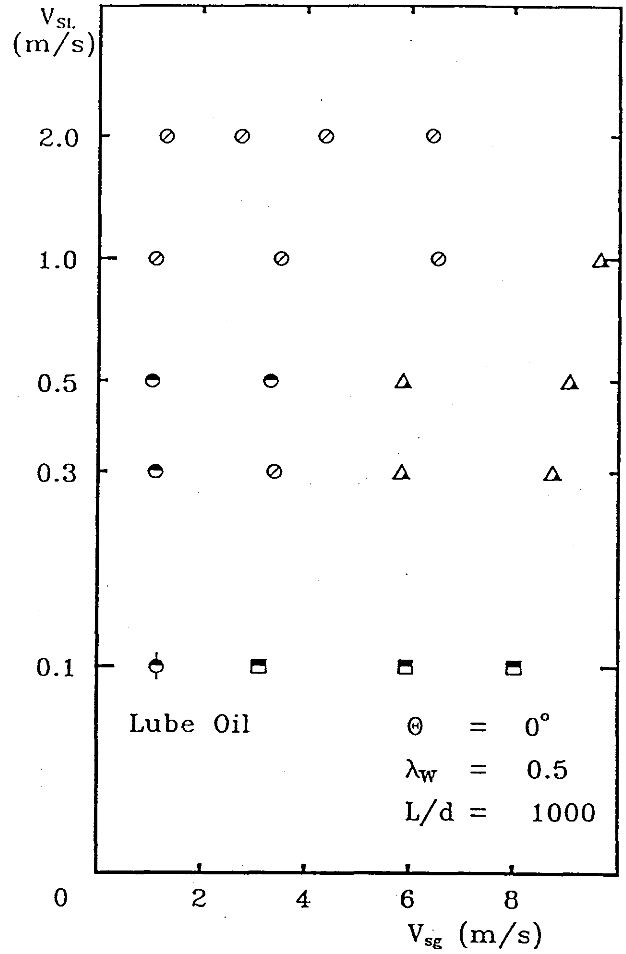


Fig 6.55 Oil/Water/Gas Flow Regime,L=1000d,Oil No2

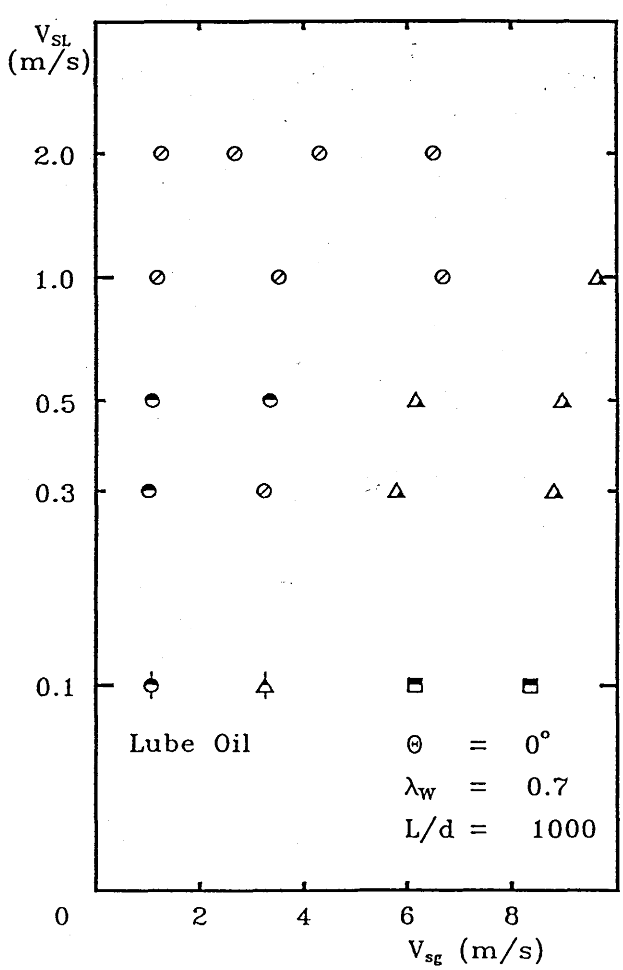
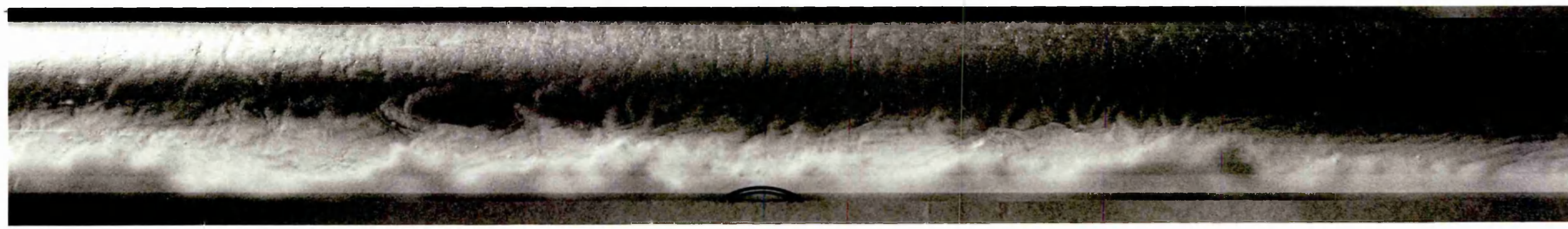


Fig 6.56 Oil/Water/Gas Flow Regime,L=1000d,Oil No2

(a) $V_{sL} = 0.1$ m/s, $\lambda_w = 0.5$, $V_{sg} = 1$ m/s (Flow R to L)

(b) $V_{sL} = 0.5$ m/s, $\lambda_w = 0.7$, $V_{sg} = 1.1$ m/s (Flow R to L)

Fig 6.57 Oil/Water/Gas Flow Regime Photographs, L=1000d, Oil No2



(c) $V_{sL} = 0.5$ m/s, $\lambda_w = 0.7$, $V_{sg} = 9$ m/s (Flow R to L)

Fig 6.57 Oil/Water/Gas Flow Regime Photographs, L=1000d, Oil No2

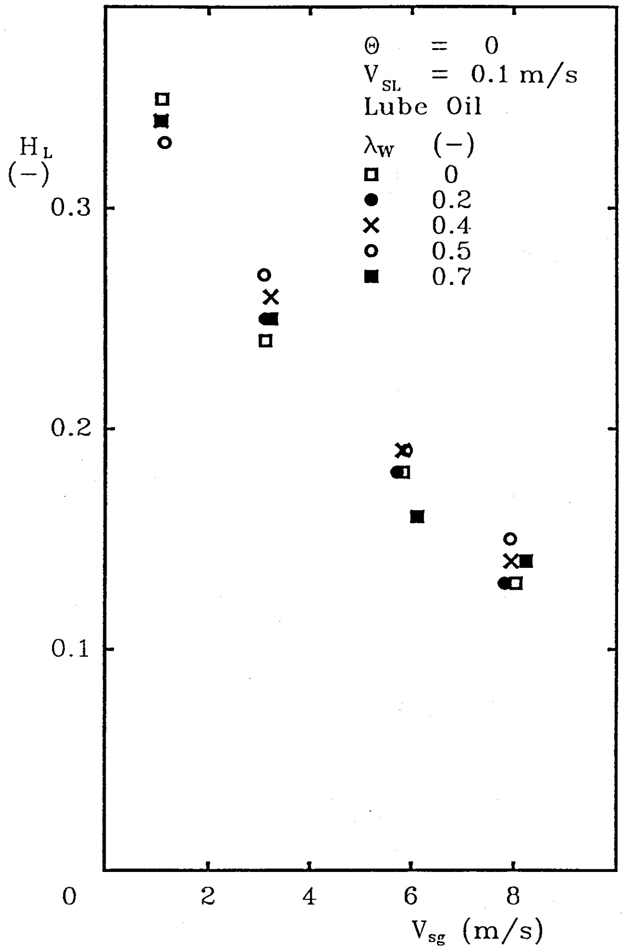


Fig 6.58 Oil/Water/Gas Total Liquids Holdup,Oil No2

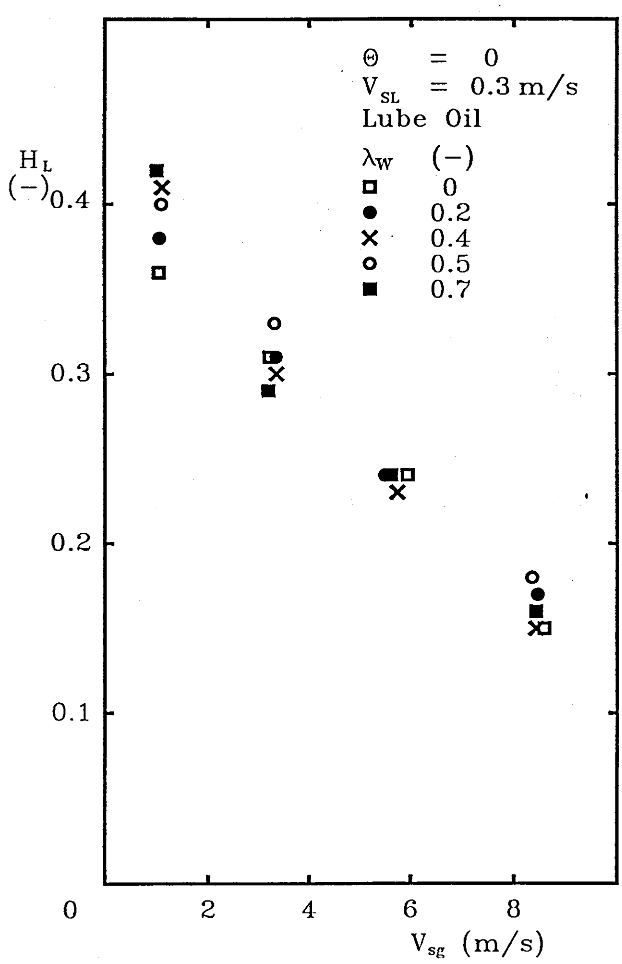


Fig 6.59 Oil/Water/Gas Total Liquids Holdup,Oil No2

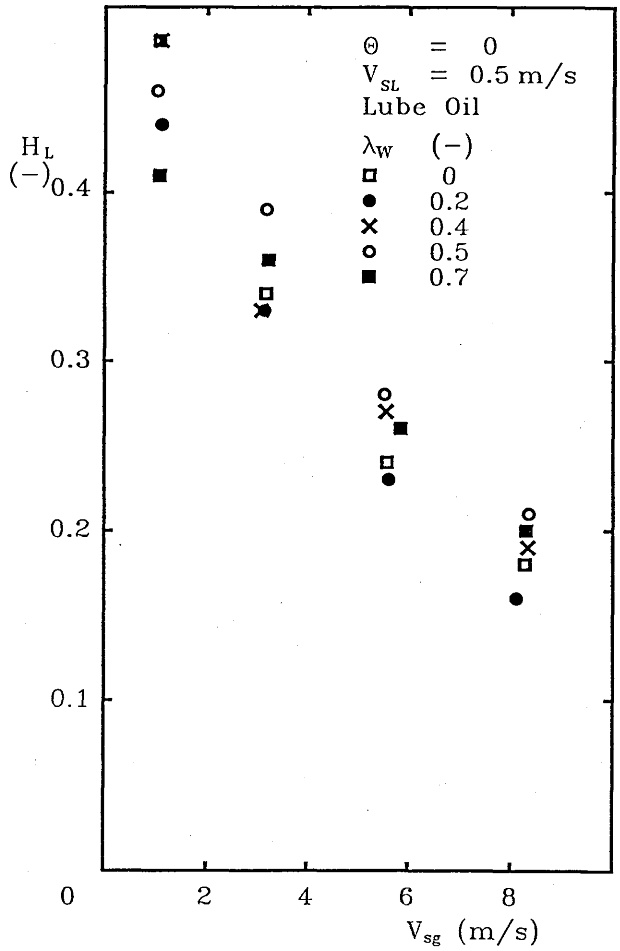


Fig 6.60 Oil/Water/Gas Total Liquids Holdup,Oil No2

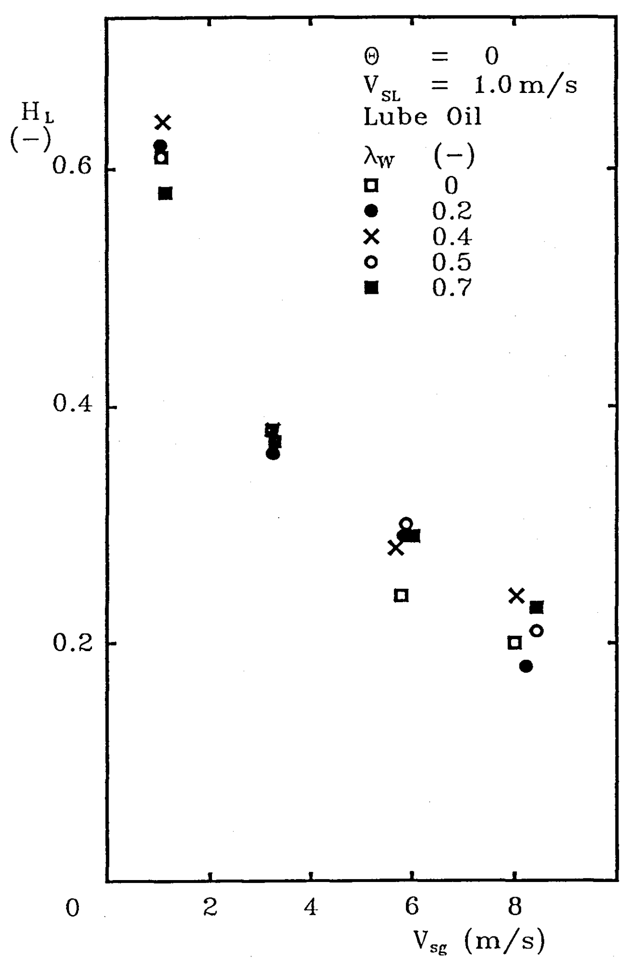


Fig 6.61 Oil/Water/Gas Total Liquids Holdup,Oil No2

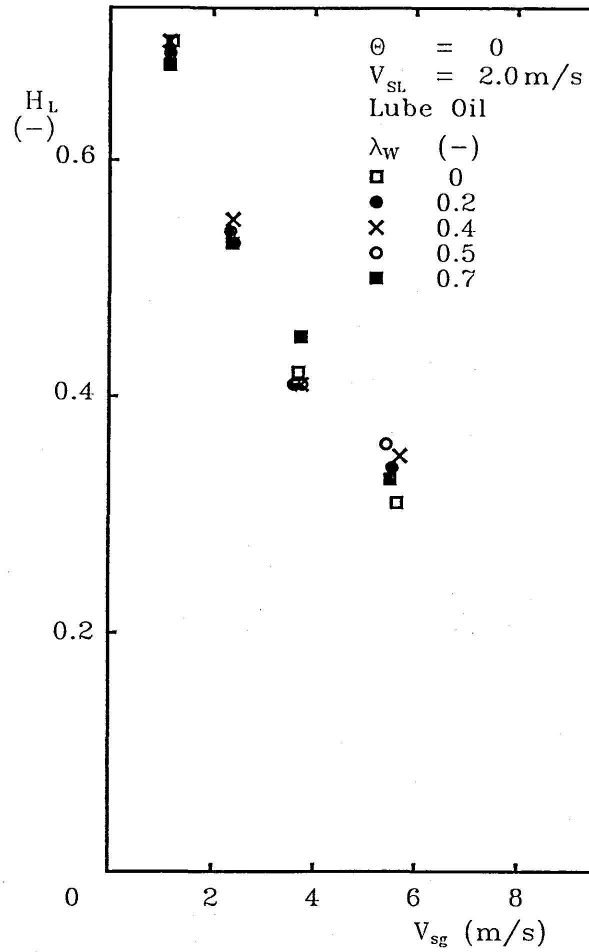


Fig 6.62 Oil/Water/Gas Total Liquids Holdup,Oil No2

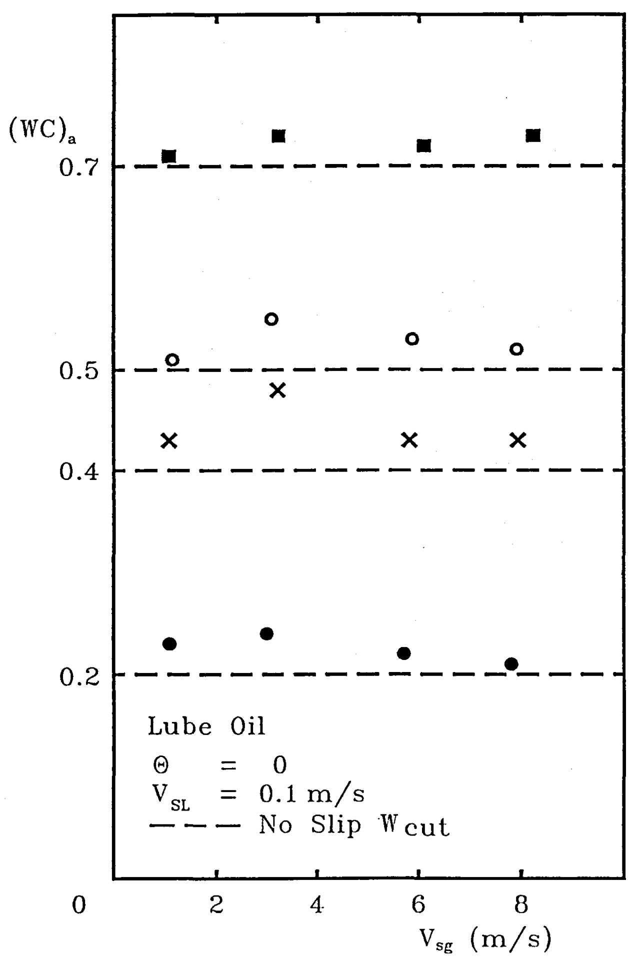


Fig 6.63 Oil/Water/Gas In-situ Water Fraction,Oil No2

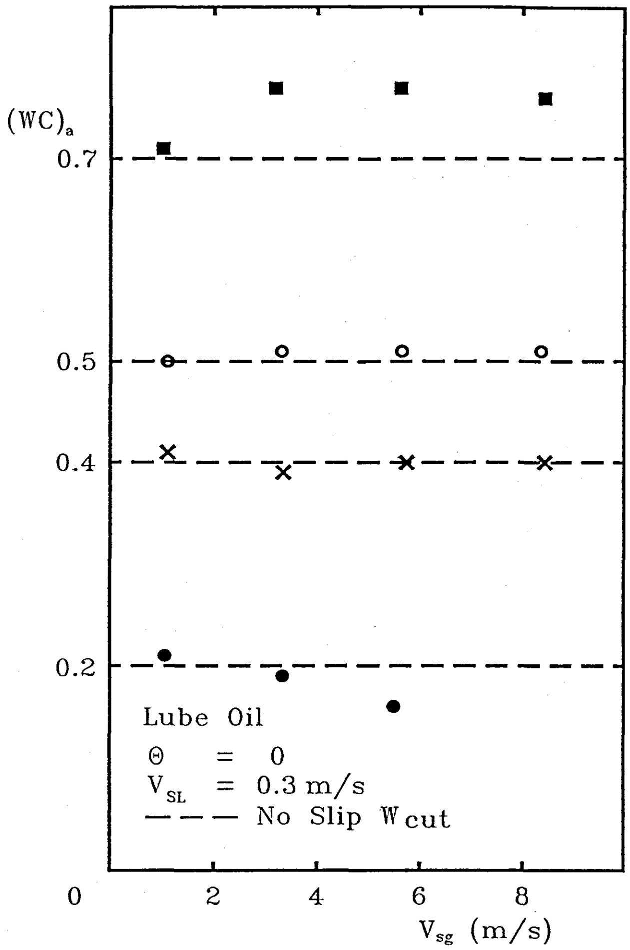


Fig 6.64 Oil/Water/Gas In-situ Water Fraction,Oil No2

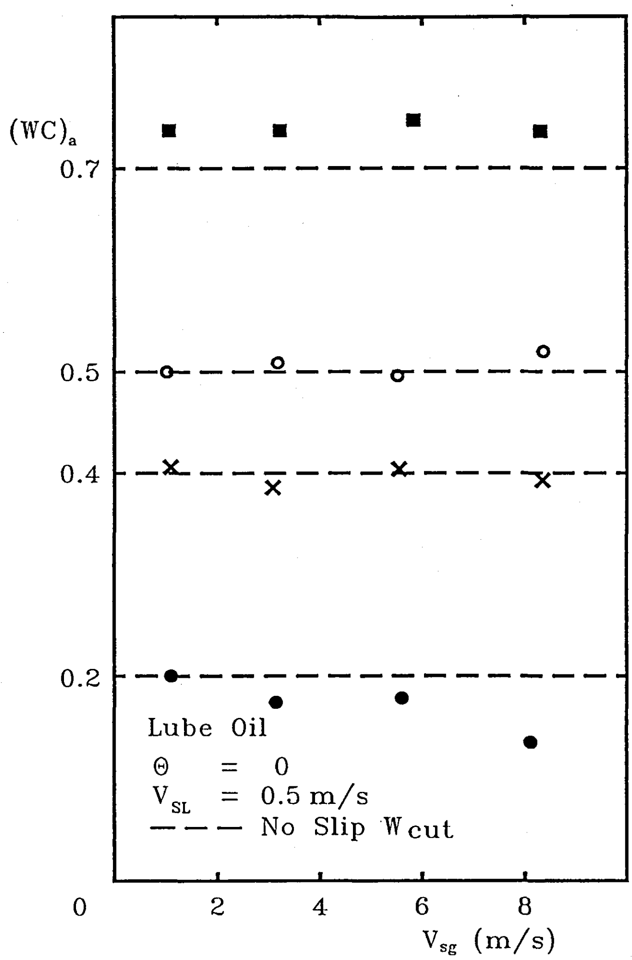


Fig 6.65 Oil/Water/Gas In-situ Water Fraction,Oil No2

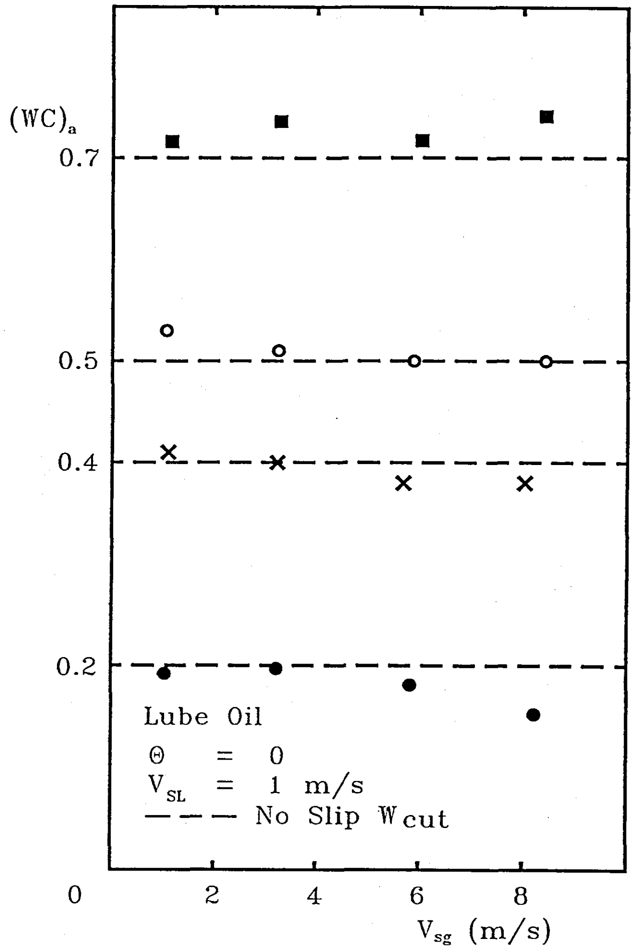


Fig 6.66 Oil/Water/Gas In-situ Water Fraction,Oil No2

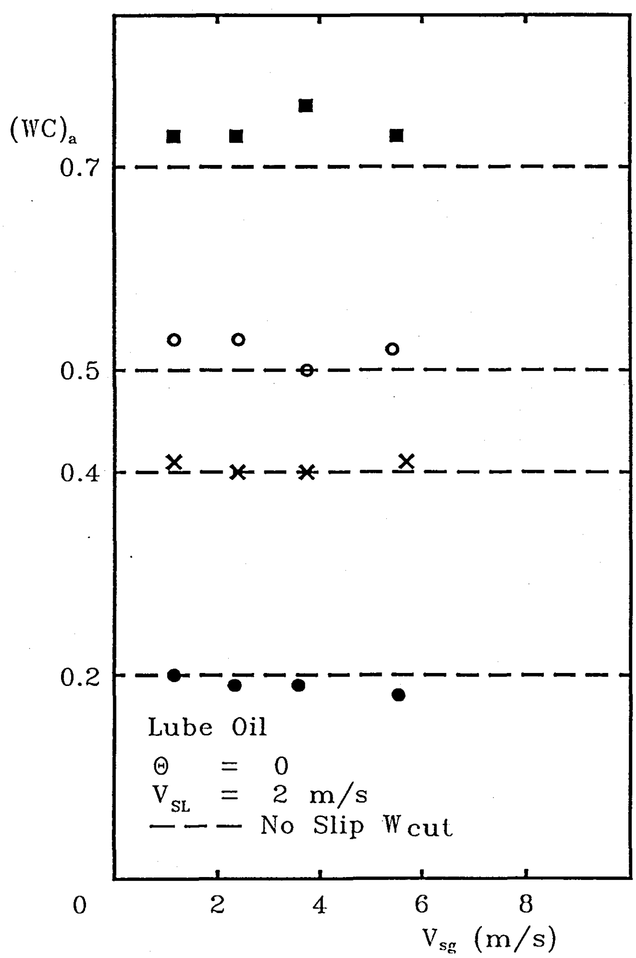


Fig 6.67 Oil/Water/Gas In-situ Water Fraction,Oil No2

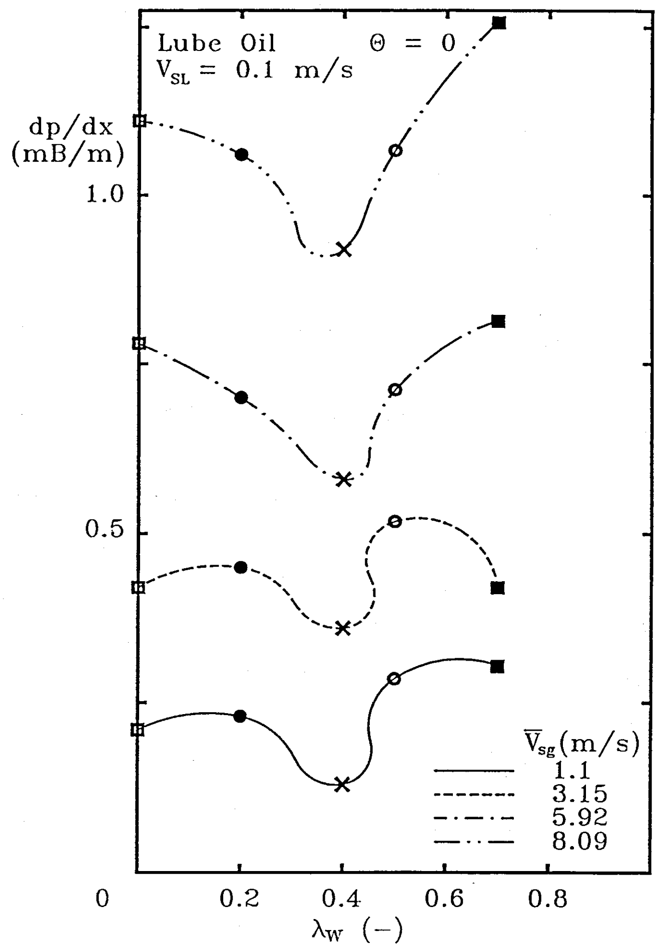


Fig 6.68 Oil/Water/Gas Pressure Loss,Oil No2

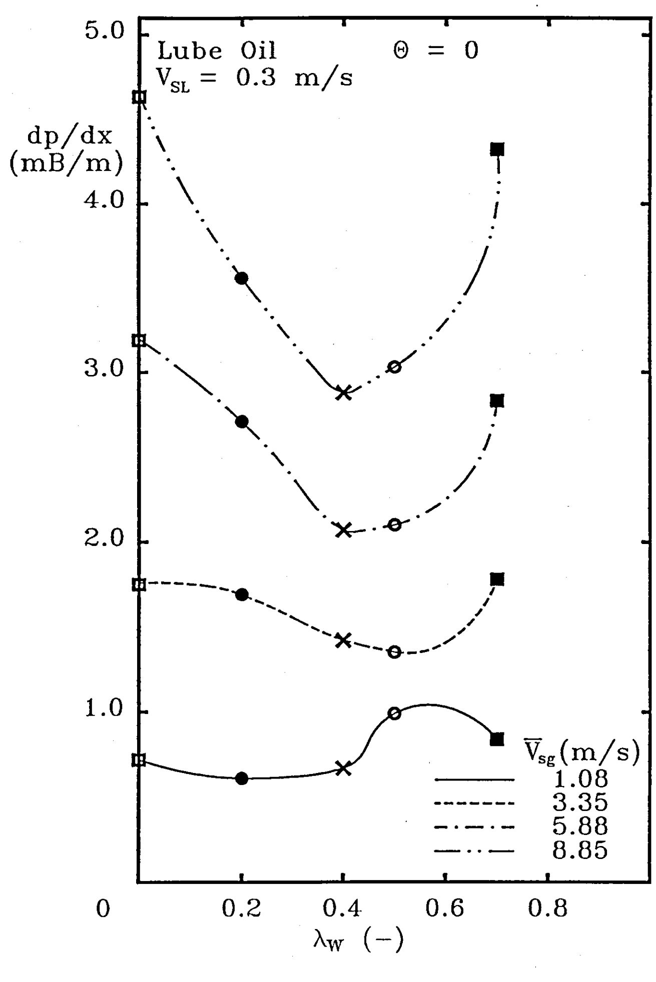


Fig 6.69 Oil/Water/Gas Pressure Loss,Oil No2

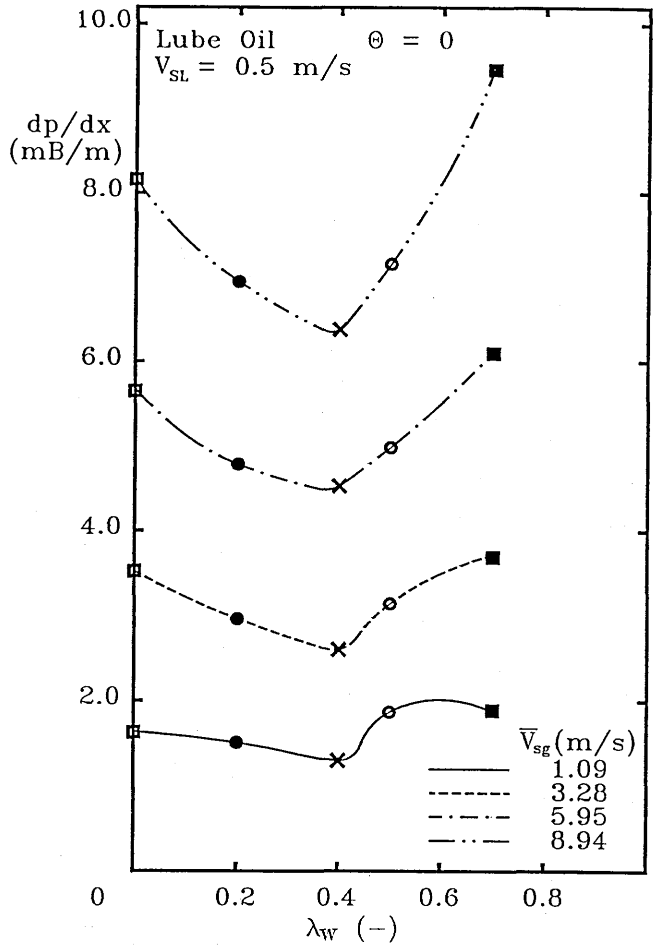


Fig 6.70 Oil/Water/Gas Pressure Loss,Oil No2

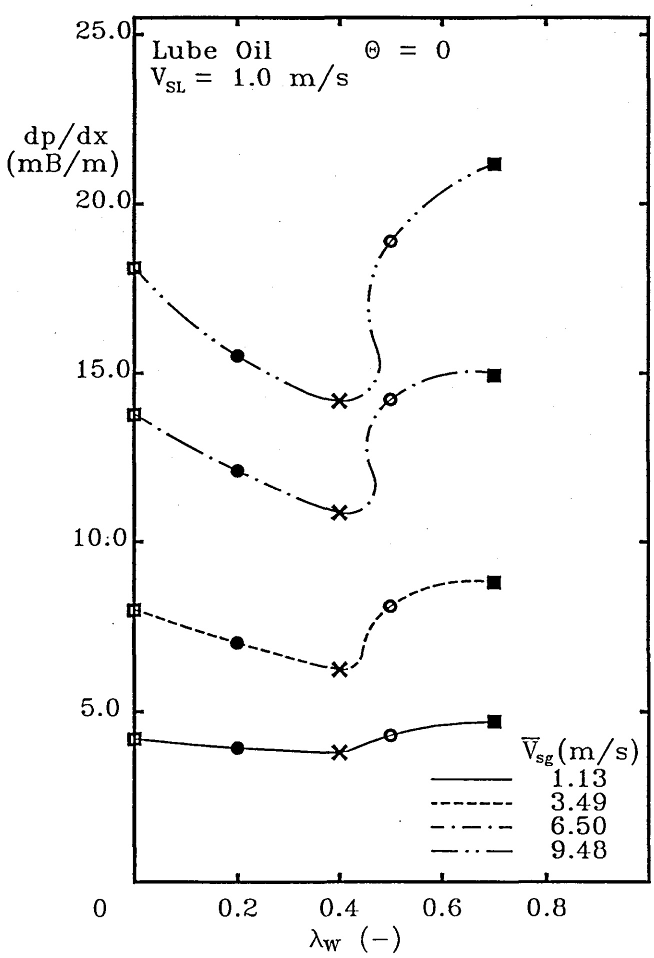


Fig 6.71 Oil/Water/Gas Pressure Loss,Oil No2

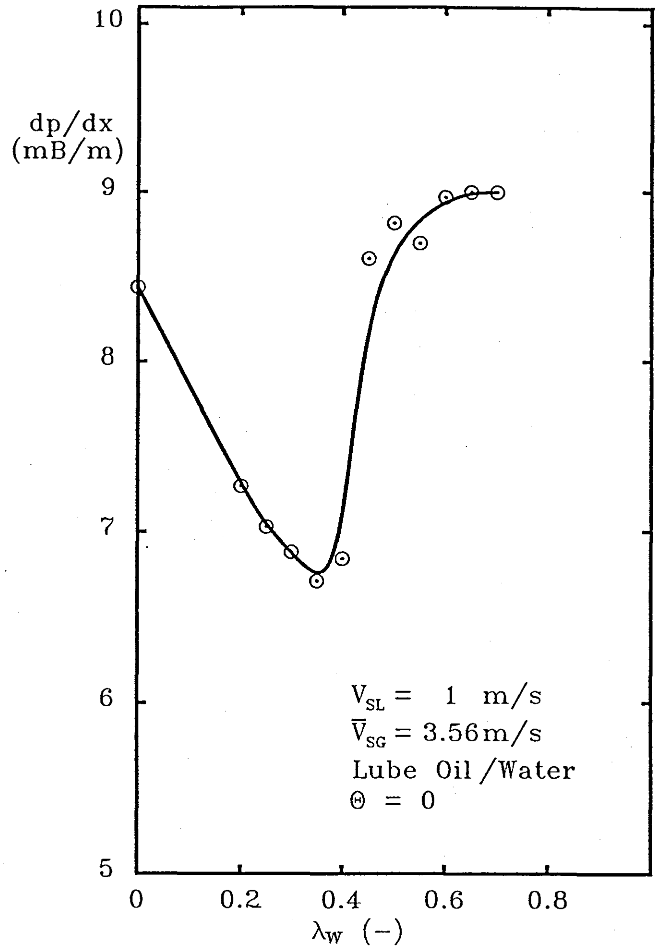


Fig 6.72 Oil/Water/Gas Pressure Loss,Oil No2

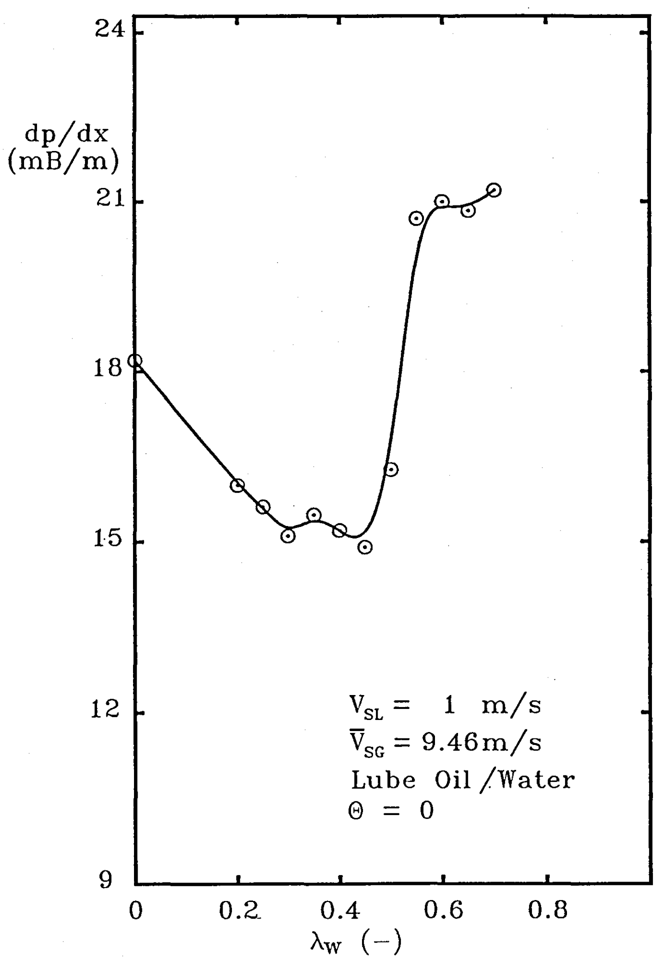


Fig 6.73 Oil/Water/Gas Pressure Loss,Oil No2

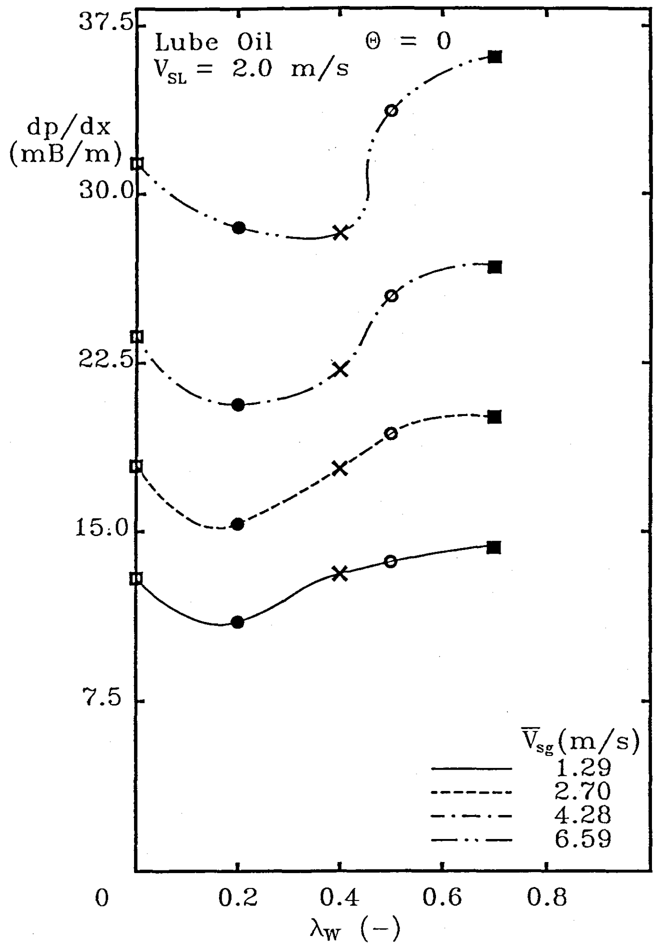


Fig 6.74 Oil/Water/Gas Pressure Loss,Oil No2

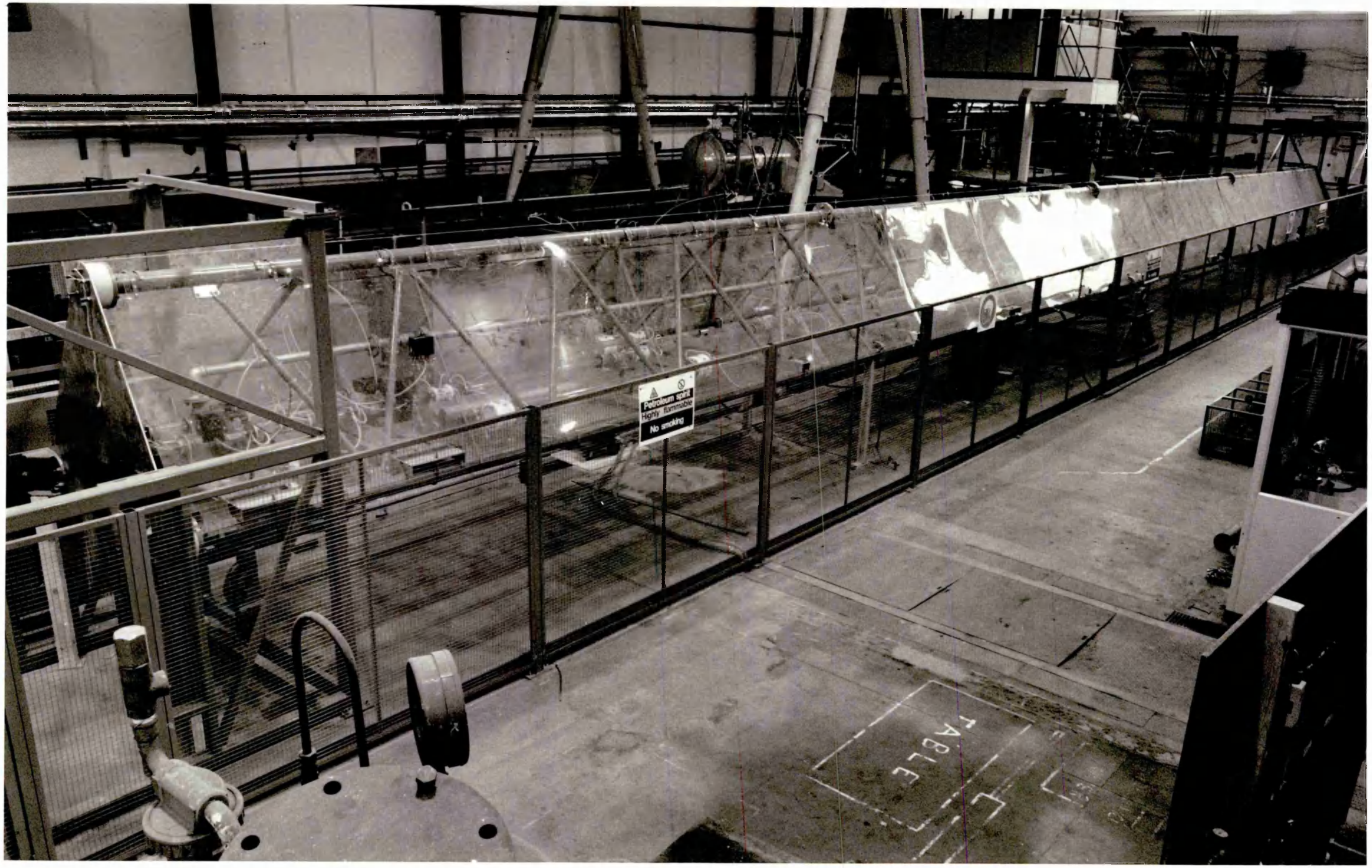


Fig 7.1 Pipe-bridge in Uphill-Downhill Configuration

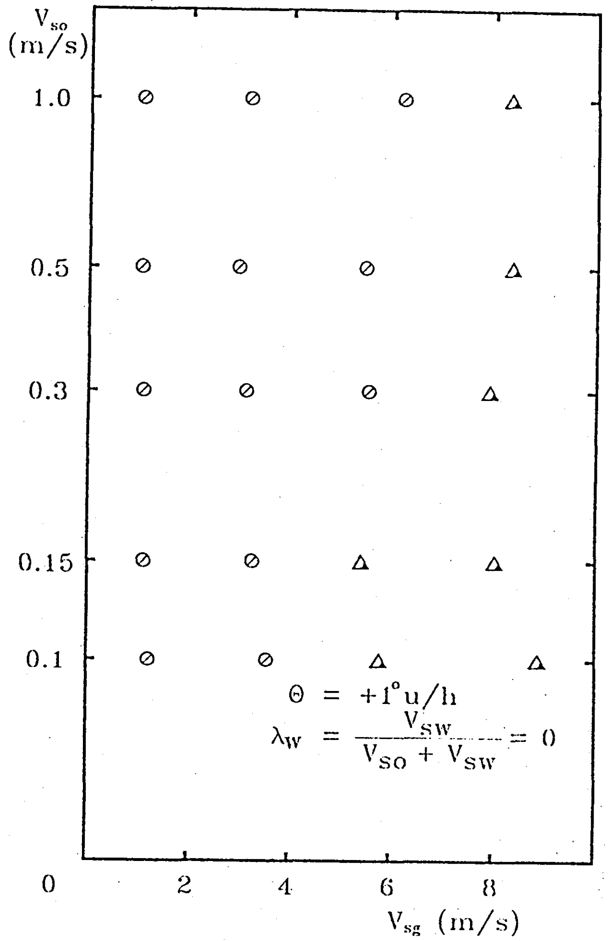


Fig 7.2 Oil/Water/Gas Flow Regime,1 deg u/h,Oil No1

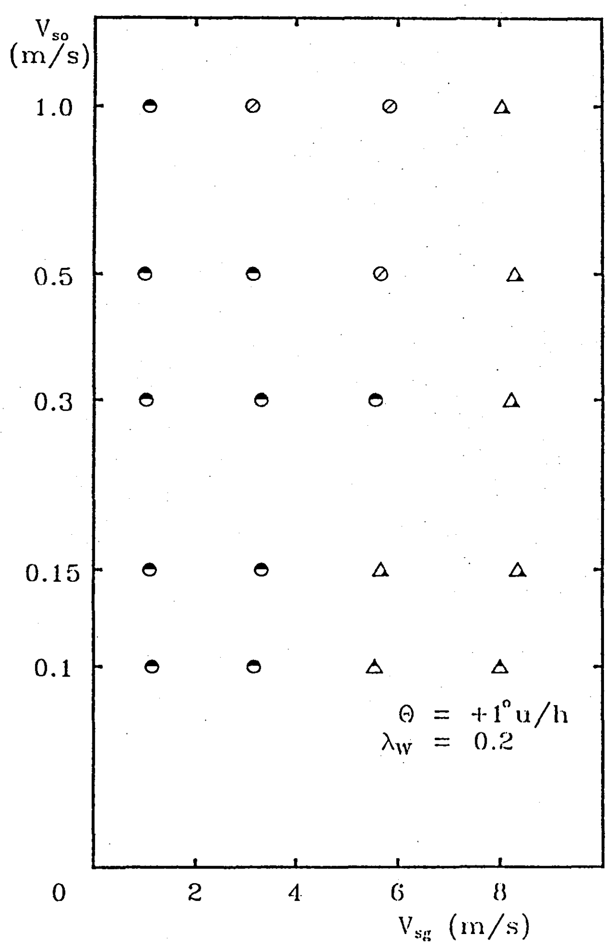


Fig 7.3 Oil/Water/Gas Flow Regime,1 deg u/h,Oil No1

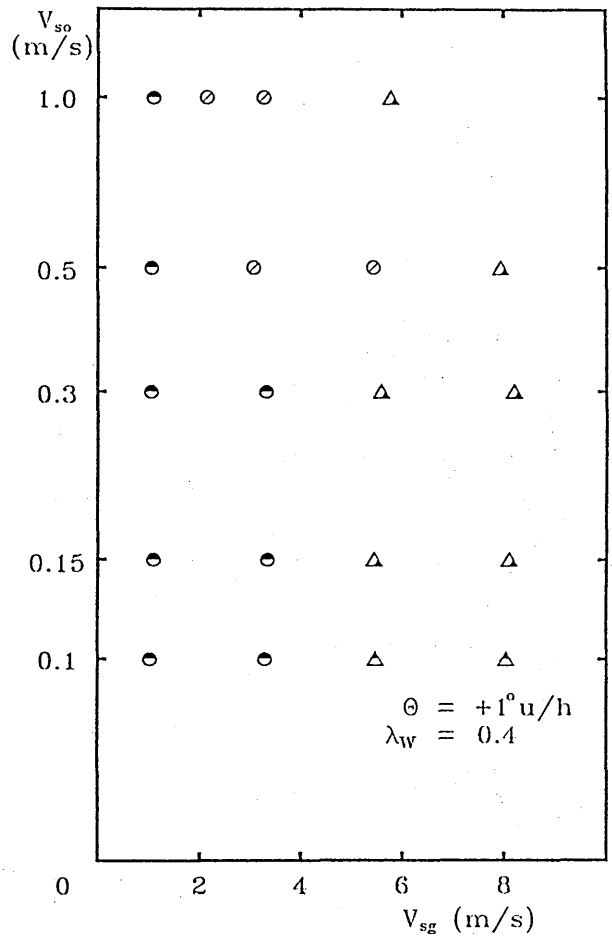


Fig 7.4 Oil/Water/Gas Flow Regime,1 deg u/h,Oil No1

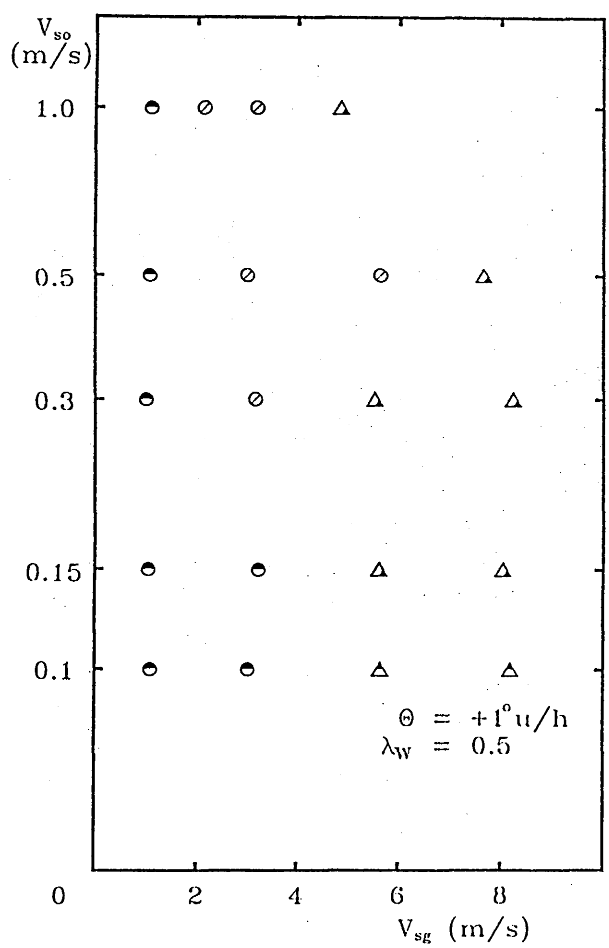


Fig 7.5 Oil/Water/Gas Flow Regime,1 deg u/h,Oil No1

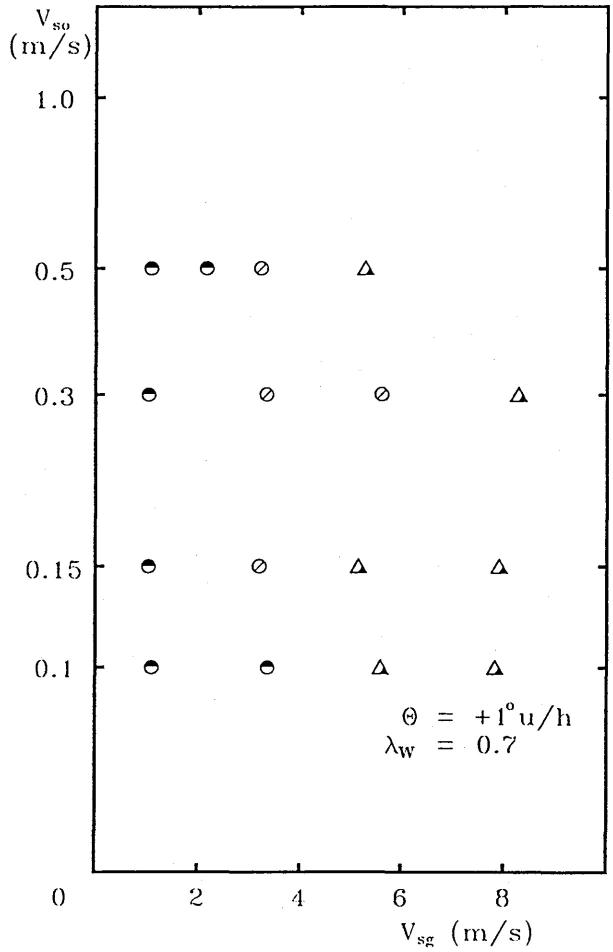


Fig 7.6 Oil/Water/Gas Flow Regime,1 deg u/h,Oil No1

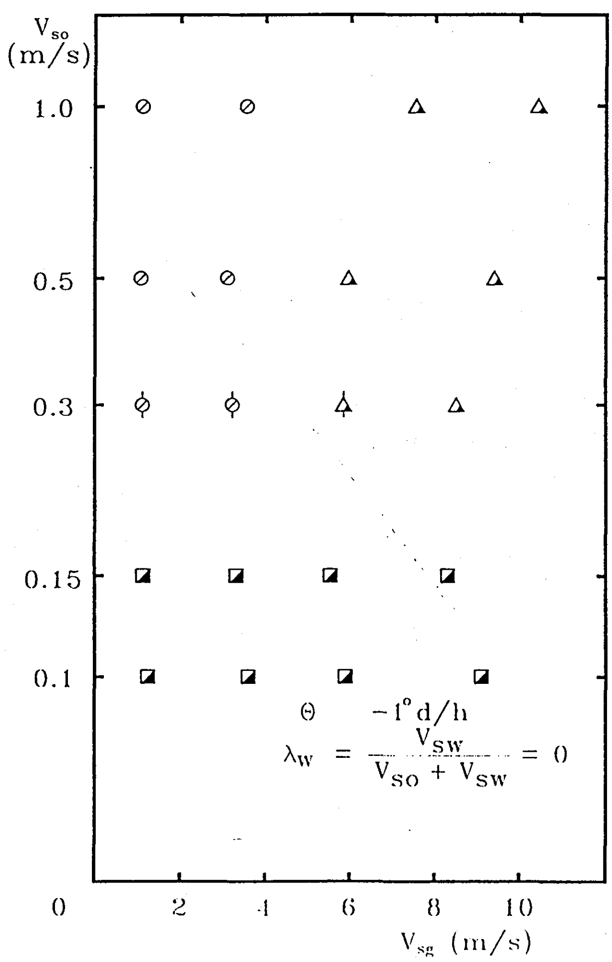


Fig 7.7 Oil/Water/Gas Flow Regime,1 deg d/h,Oil No1

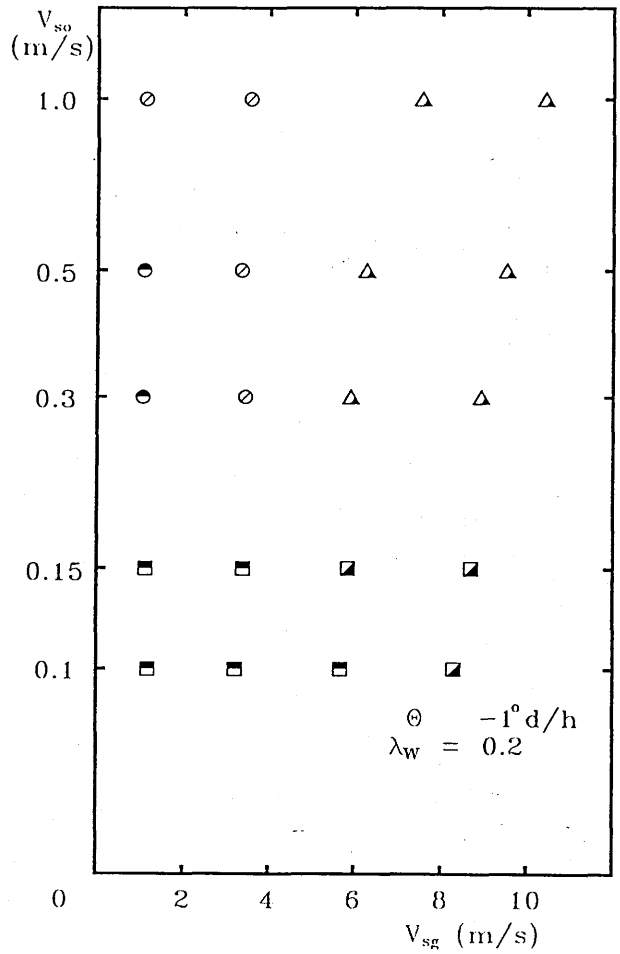


Fig 7.8 Oil/Water/Gas Flow Regime,1 deg d/h,Oil No1

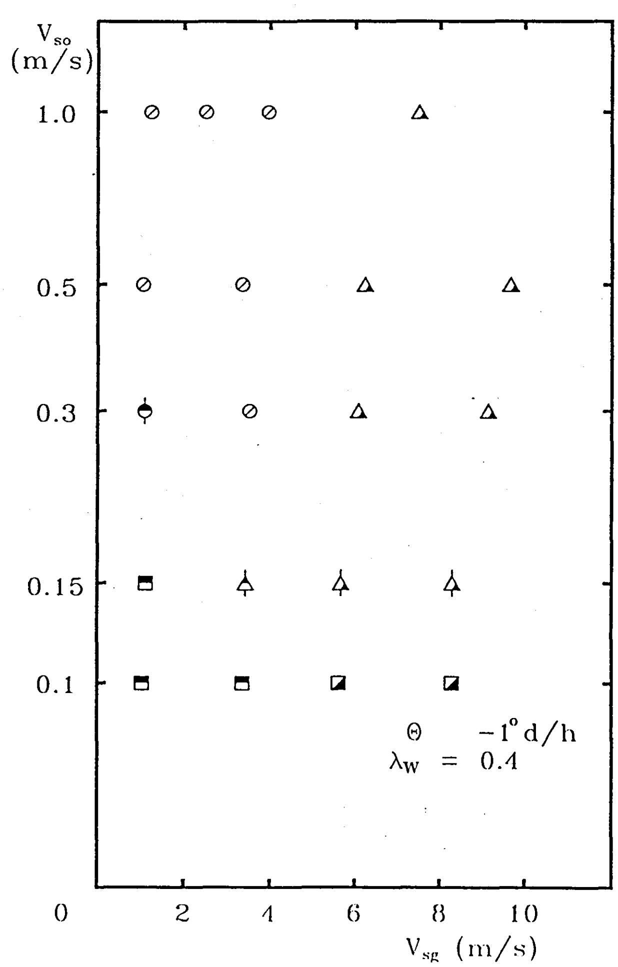


Fig 7.9 Oil/Water/Gas Flow Regime,1 deg d/h,Oil No1

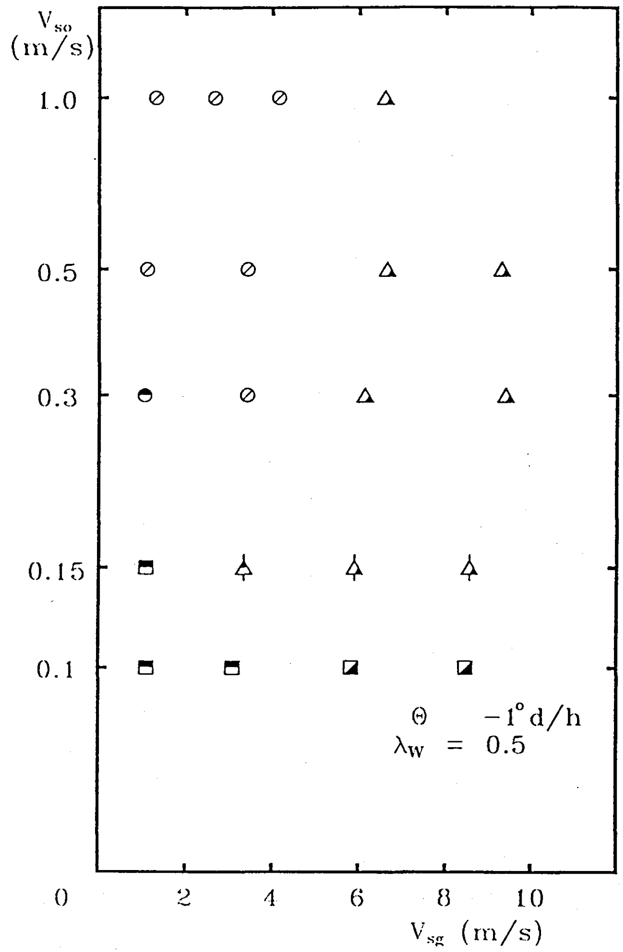


Fig 7.10 Oil/Water/Gas Flow Regime,1 deg d/h,Oil No1

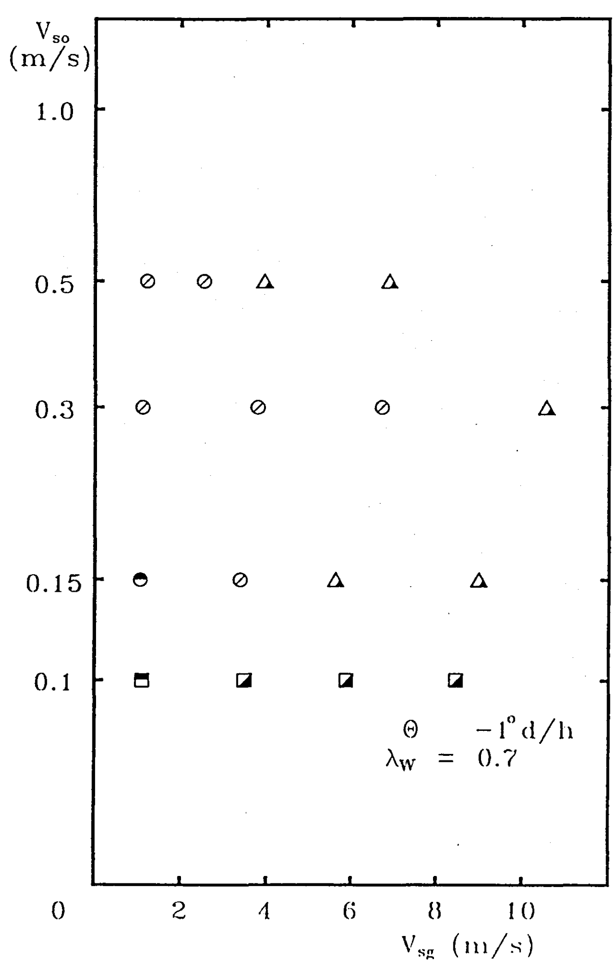


Fig 7.11 Oil/Water/Gas Flow Regime,1 deg d/h,Oil No1

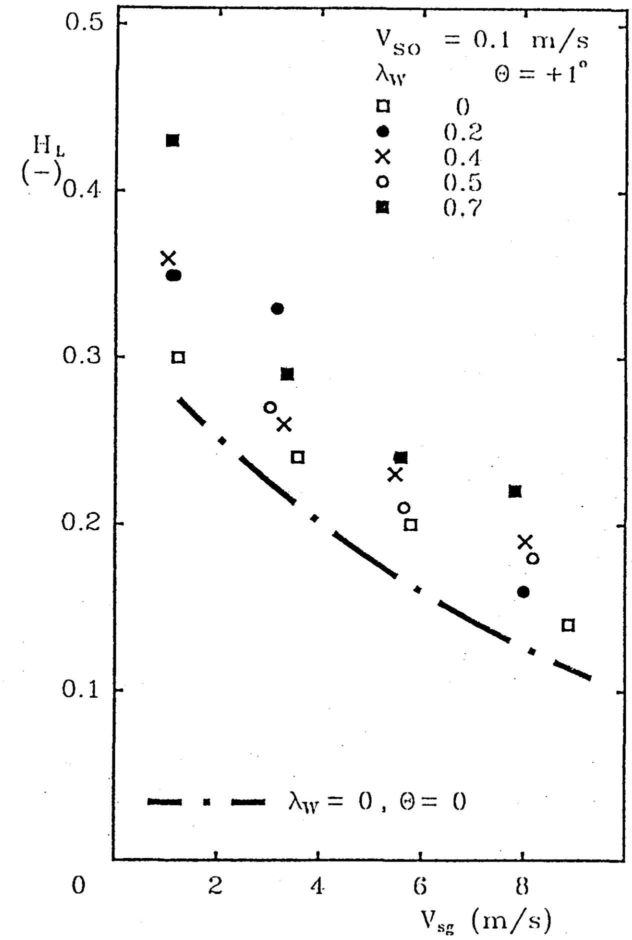


Fig 7.12 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

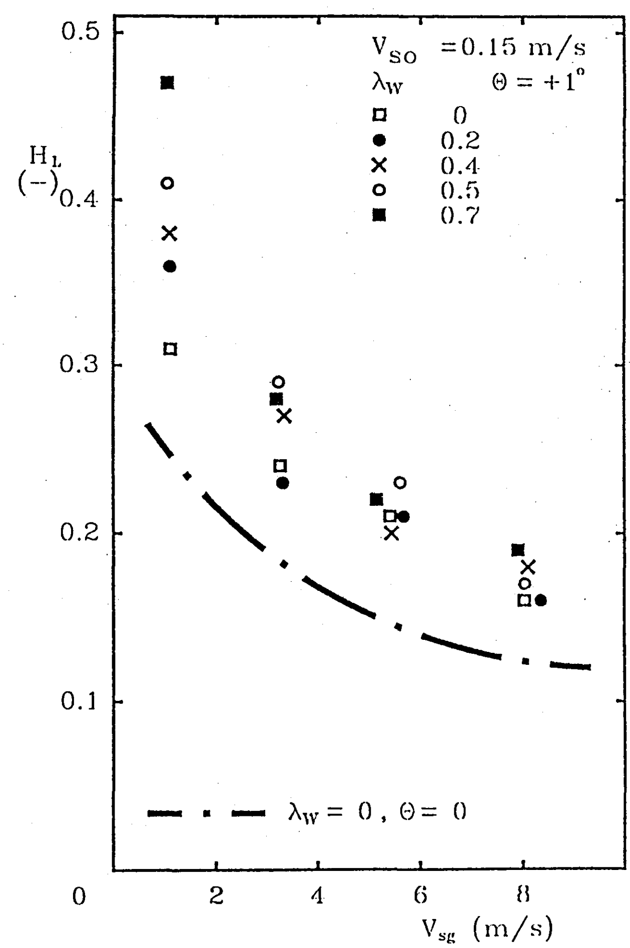


Fig 7.13 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

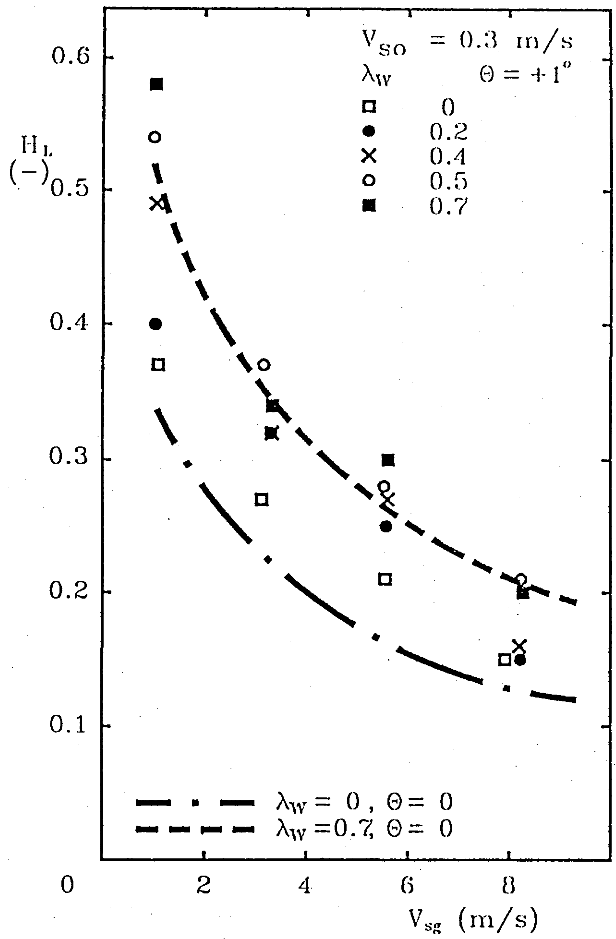


Fig 7.14 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

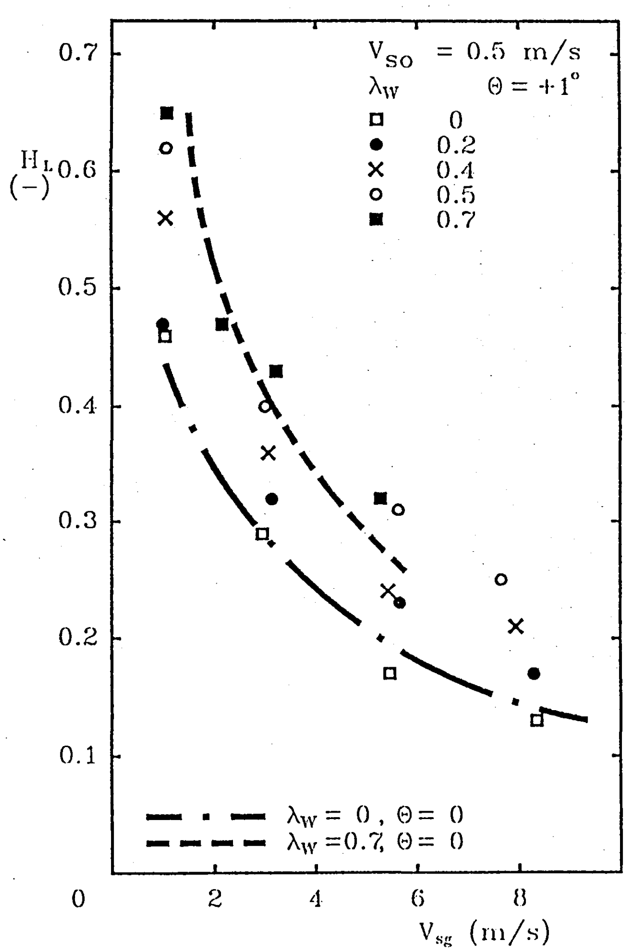


Fig 7.15 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

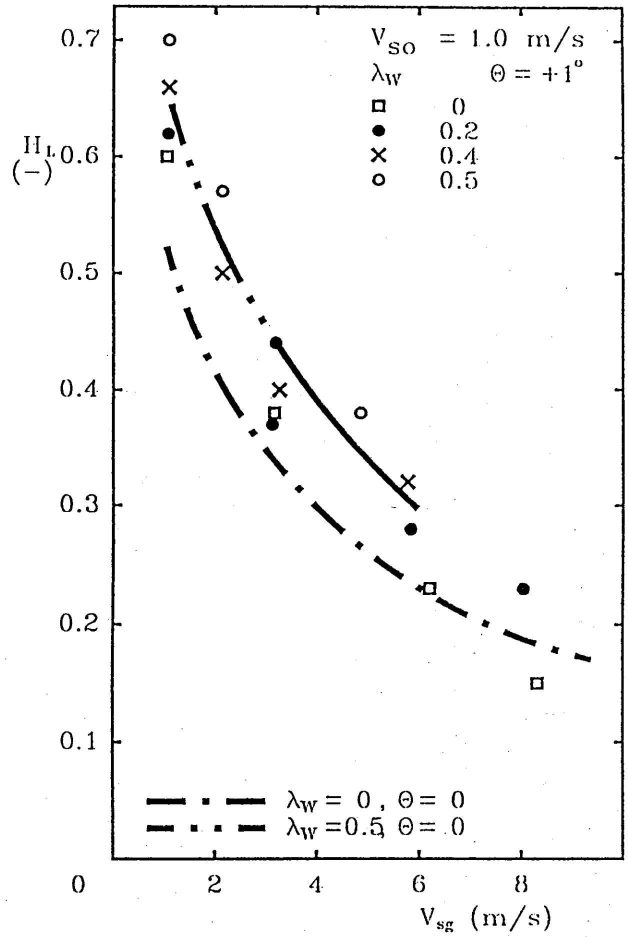


Fig 7.16 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

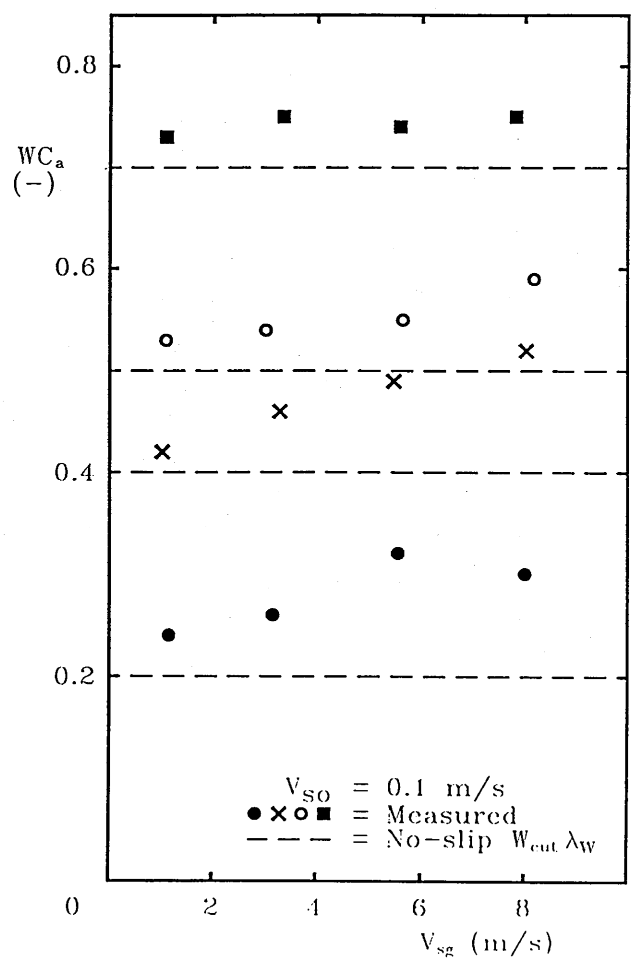


Fig 7.17 Oil/Water/Gas In-situ Water Fractions,1 deg u/h,Oil No1

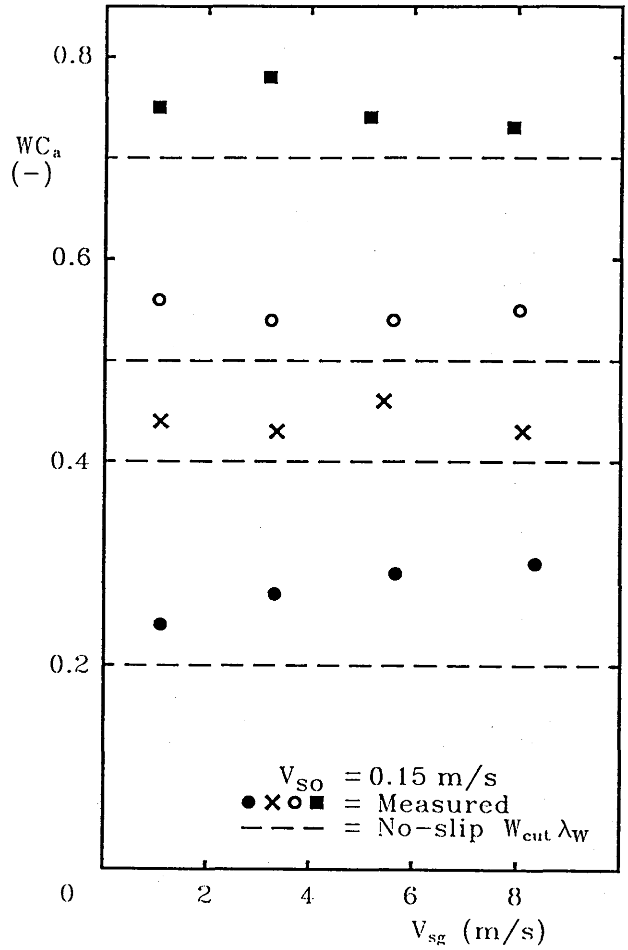


Fig 7.18 Oil/Water/Gas In-situ Water Fractions,1 deg u/h,Oil No1

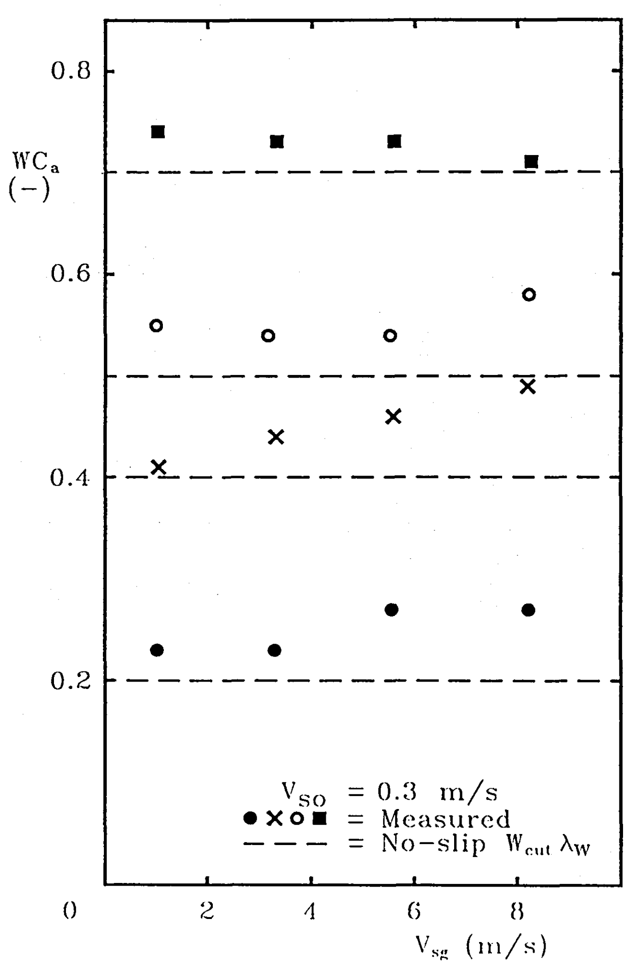


Fig 7.19 Oil/Water/Gas In-situ Water Fractions,1 deg u/h,Oil No1

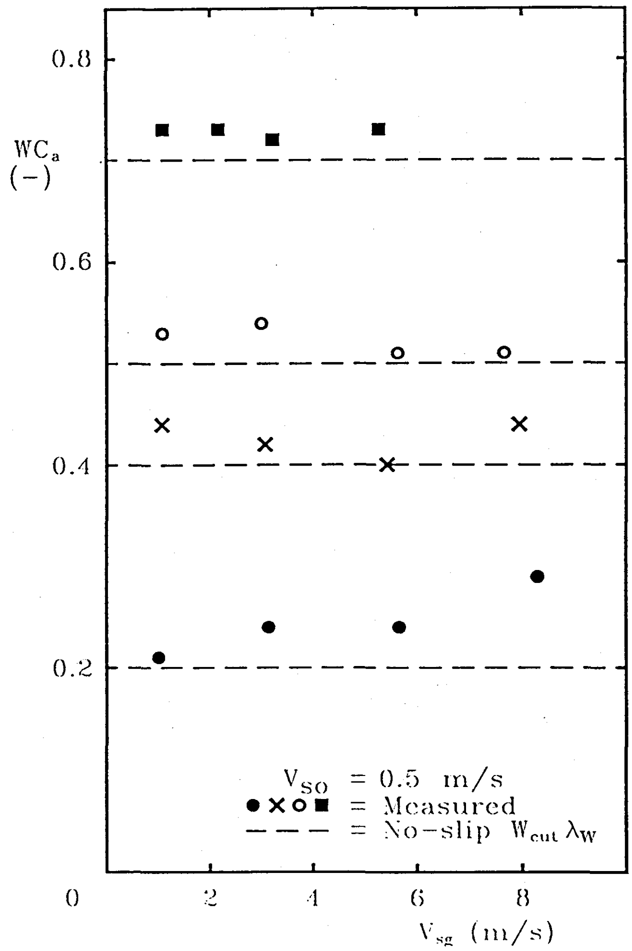


Fig 7.20 Oil/Water/Gas In-situ Water Fractions,1 deg u/h,Oil No1

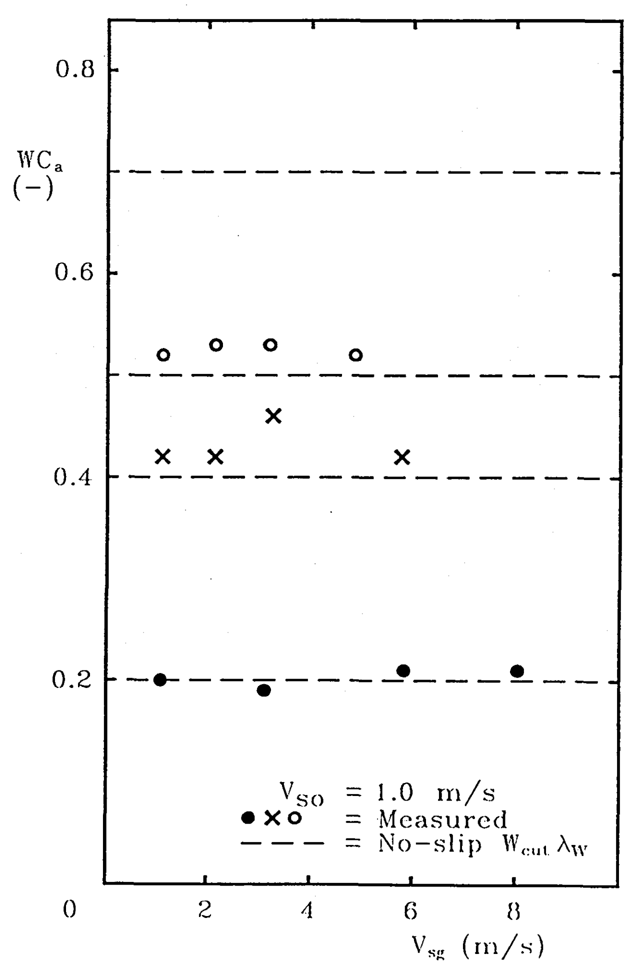


Fig 7.21 Oil/Water/Gas In-situ Water Fractions,1 deg u/h,Oil No1

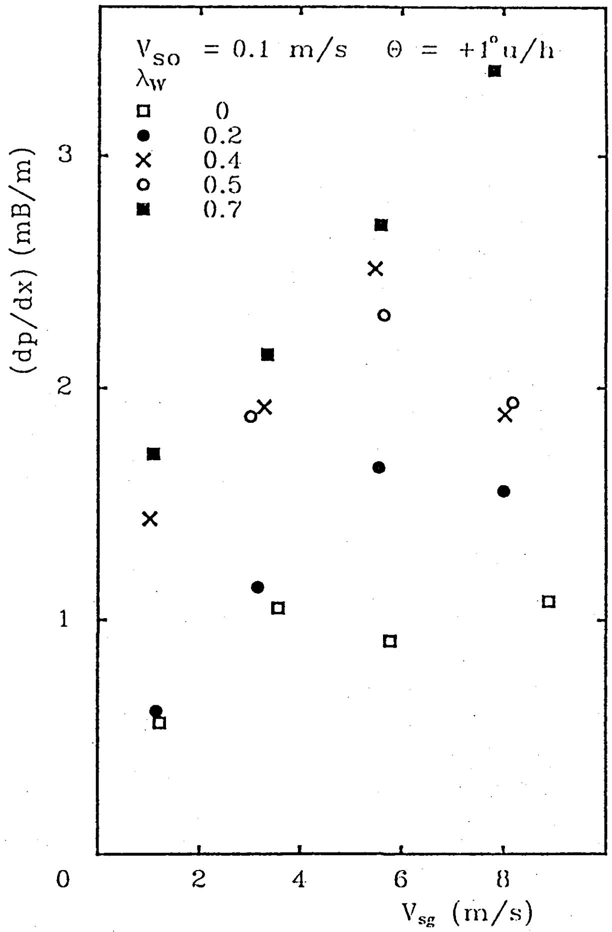


Fig 7.22 Oil/Water/Gas Pressure Loss,1 deg u/h,Oil No1

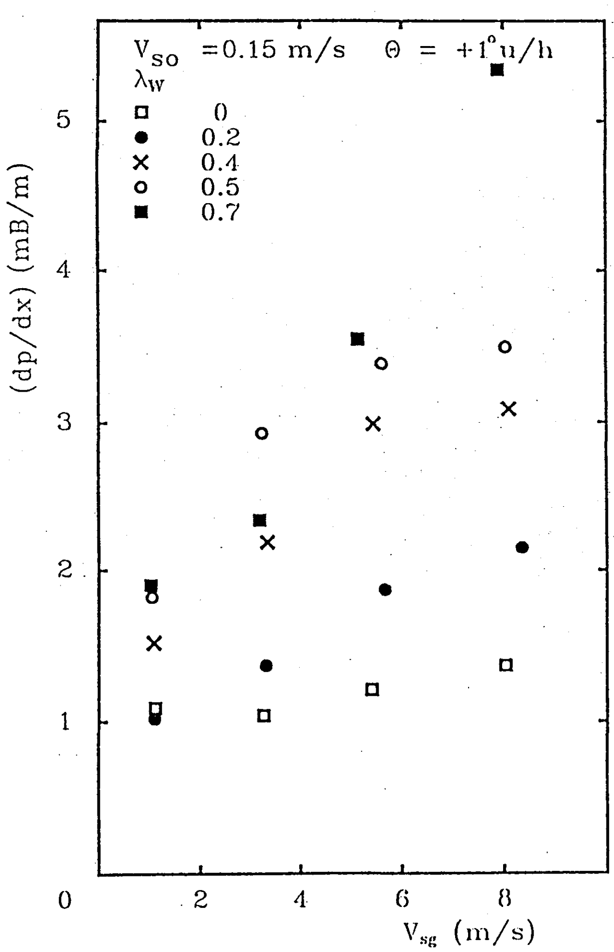


Fig 7.23 Oil/Water/Gas Pressure Loss,1 deg u/h,Oil No1

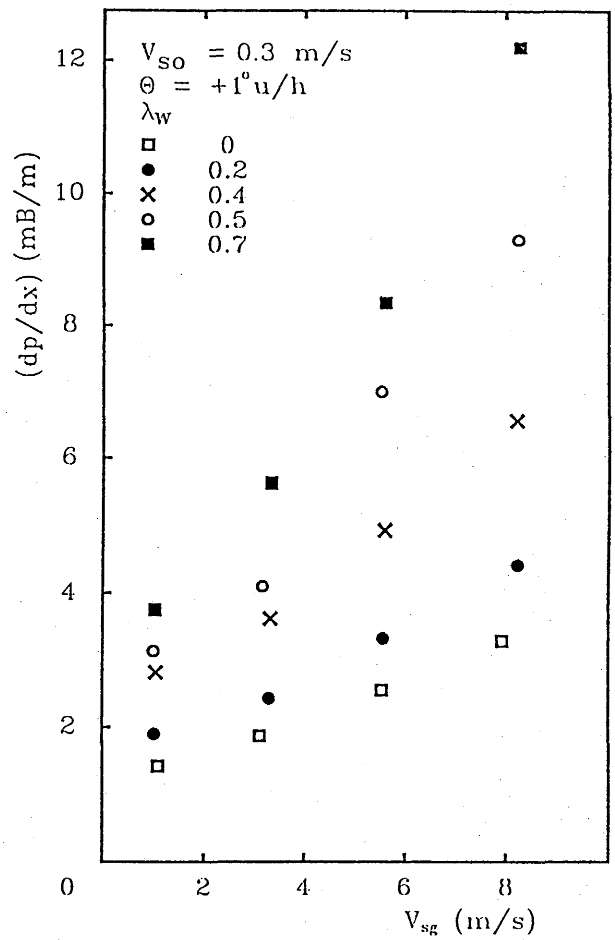


Fig 7.24 Oil/Water/Gas Pressure Loss,1 deg u/h,Oil No1

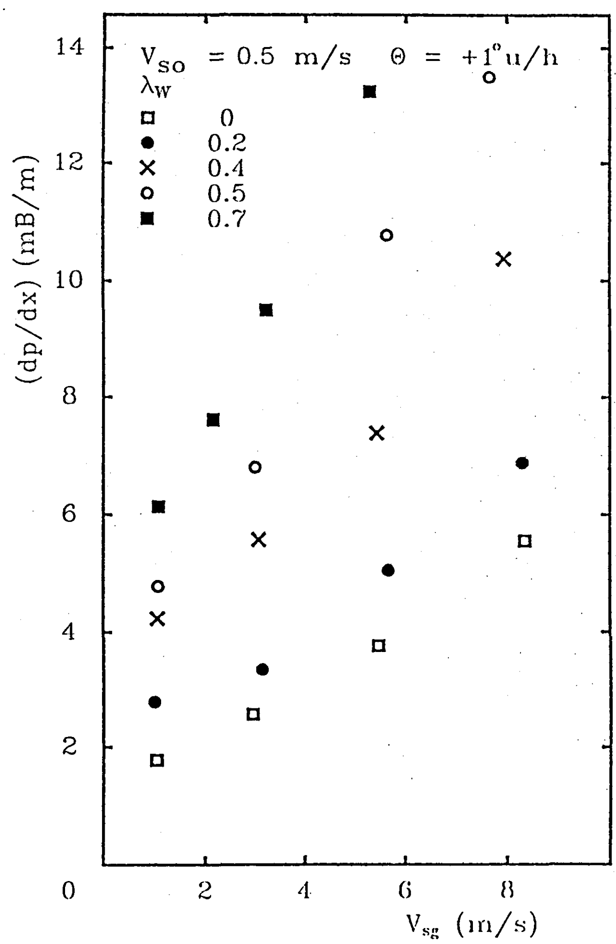


Fig 7.25 Oil/Water/Gas Pressure Loss,1 deg u/h,Oil No1

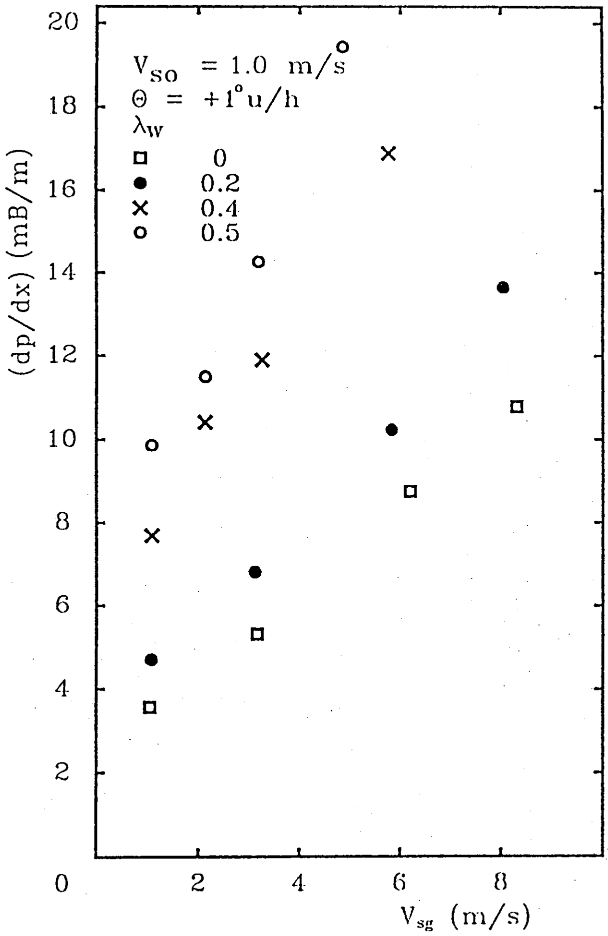


Fig 7.26 Oil/Water/Gas Pressure Loss,1 deg u/h,Oil No1

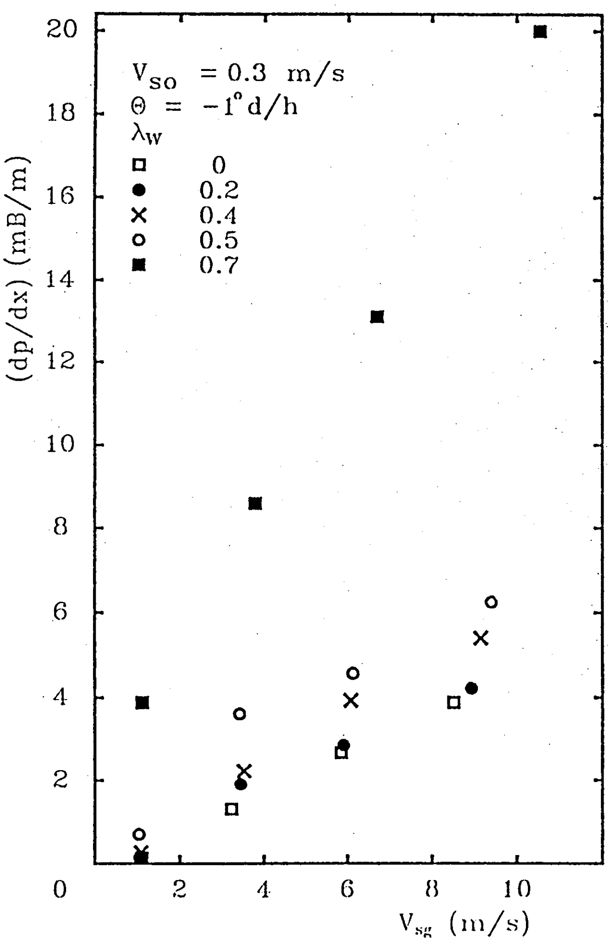


Fig 7.27 Oil/Water/Gas Pressure Loss,1 deg d/h,Oil No1

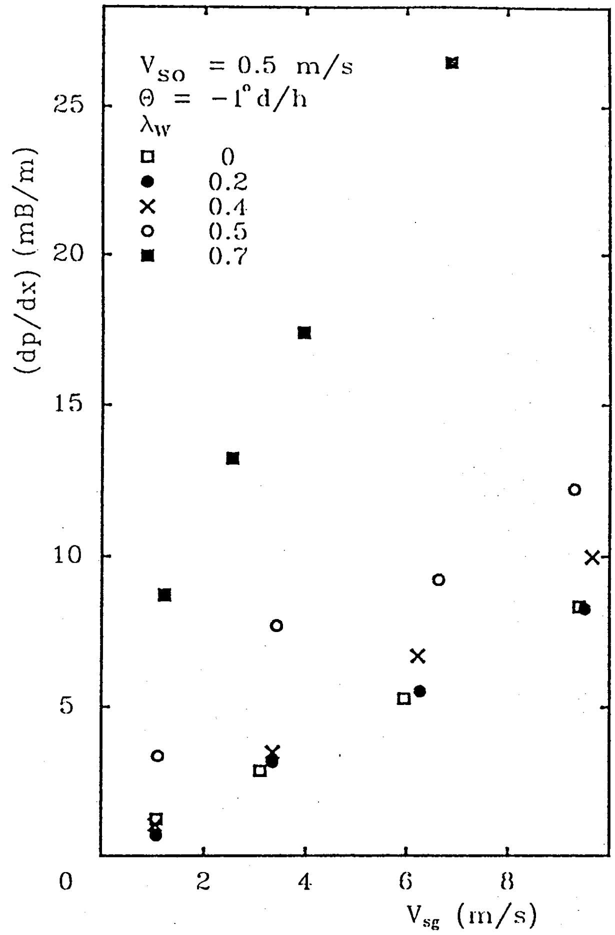


Fig 7.28 Oil/Water/Gas Pressure Loss,1 deg d/h,Oil No1

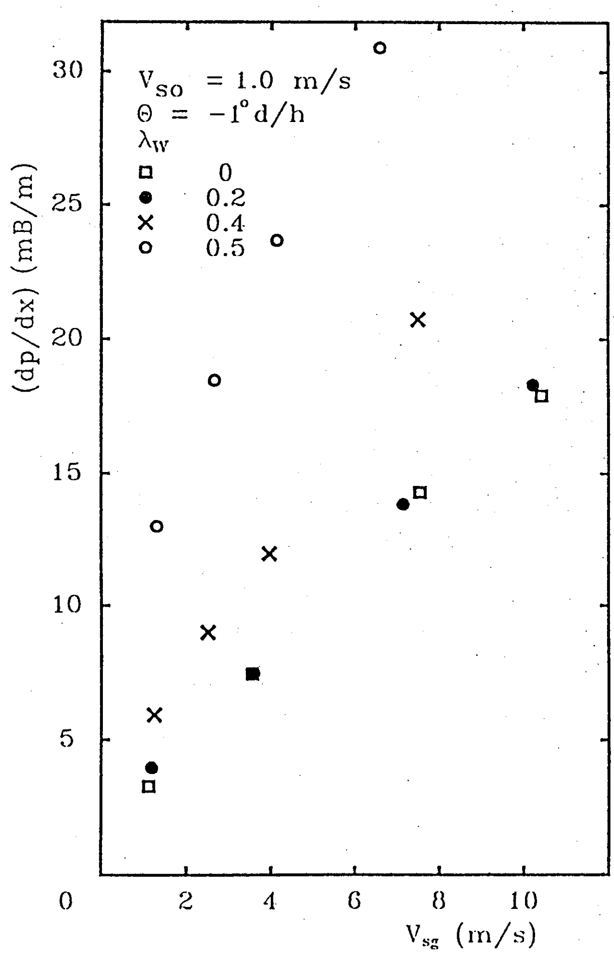


Fig 7.29 Oil/Water/Gas Pressure Loss,1 deg d/h,Oil No1

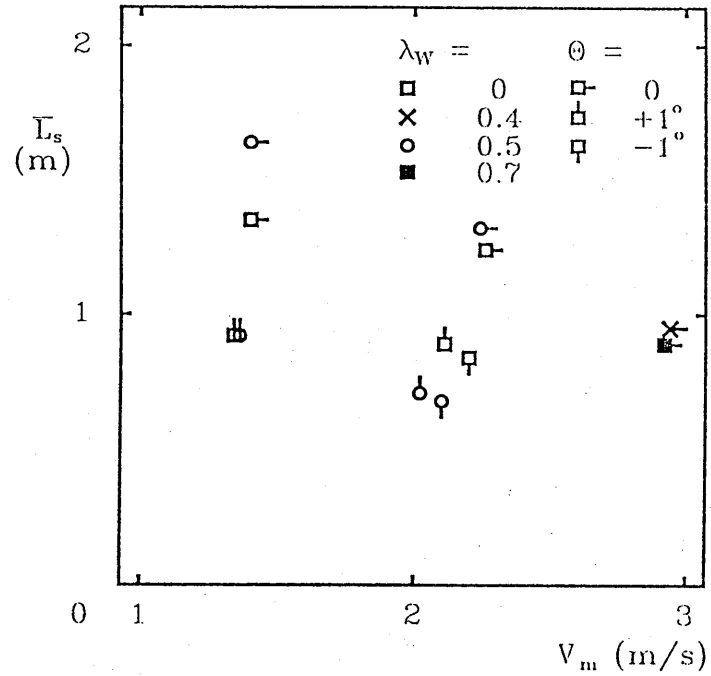


Fig 7.30 Oil/Water/Gas Average Slug Length,Oil No1

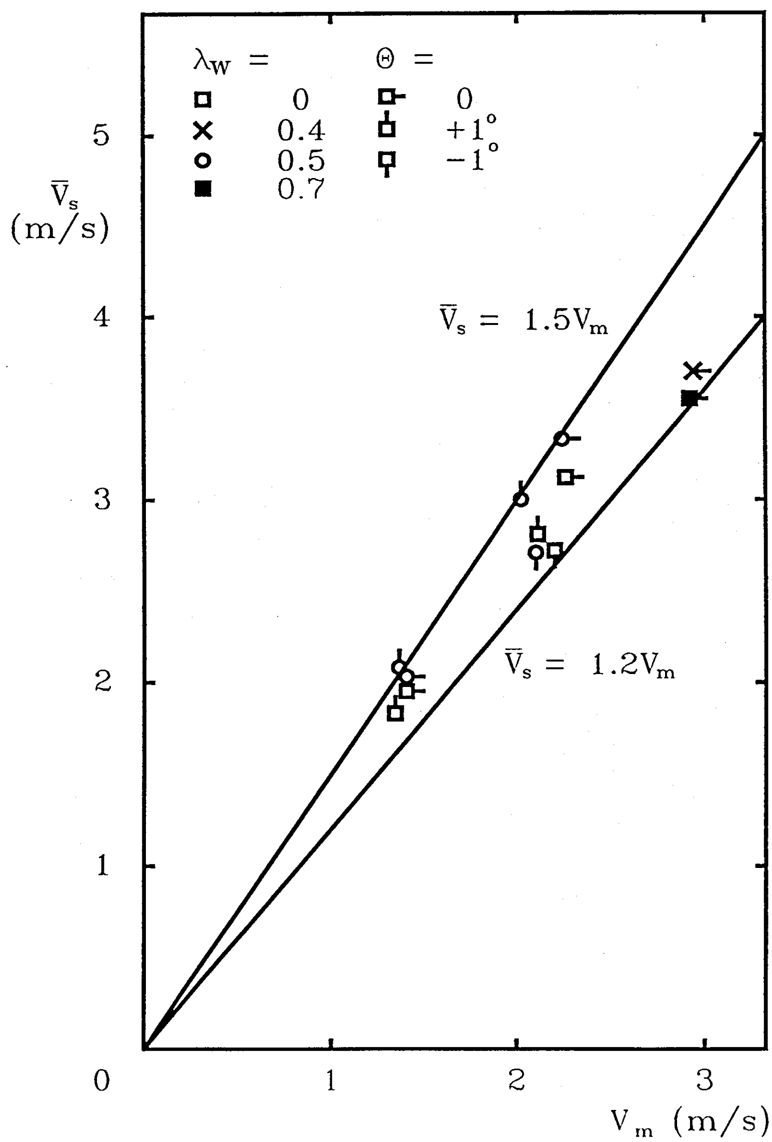


Fig 7.31 Oil/Water/Gas Average Slug Front Velocity,Oil No1

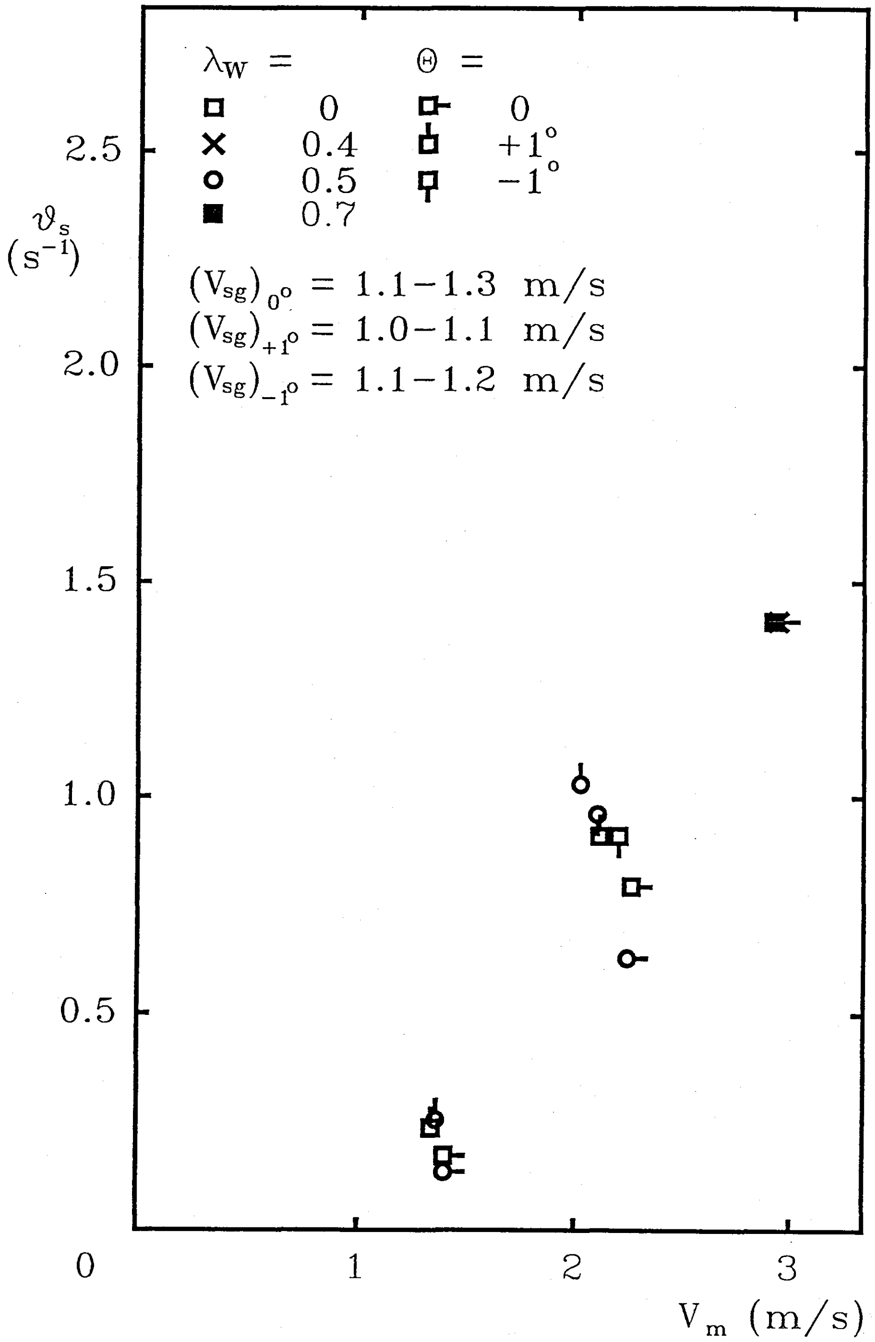


Fig 7.32 Oil/Water/Gas Slug Frequency,Oil No1

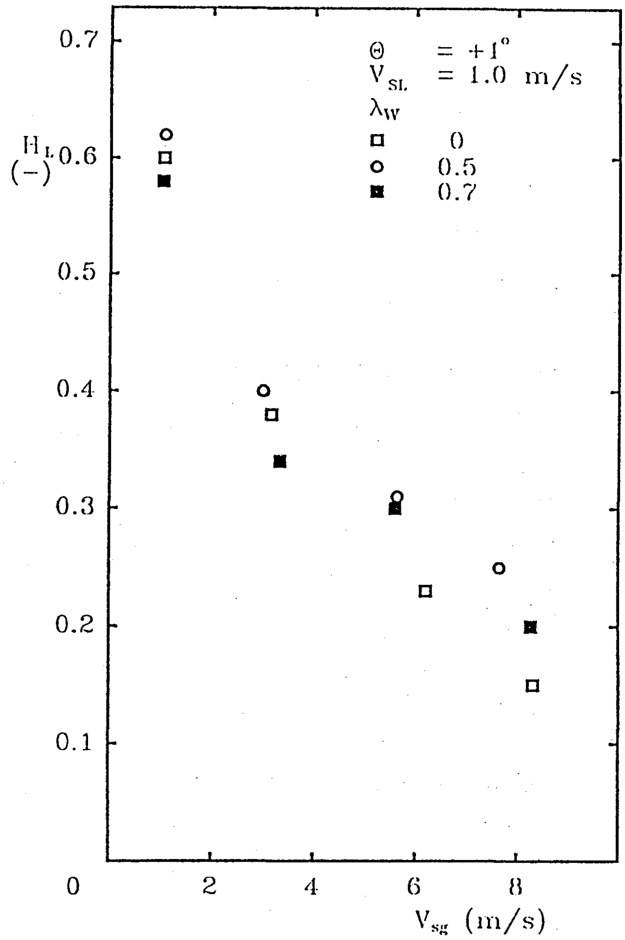


Fig 7.33 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

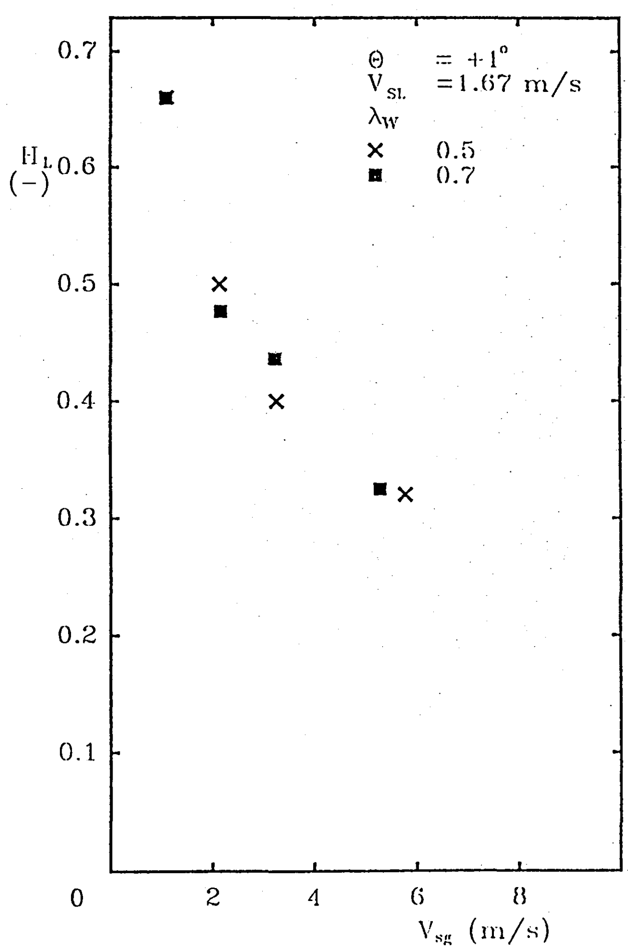


Fig 7.34 Oil/Water/Gas Total Liquids Holdup,1 deg u/h,Oil No1

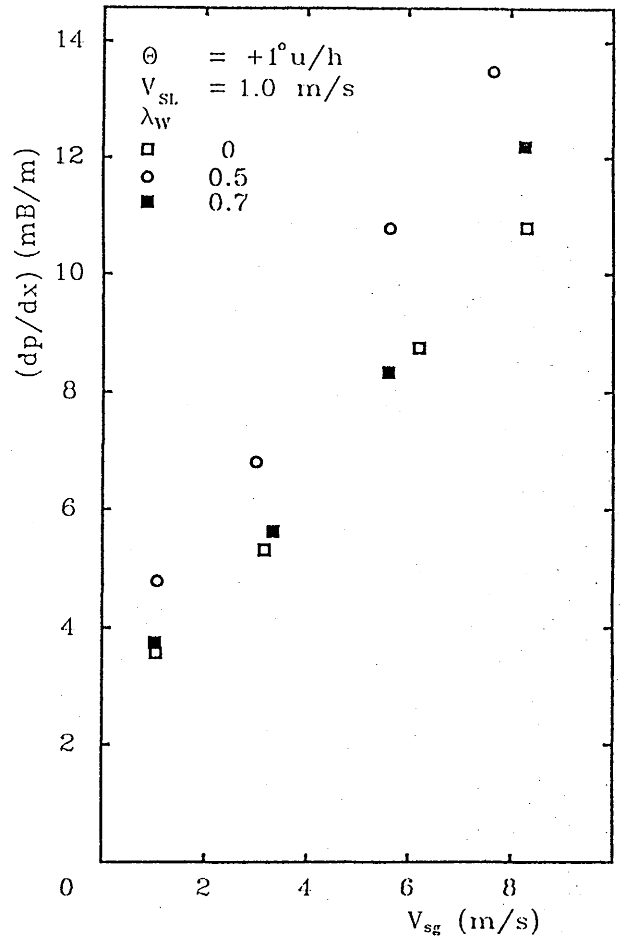


Fig 7.35 Oil/Water/Gas Pressure Drop,1 deg u/h,Oil No1

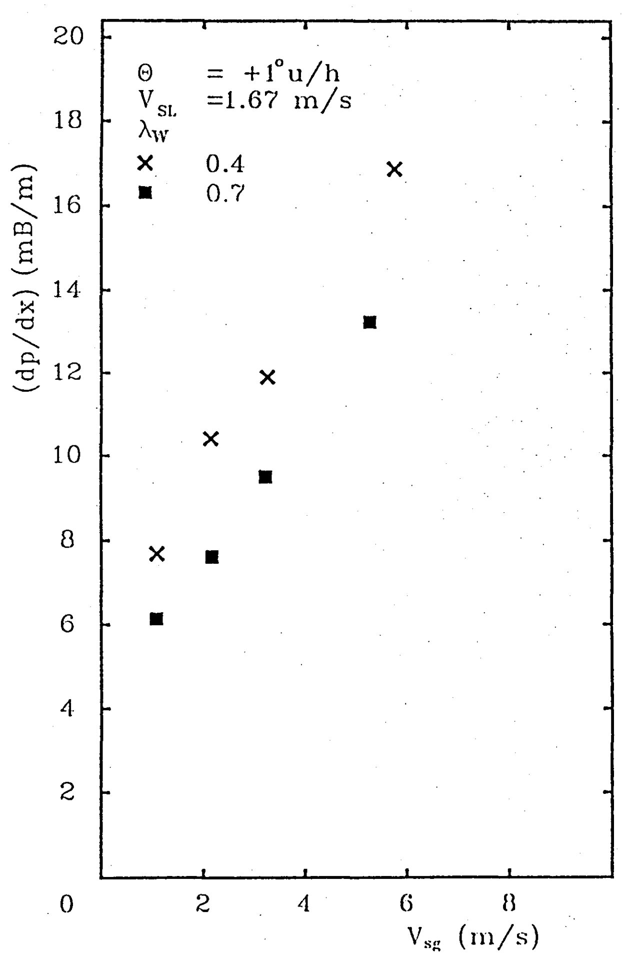


Fig 7.36 Oil/Water/Gas Pressure Drop,1 deg u/h,Oil No1

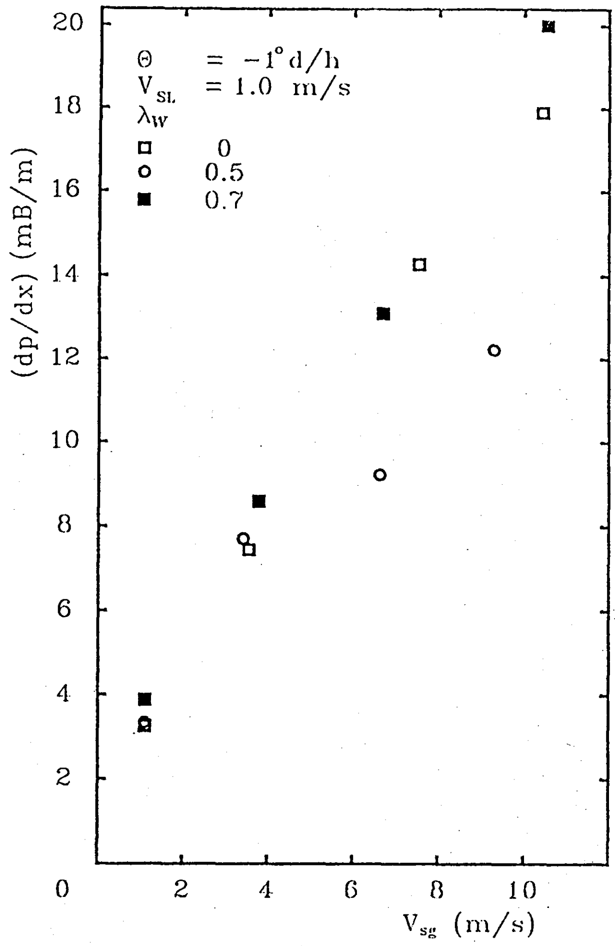


Fig 7.37 Oil/Water/Gas Pressure Drop,1 deg d/h,Oil No1

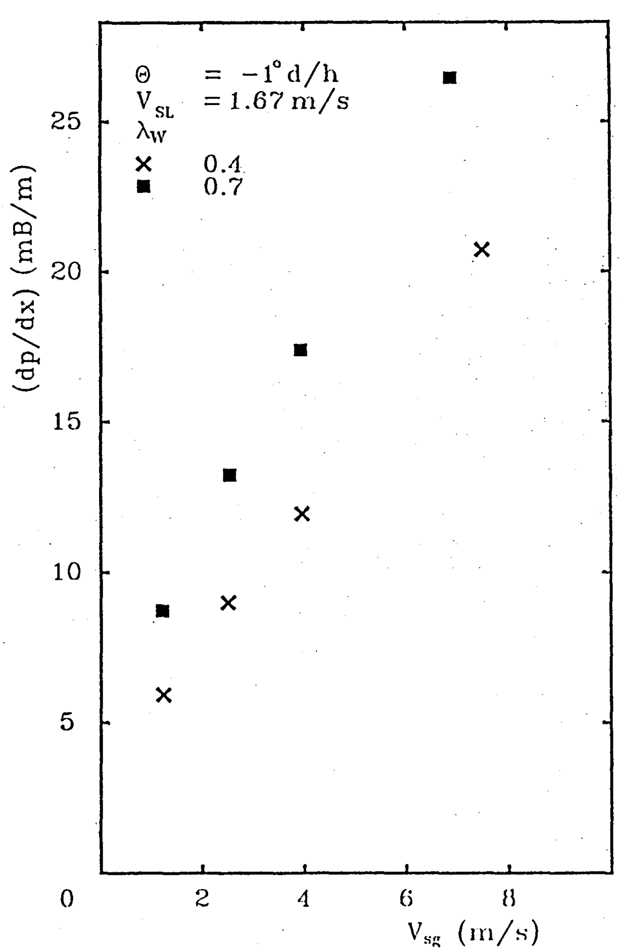


Fig 7.38 Oil/Water/Gas Pressure Drop,1 deg d/h,Oil No1

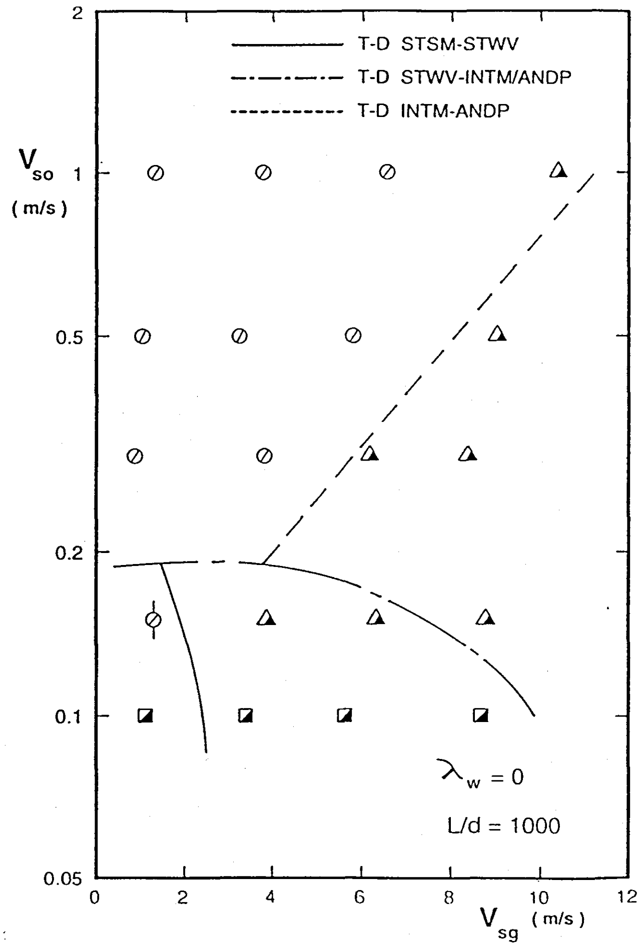


Fig 8.1 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No1

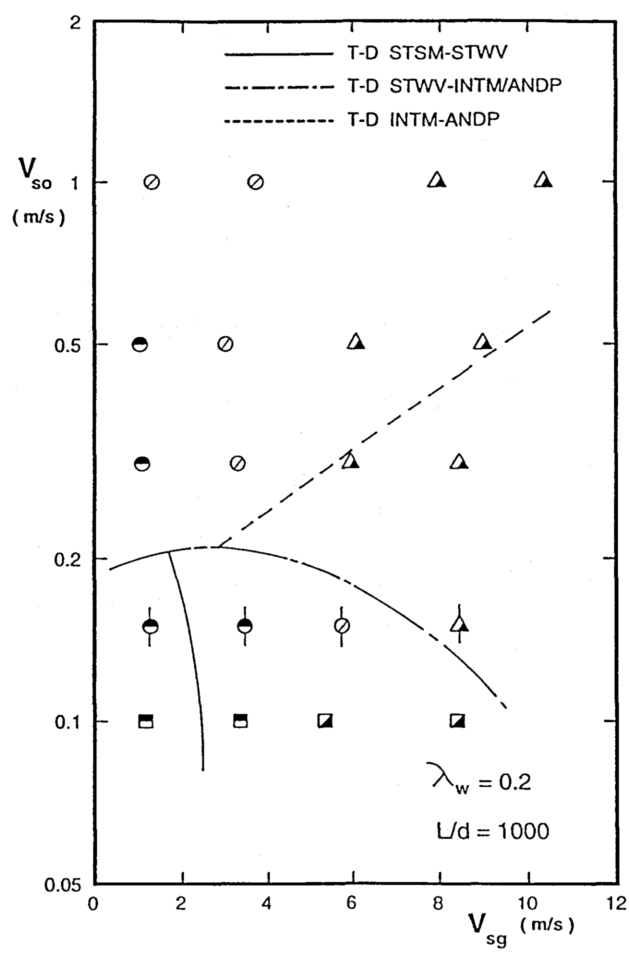


Fig 8.2 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No1

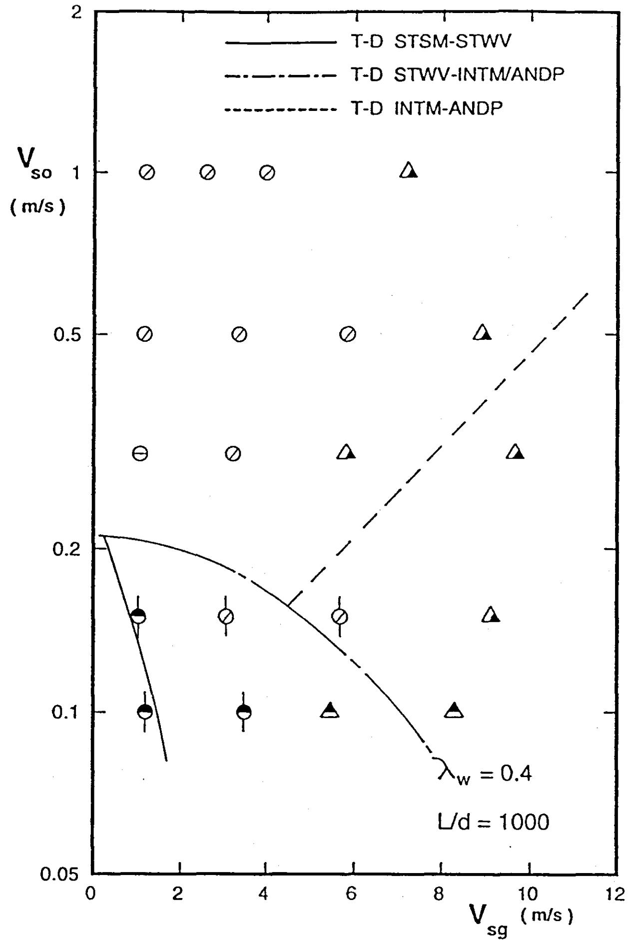


Fig 8.3 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No1

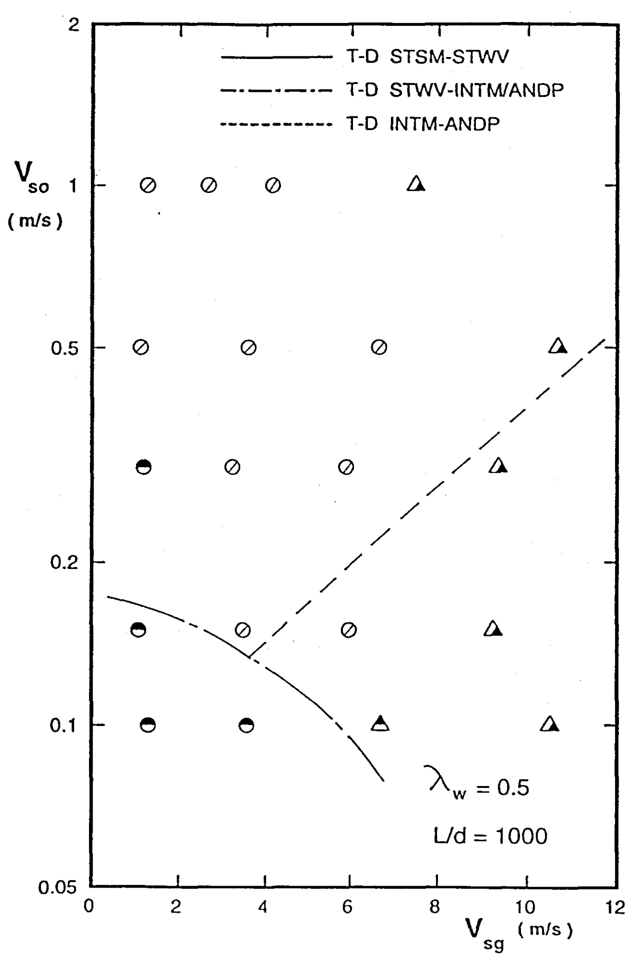


Fig 8.4 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No1

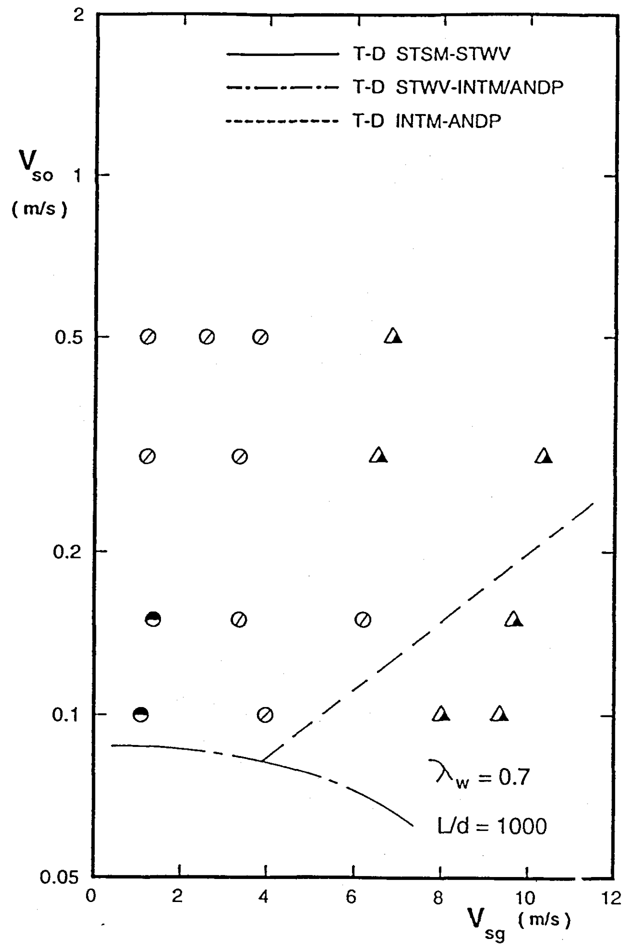


Fig 8.5 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No1

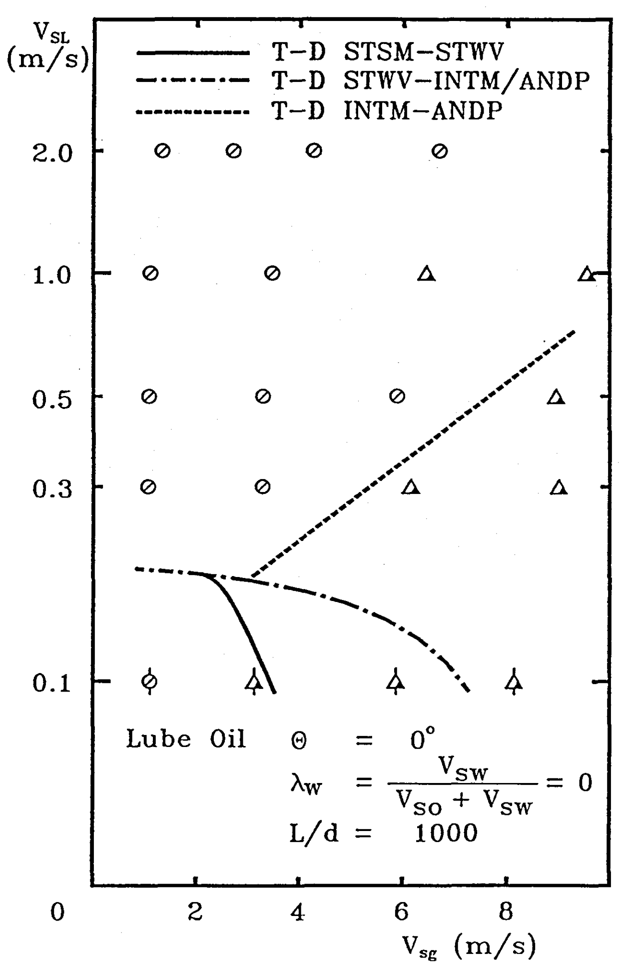


Fig 8.6 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No2

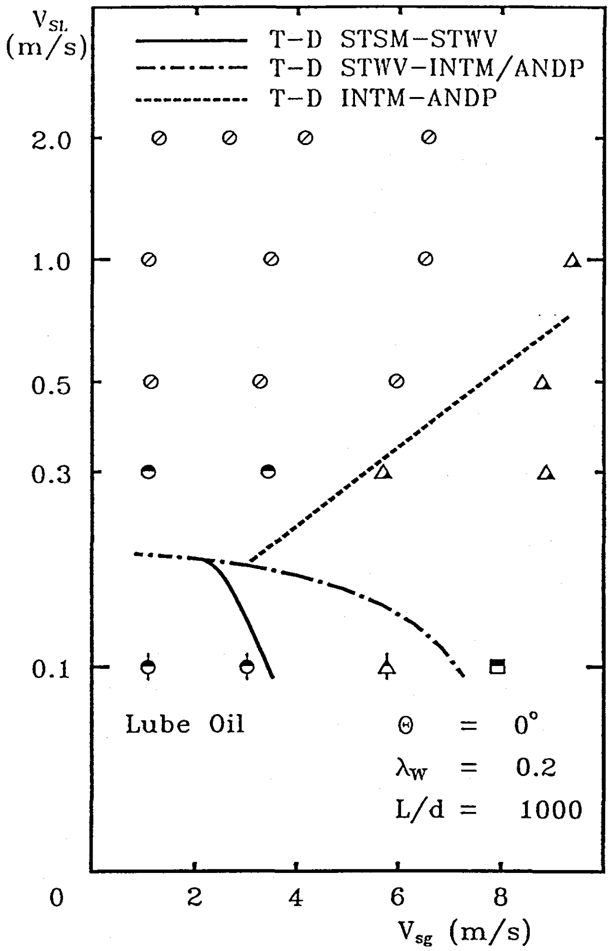


Fig 8.7 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No2

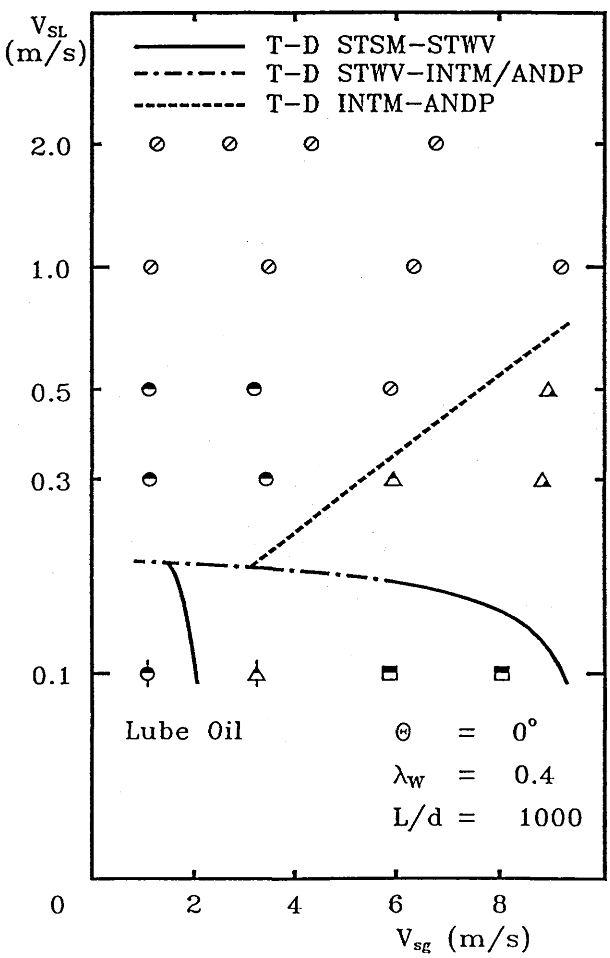


Fig 8.8 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No2

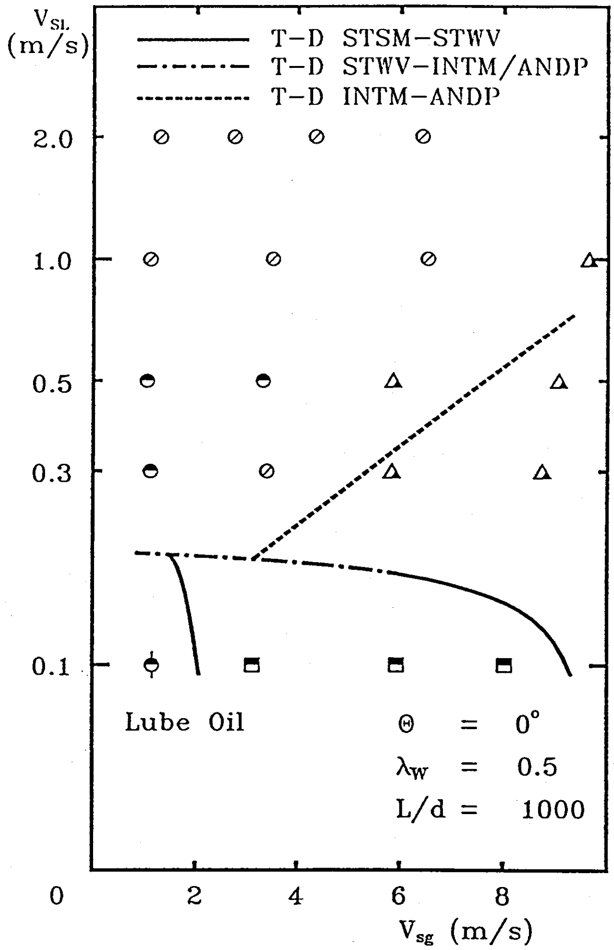


Fig 8.9 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No2

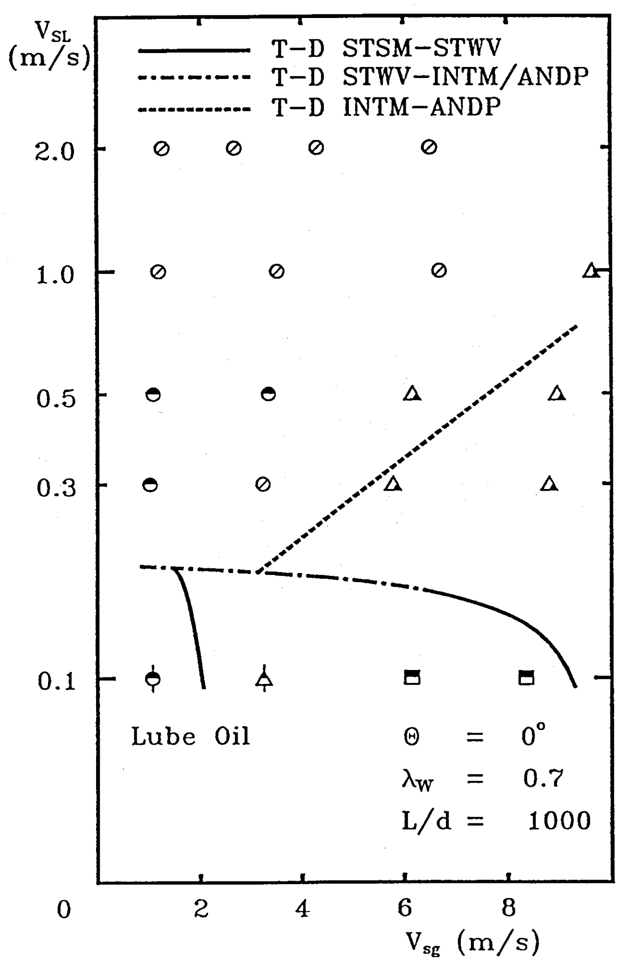


Fig 8.10 Oil/Water/Gas Flow Pattern at L=1000d ,Oil No2

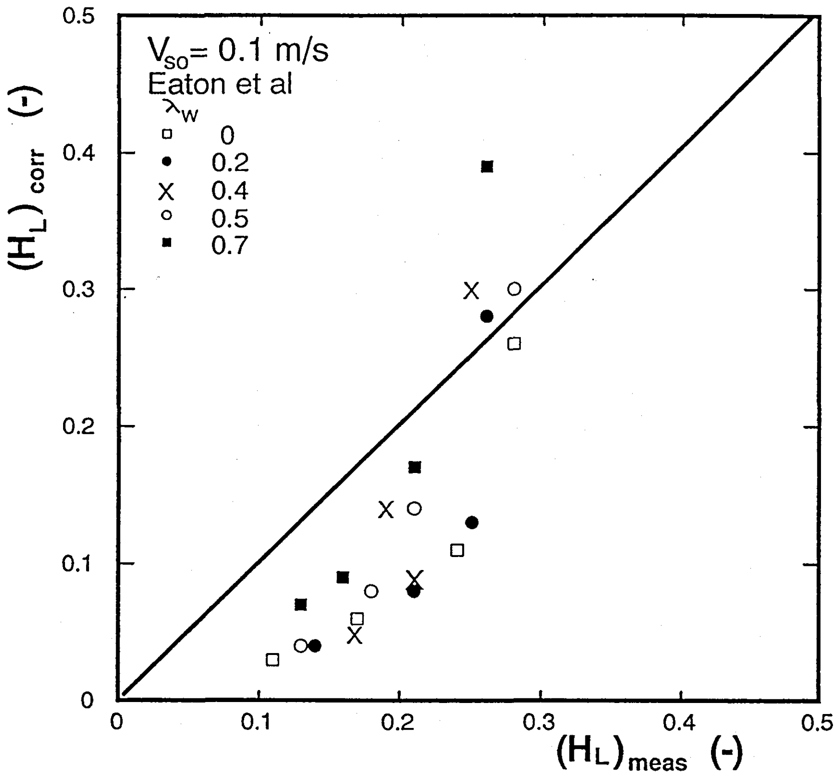


Fig 8.11 Liquid Holdup Compared to Eaton et al correlation,Oil No1

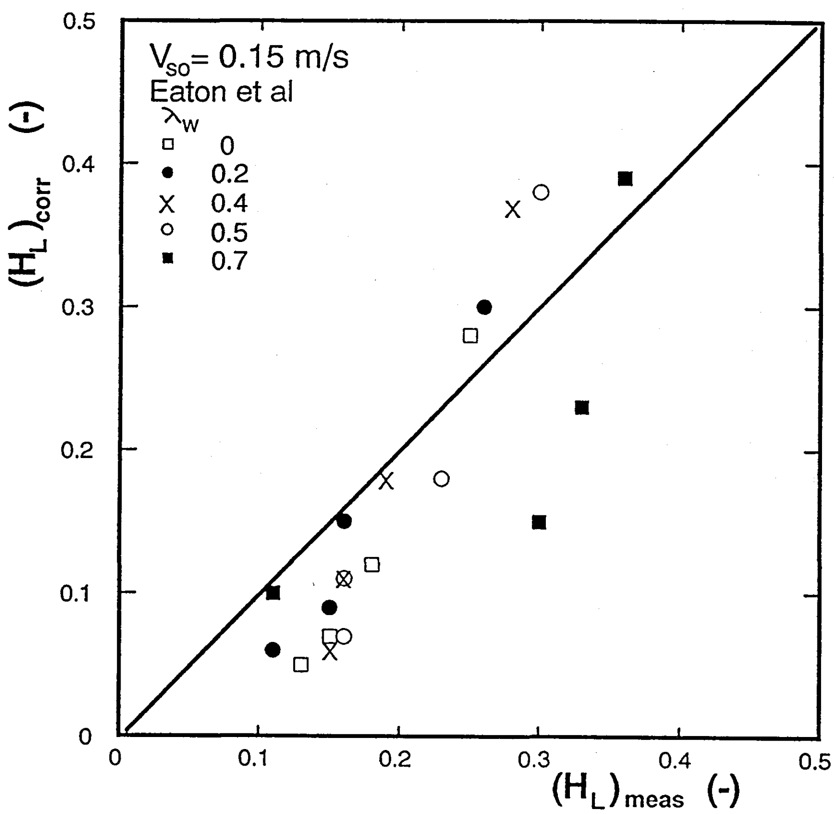


Fig 8.12 Liquid Holdup Compared to Eaton et al correlation,Oil No1

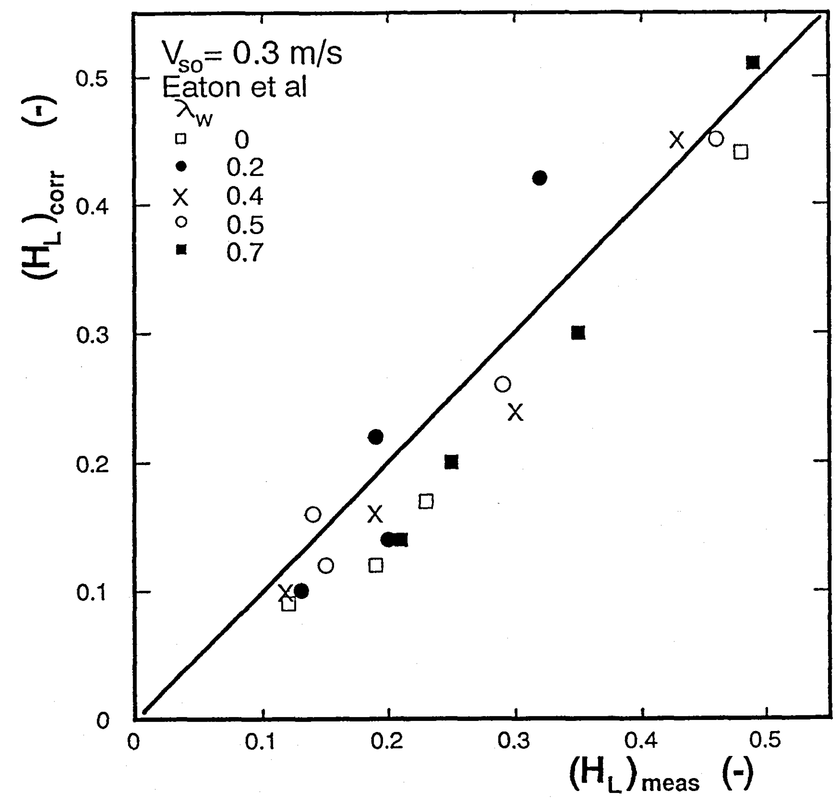


Fig 8.13 Liquid Holdup Compared to Eaton et al correlation,Oil No1

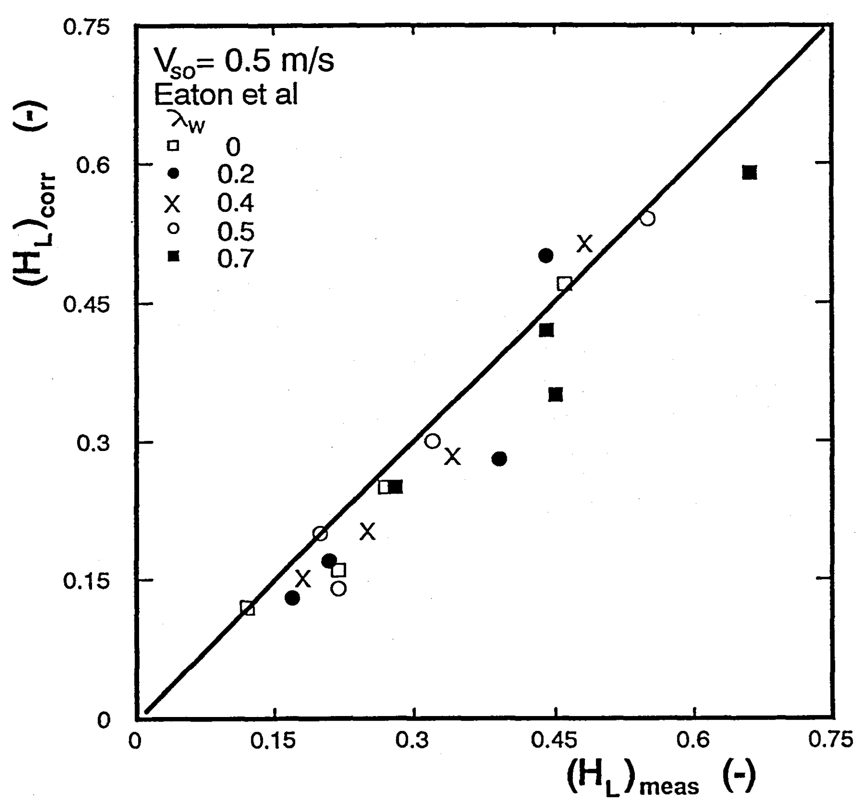


Fig 8.14 Liquid Holdup Compared to Eaton et al correlation,Oil No1

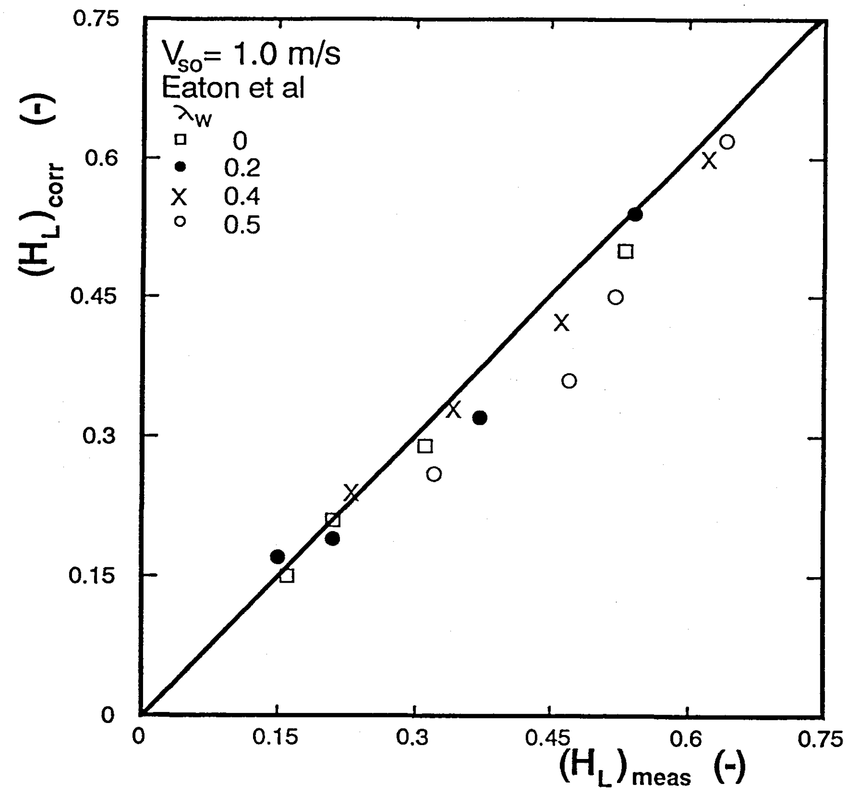


Fig 8.15 Liquid Holdup Compared to Eaton et al correlation,Oil No1

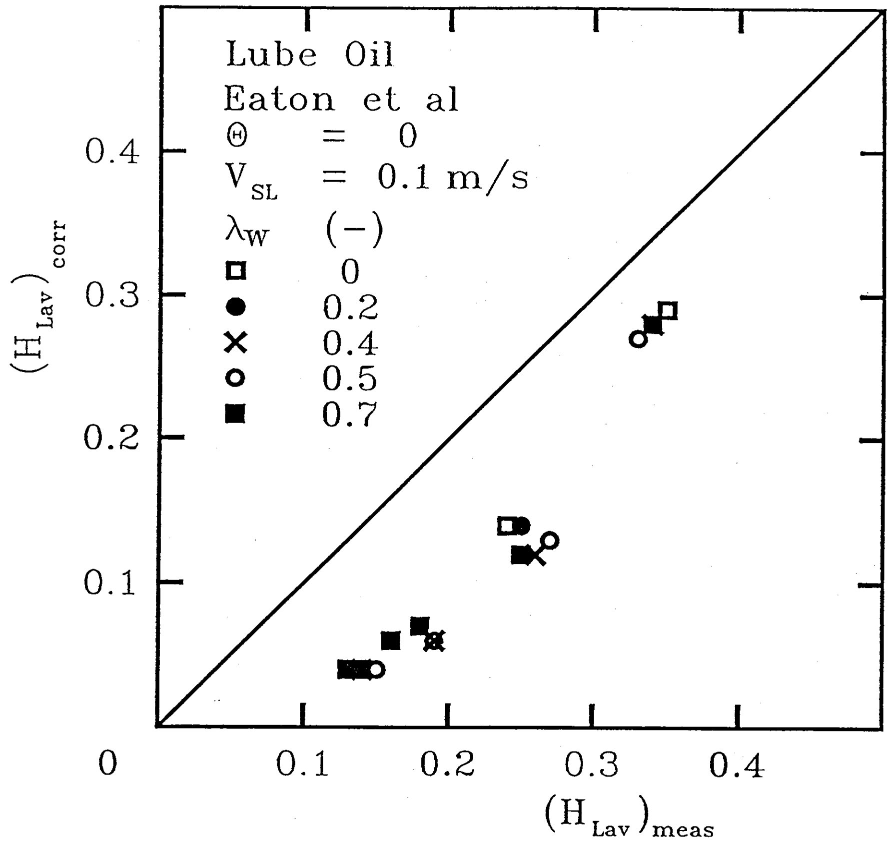


Fig 8.16 Liquid Holdup Compared to Eaton et al correlation,Oil No2

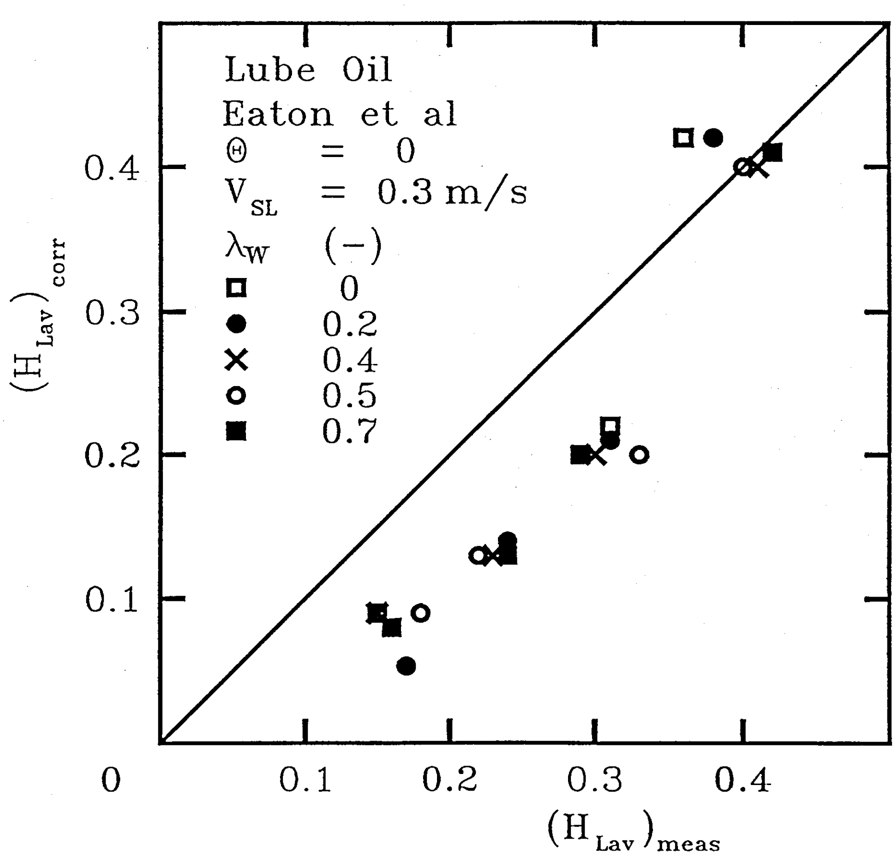


Fig 8.17 Liquid Holdup Compared to Eaton et al correlation,Oil No2

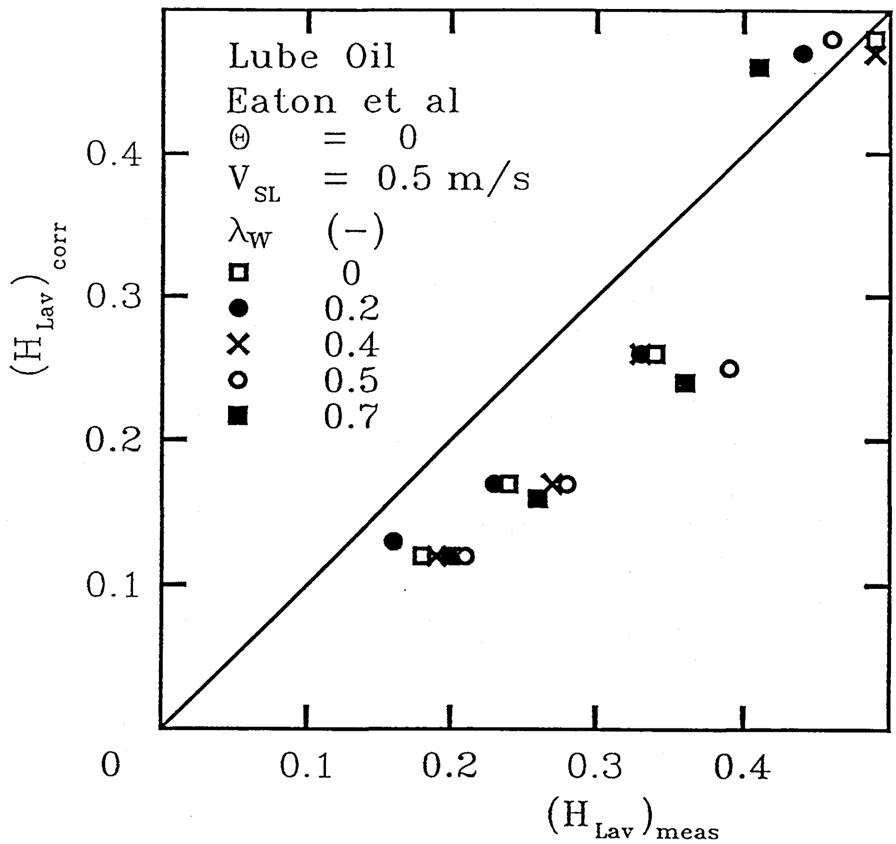


Fig 8.18 Liquid Holdup Compared to Eaton et al correlation,Oil No2

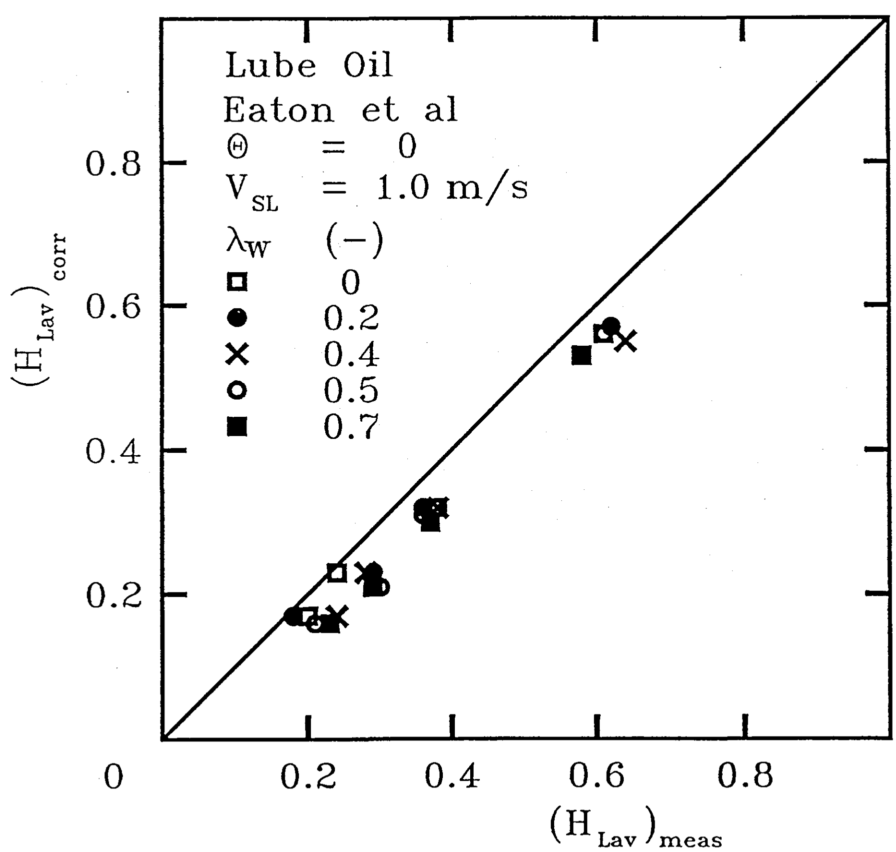


Fig 8.19 Liquid Holdup Compared to Eaton et al correlation,Oil No2

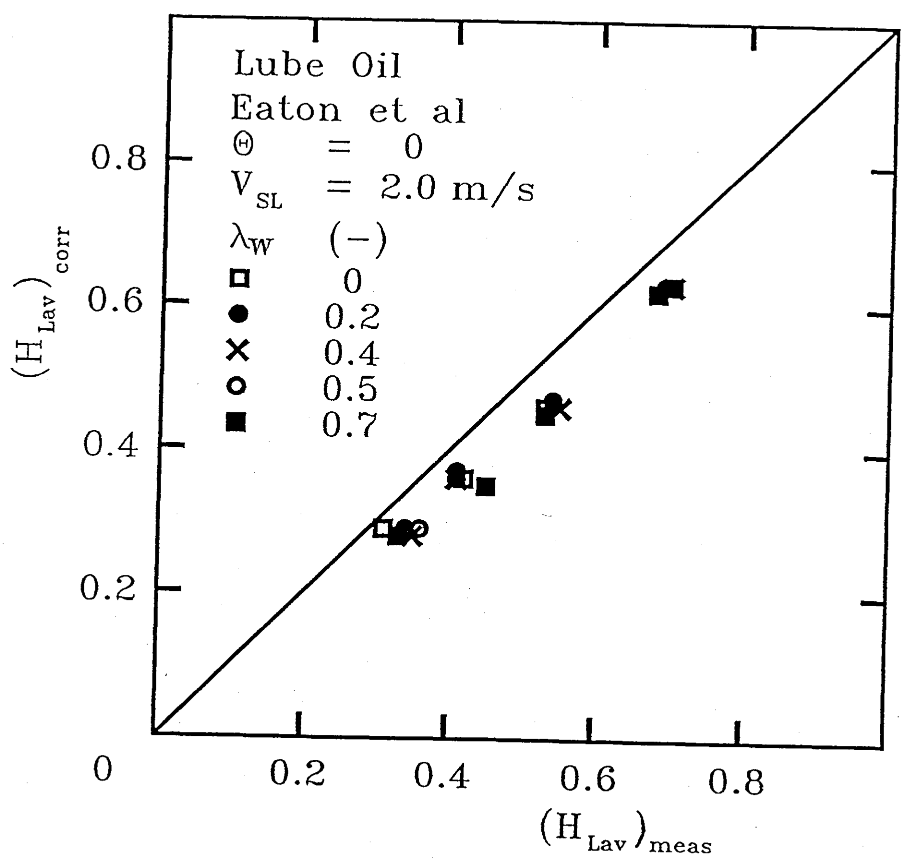


Fig 8.20 Liquid Holdup Compared to Eaton et al correlation,Oil No2

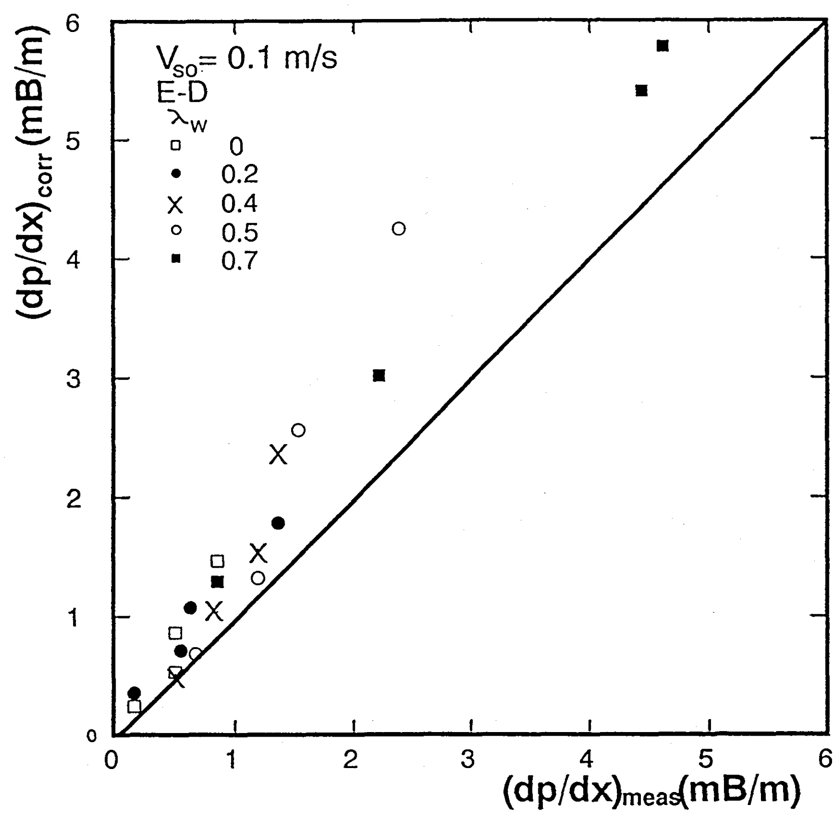


Fig 8.21 Pressure Loss Compared to Eaton-Dukler correlation,Oil No1

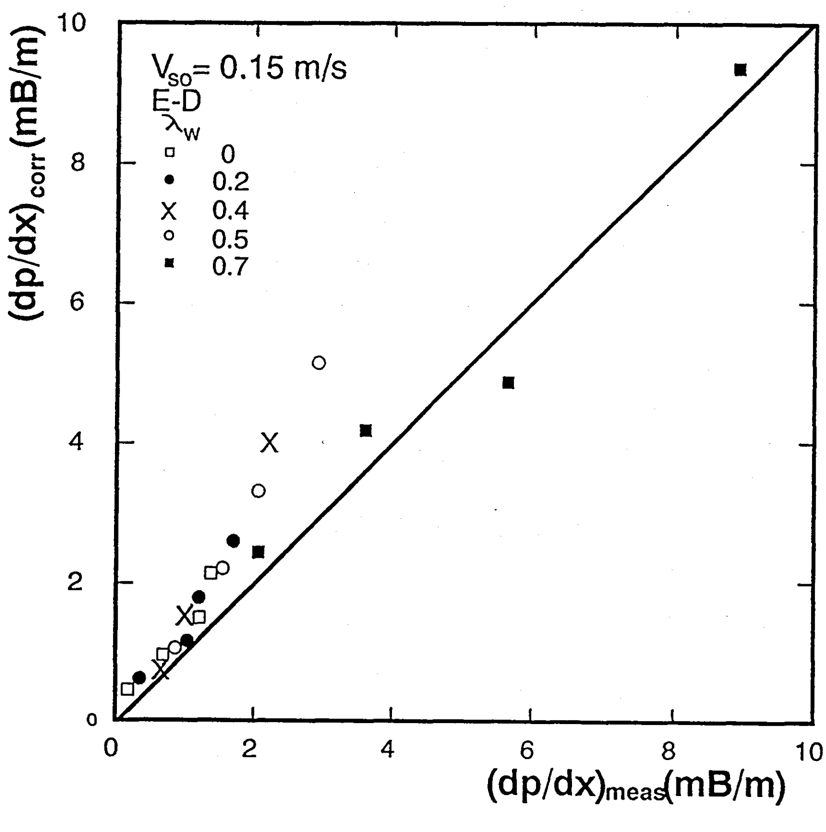


Fig 8.22 Pressure Loss Compared to Eaton-Dukler correlation,Oil No1

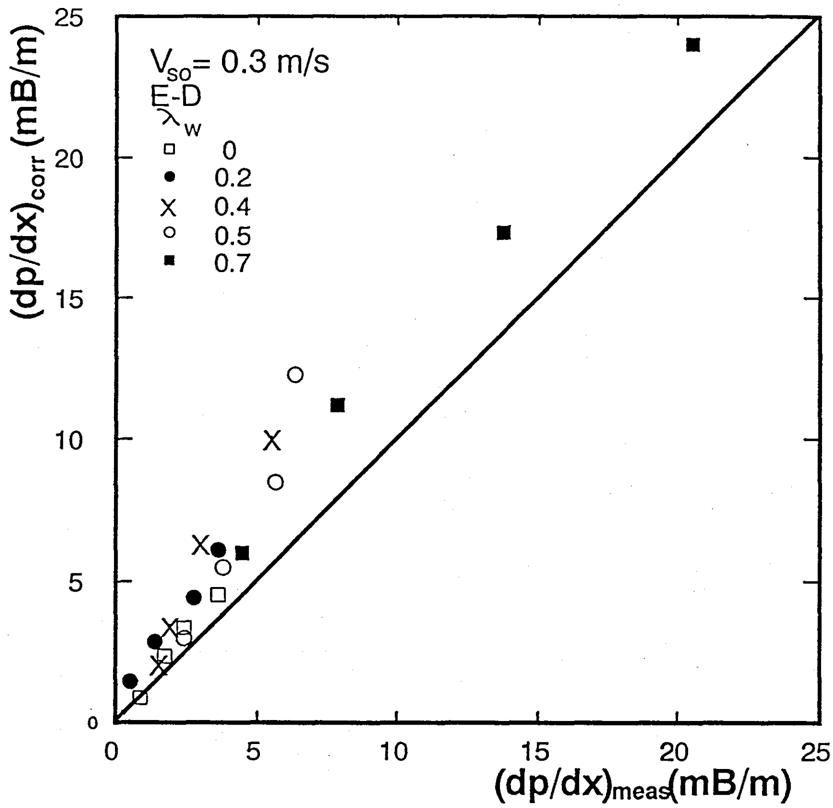


Fig 8.23 Pressure Loss Compared to Eaton-Dukler correlation,Oil No1

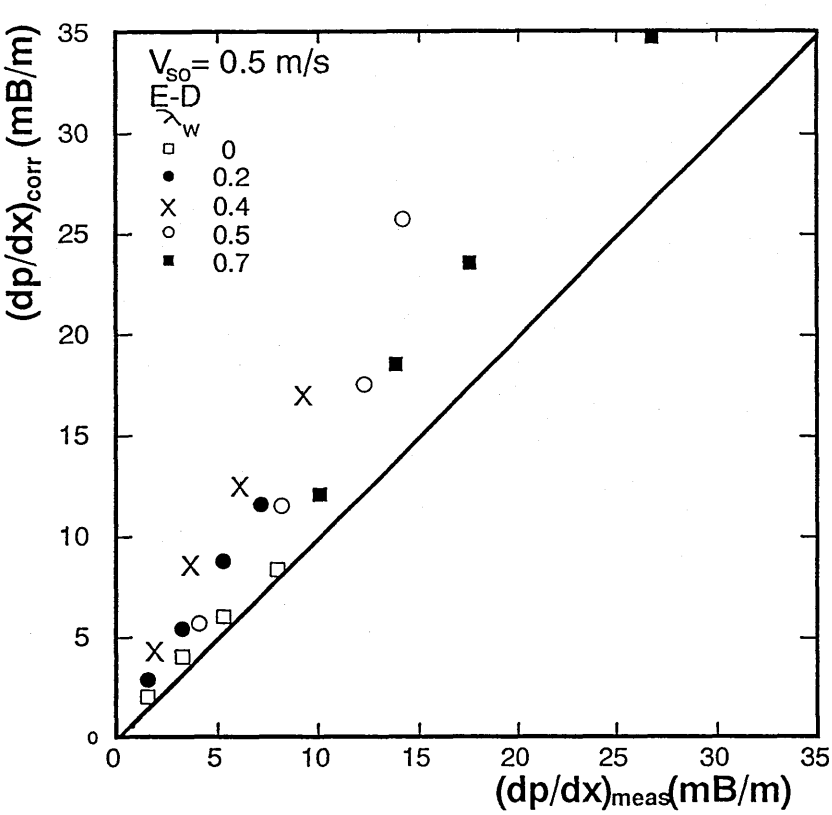


Fig 8.24 Pressure Loss Compared to Eaton-Dukler correlation,Oil No1

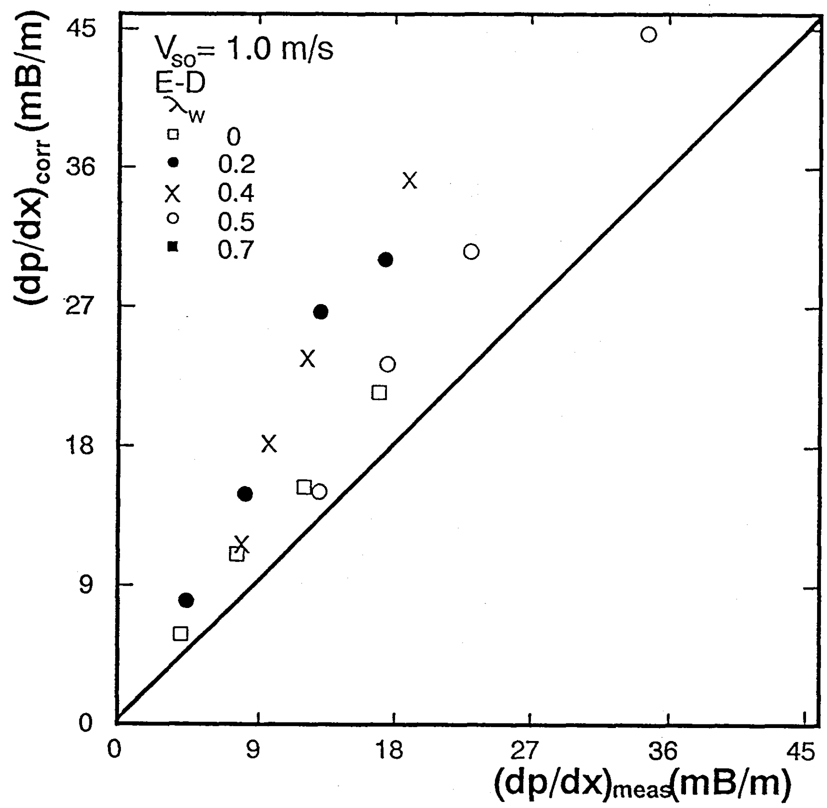


Fig 8.25 Pressure Loss Compared to Eaton-Dukler correlation,Oil No1

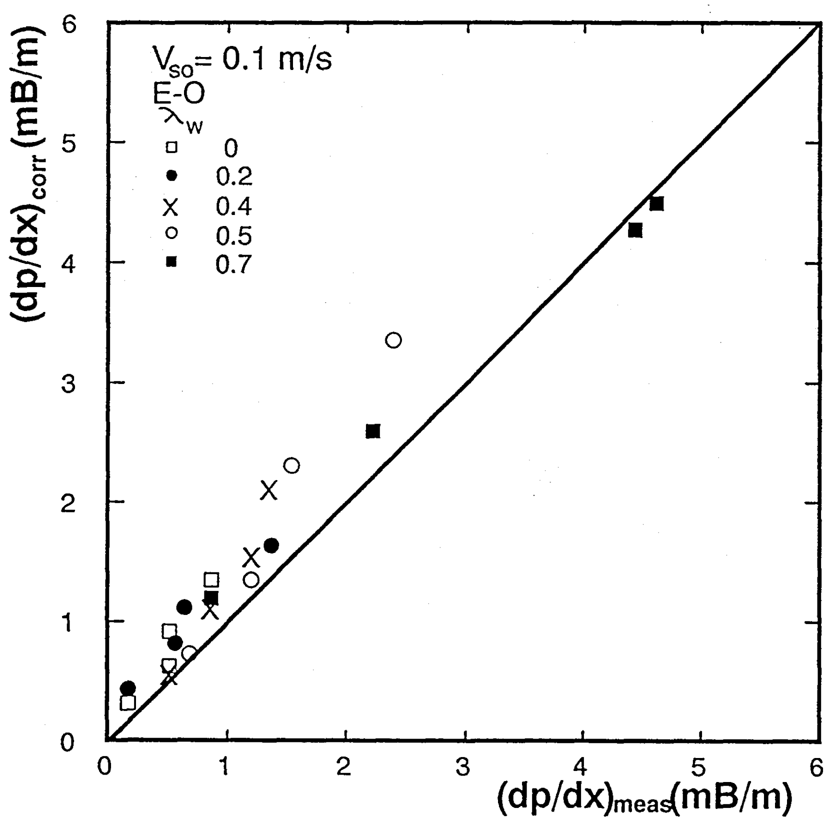


Fig 8.26 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No1

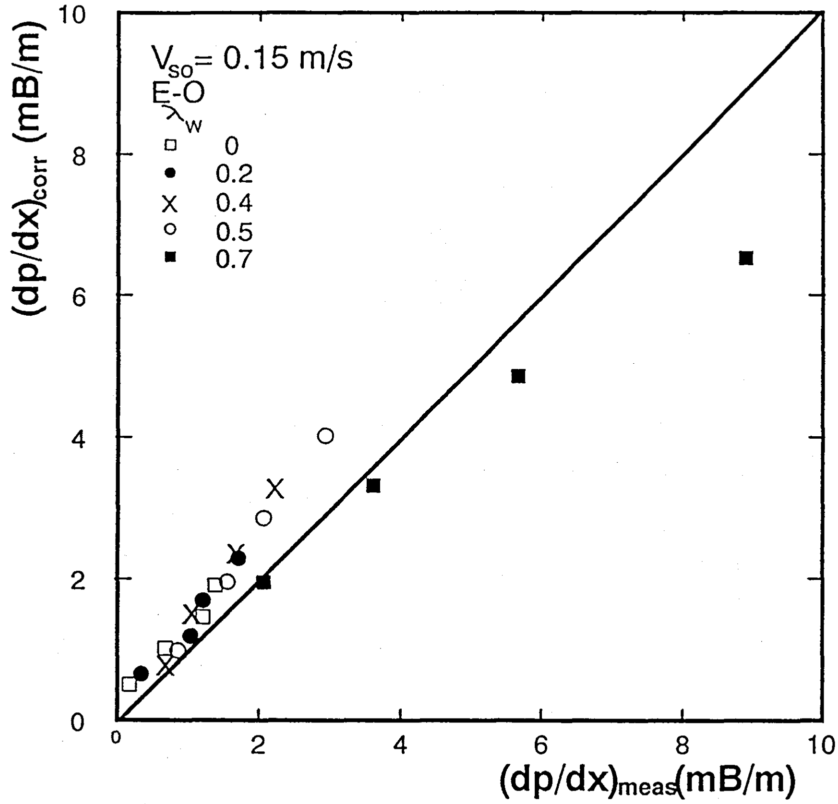


Fig 8.27 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No1

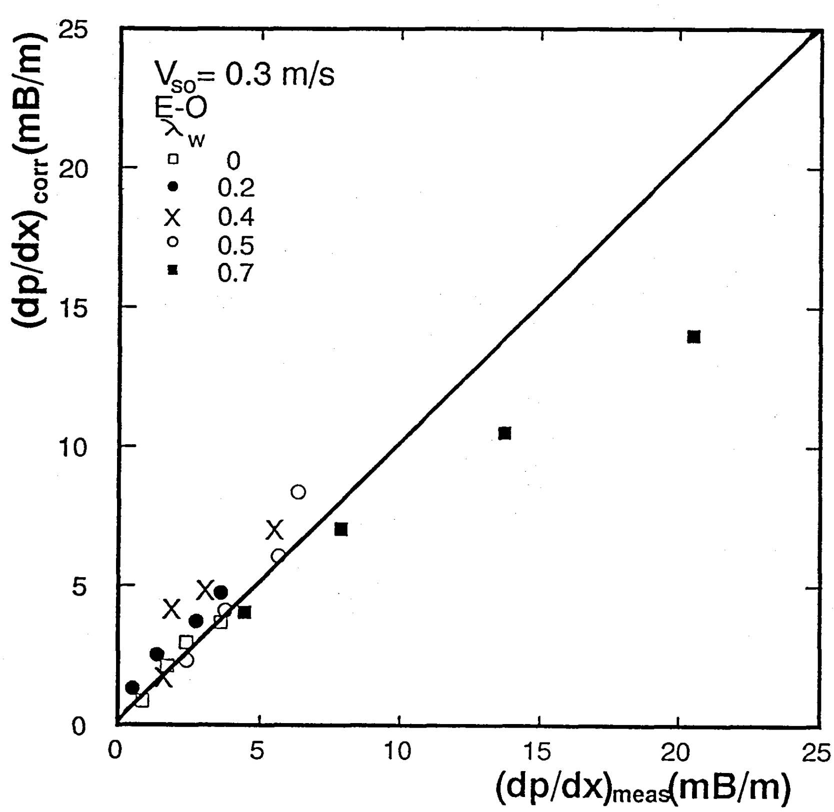


Fig 8.28 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No1

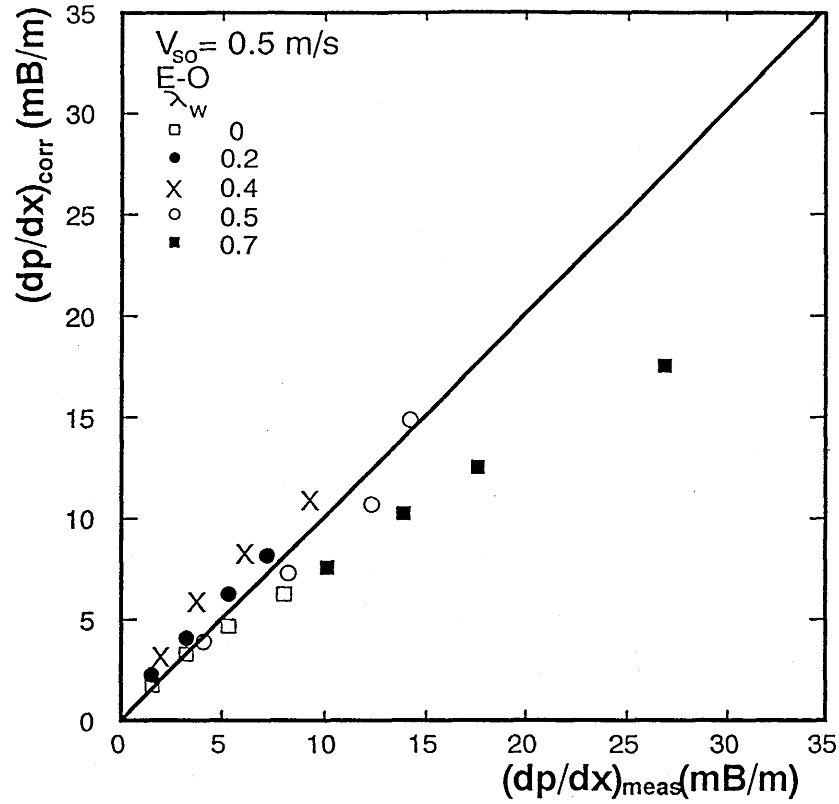


Fig 8.29 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No1

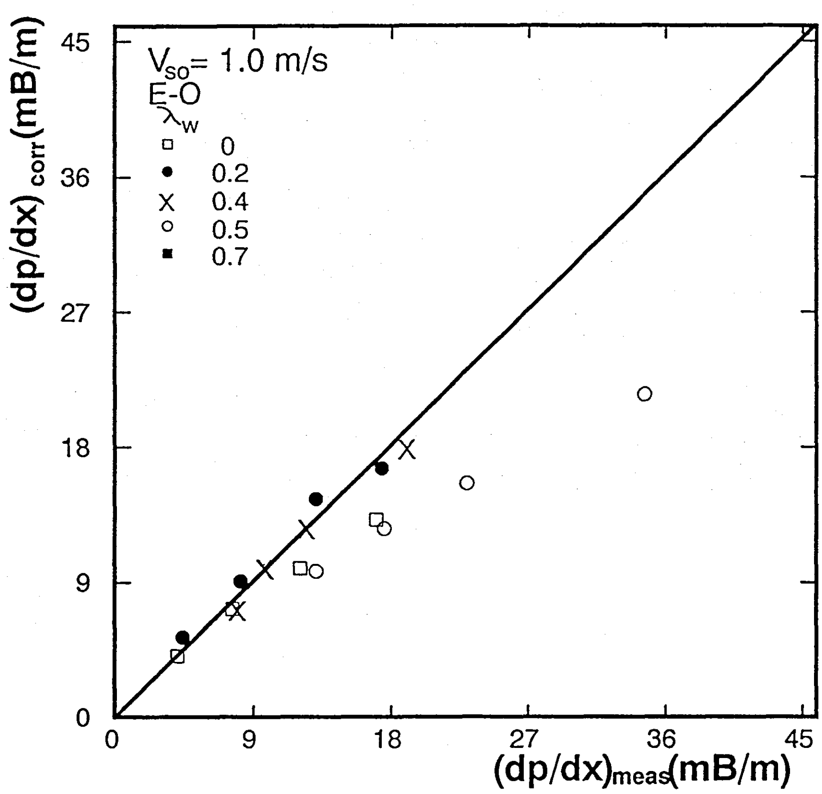


Fig 8.30 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No1

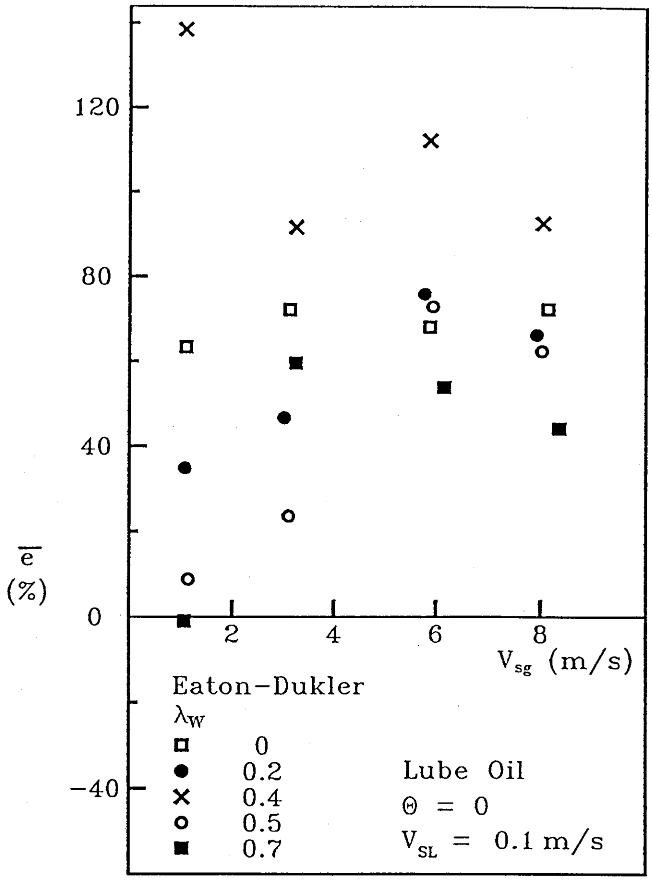


Fig 8.31 Pressure Loss Compared to Eaton-Dukler correlation,Oil No2

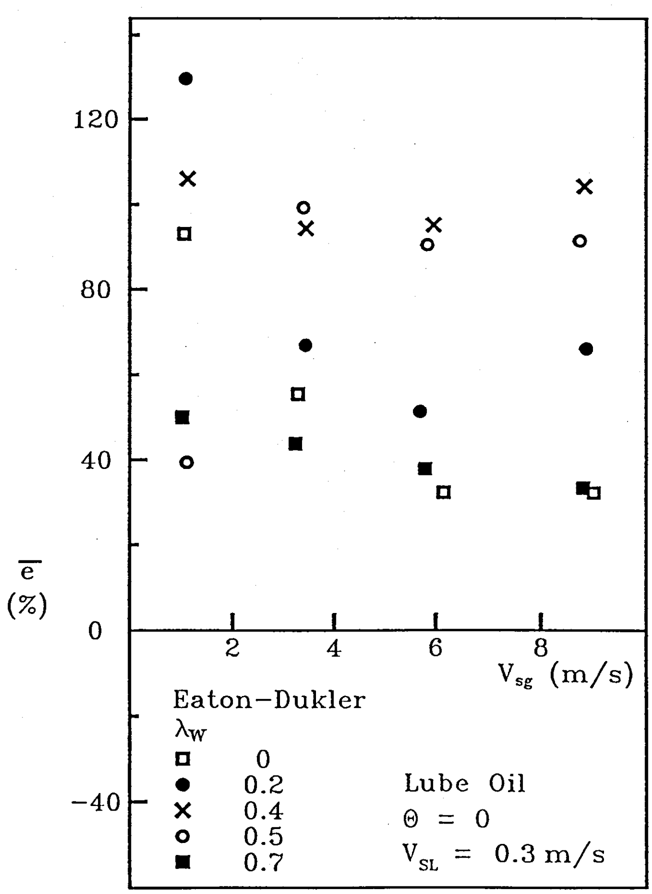


Fig 8.32 Pressure Loss Compared to Eaton-Dukler correlation,Oil No2

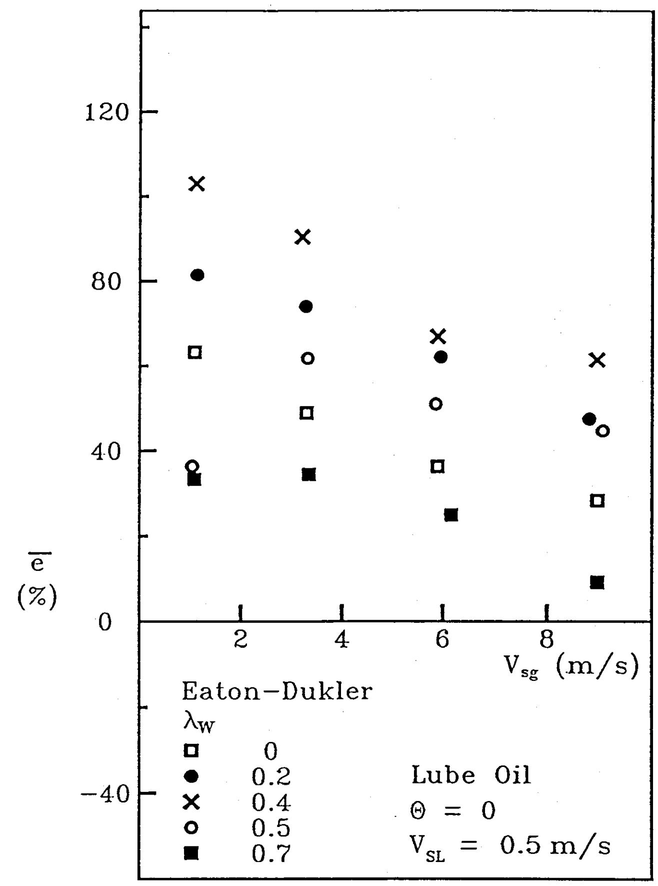


Fig 8.33 Pressure Loss Compared to Eaton-Dukler correlation,Oil No2

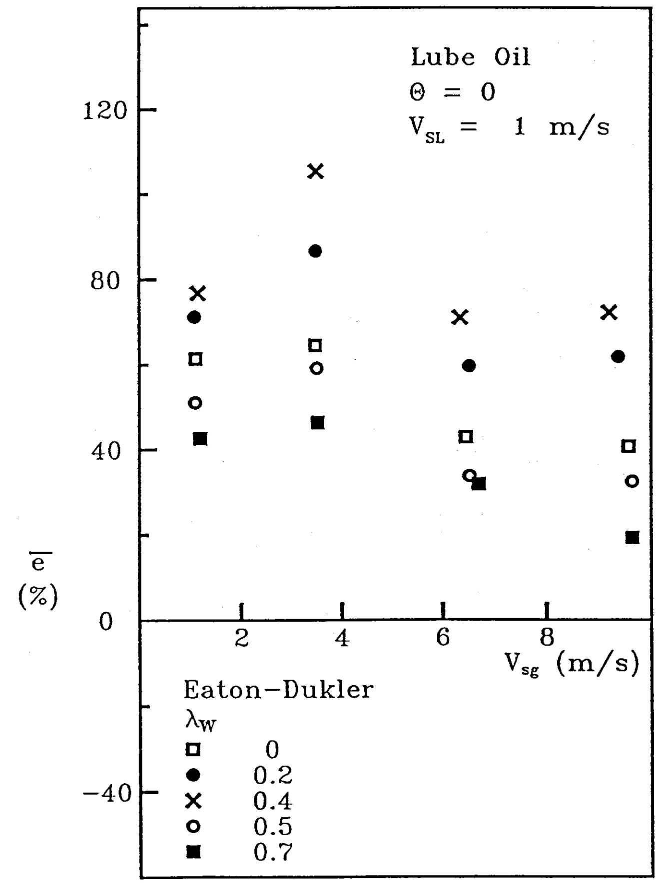


Fig 8.34 Pressure Loss Compared to Eaton-Dukler correlation,Oil No2

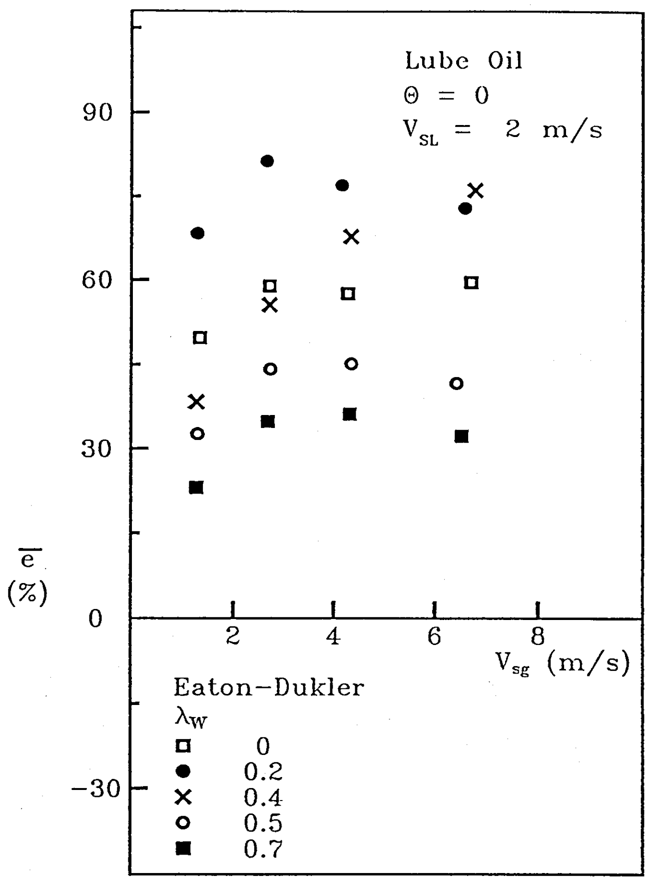


Fig 8.35 Pressure Loss Compared to Eaton-Dukler correlation,Oil No2

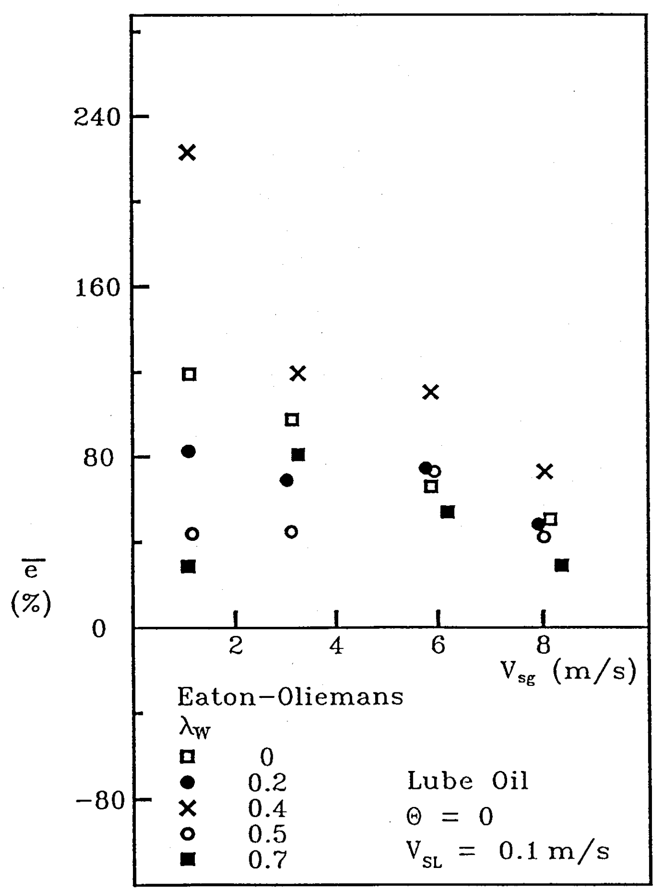


Fig 8.36 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No2

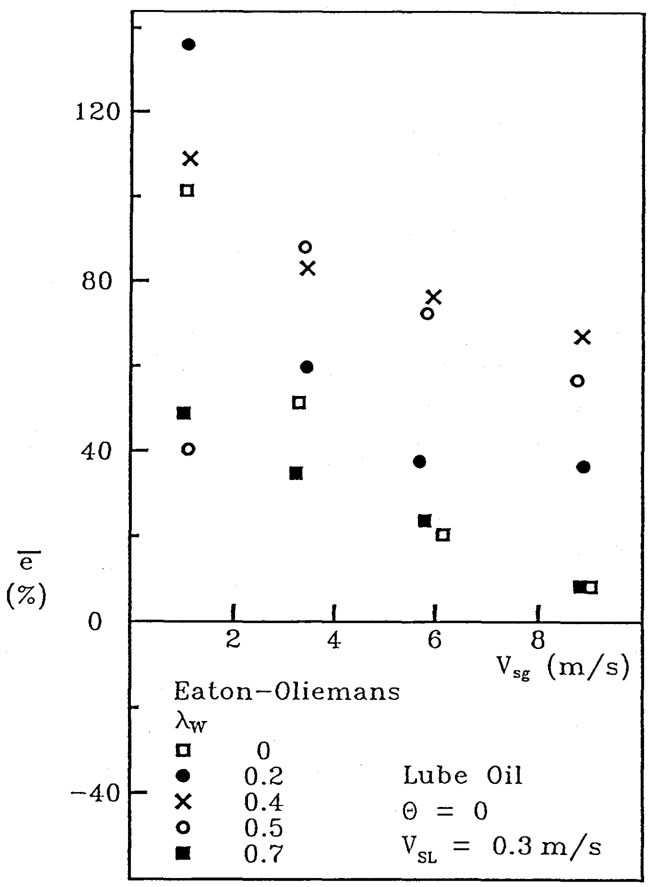


Fig 8.37 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No2

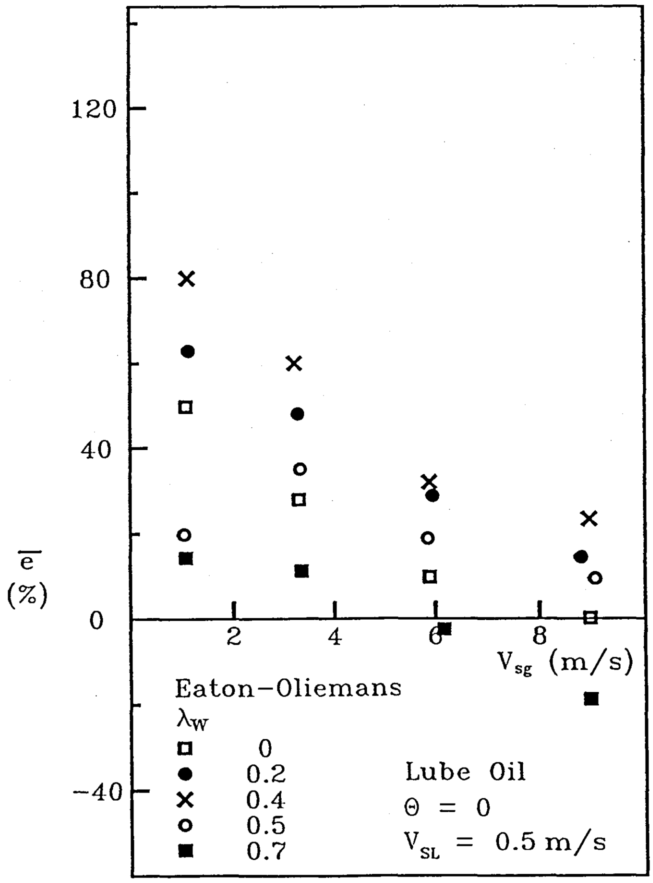


Fig 8.38 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No2

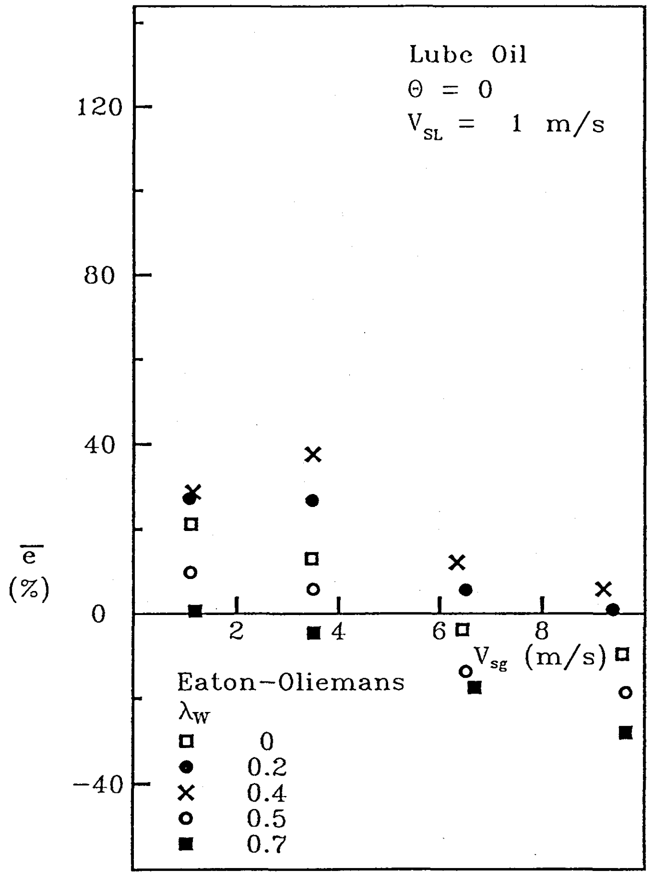


Fig 8.39 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No2

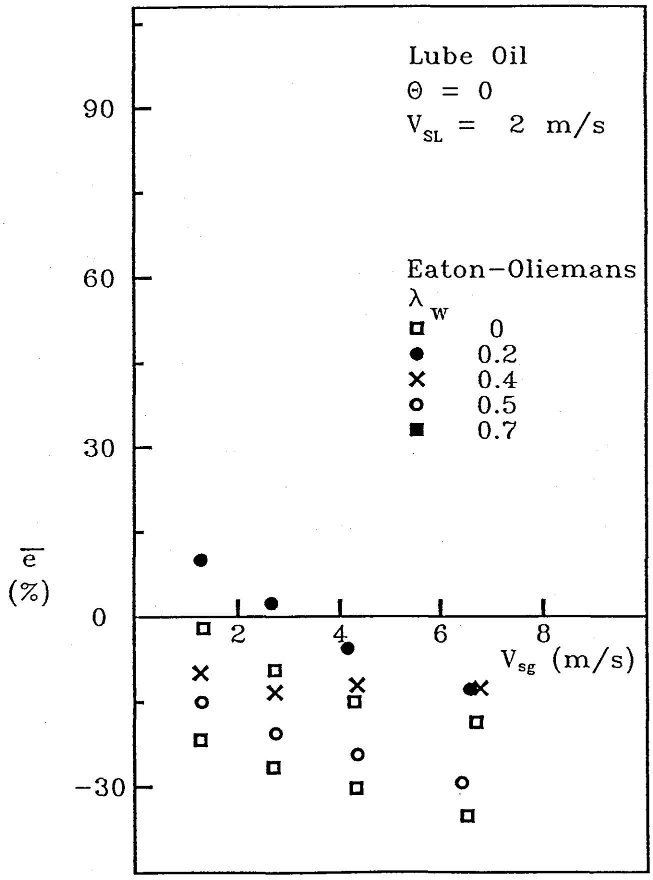


Fig 8.40 Pressure Loss Compared to Eaton-Oliemans correlation,Oil No2

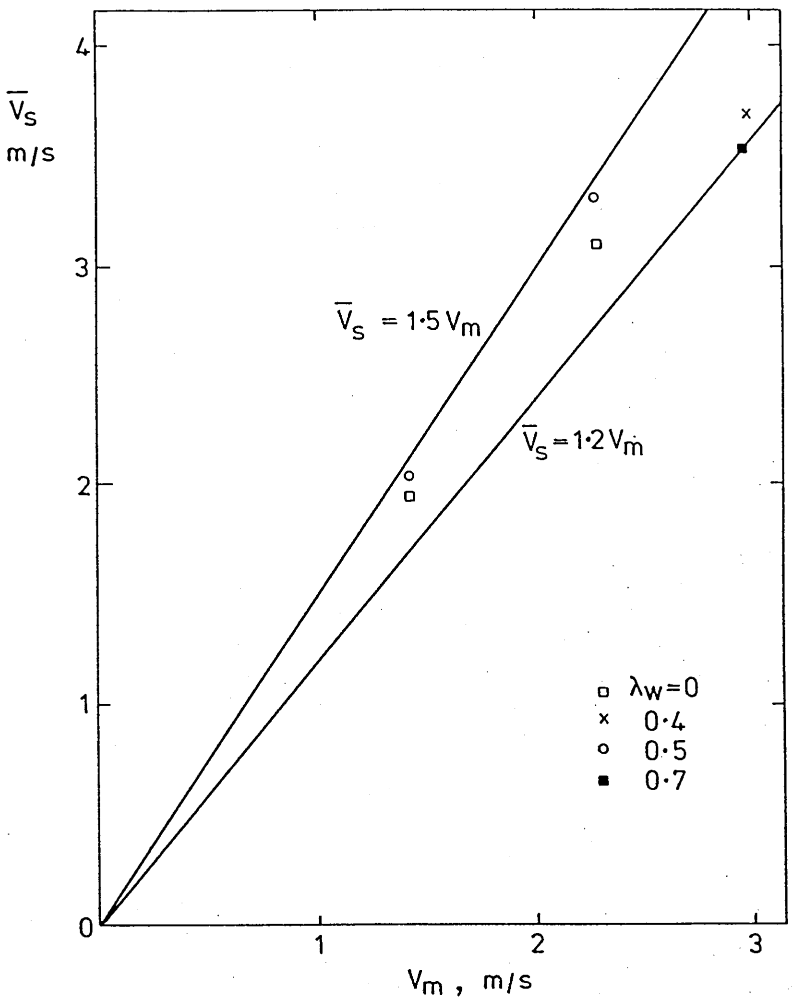


Fig 8.41 Average Slug Front Velocity,Oil No1

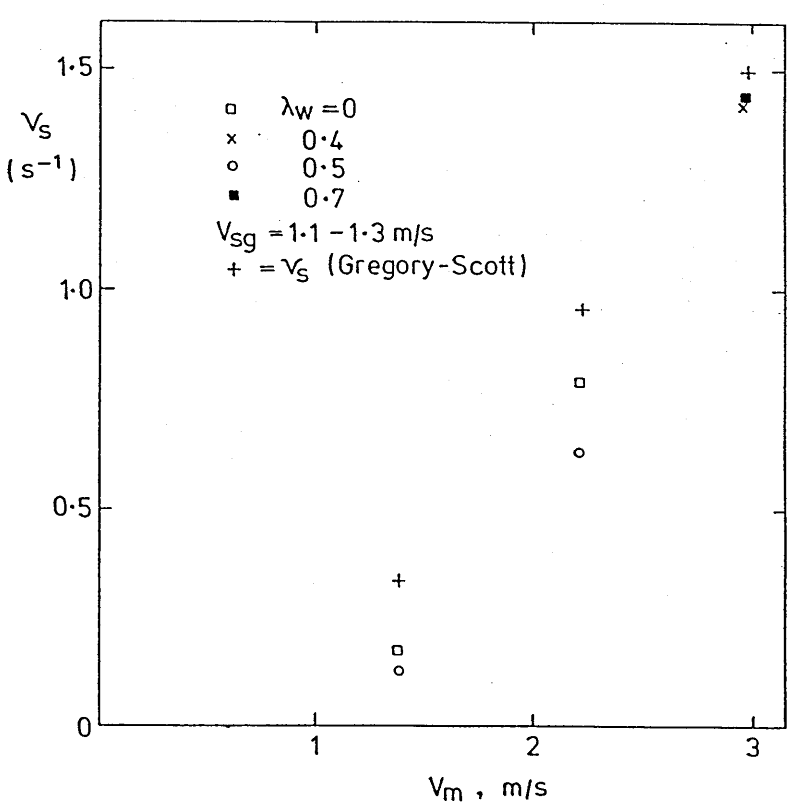


Fig 8.42 Average Slug Frequency,Oil No1

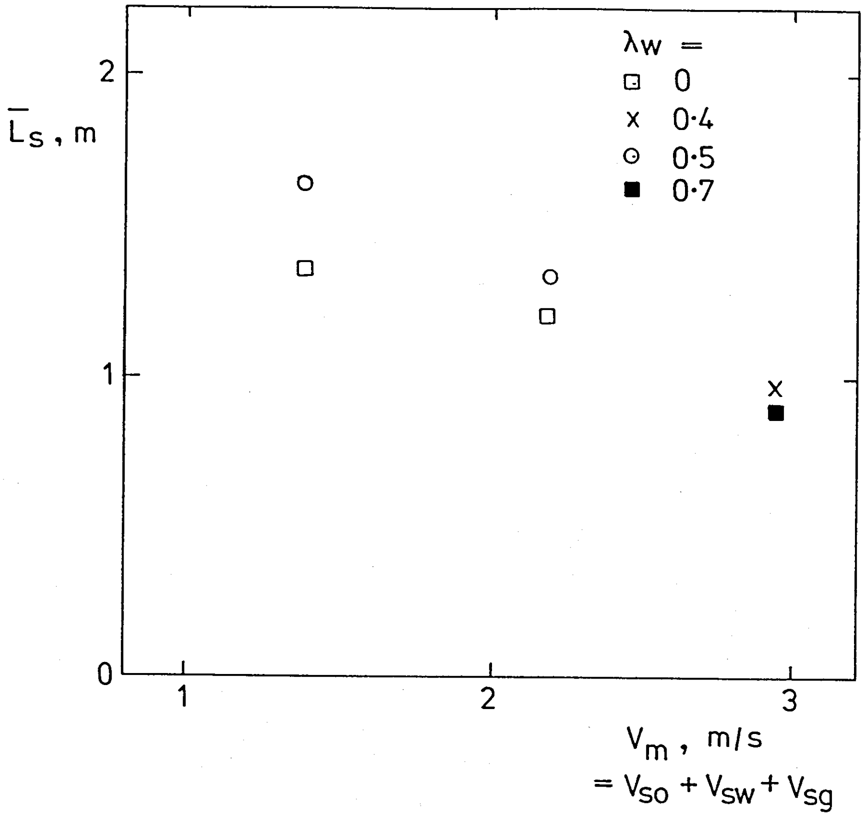


Fig 8.43 Average Slug Length,Oil No1

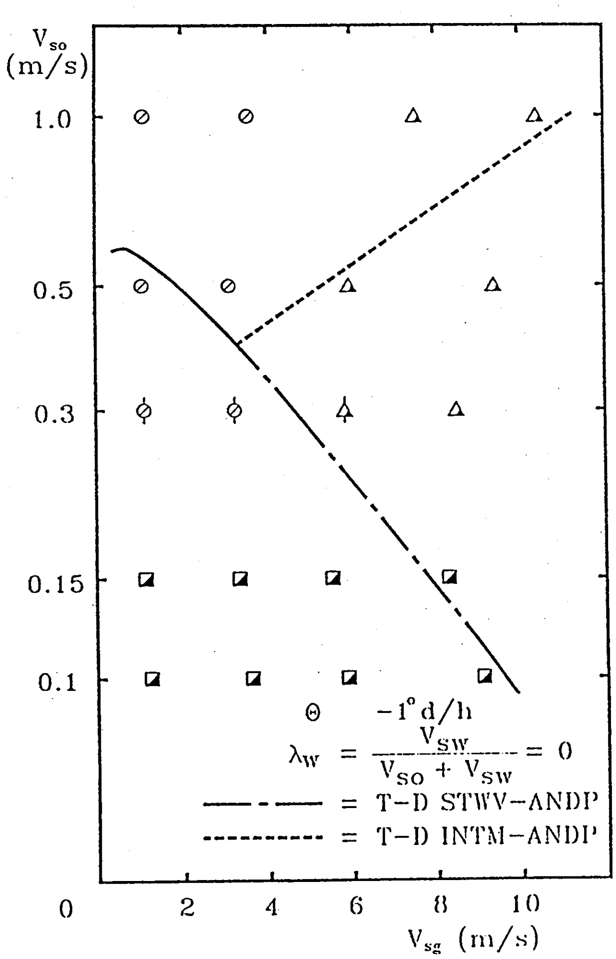


Fig 8.44 Oil/Water/Gas Flow Pattern,1 deg d/h,Oil No1

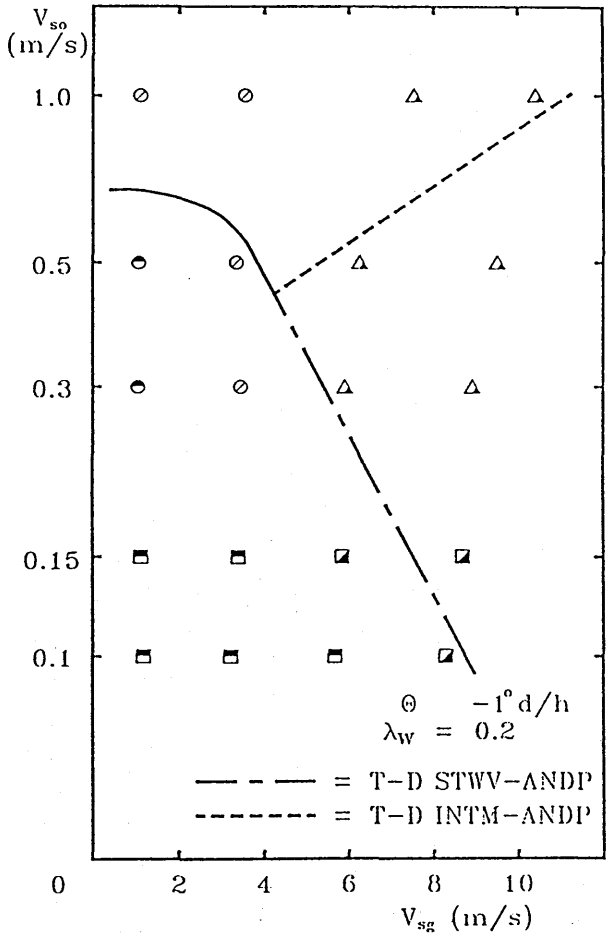


Fig 8.45 Oil/Water/Gas Flow Pattern,1 deg d/h,Oil No1

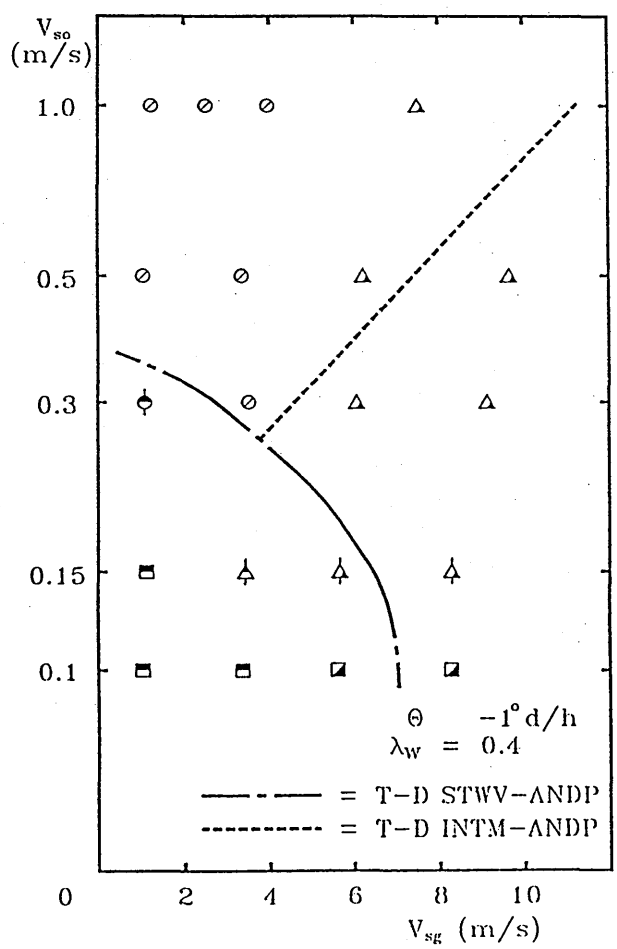


Fig 8.46 Oil/Water/Gas Flow Pattern,1 deg d/h,Oil No1

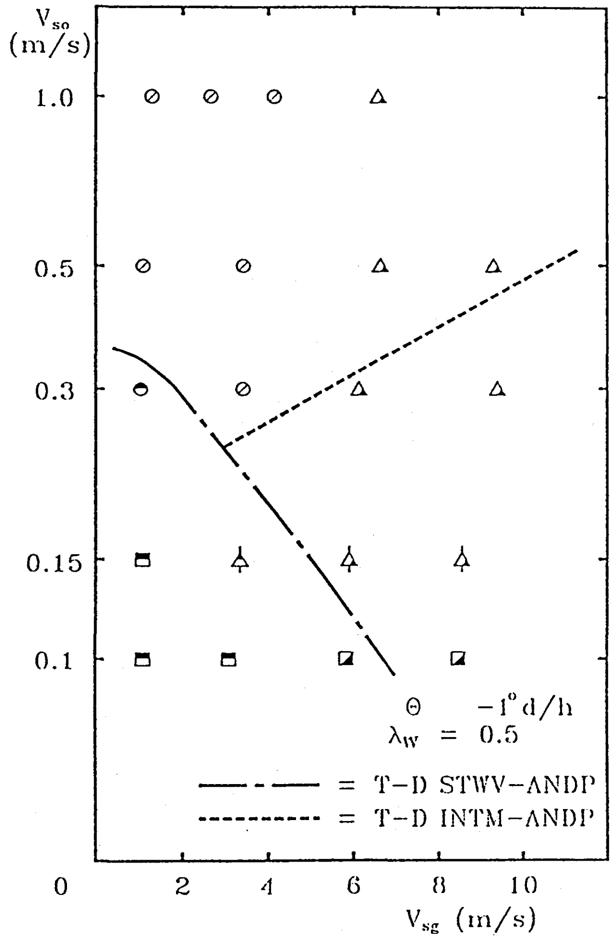


Fig 8.47 Oil/Water/Gas Flow Pattern,1 deg d/h,Oil No1

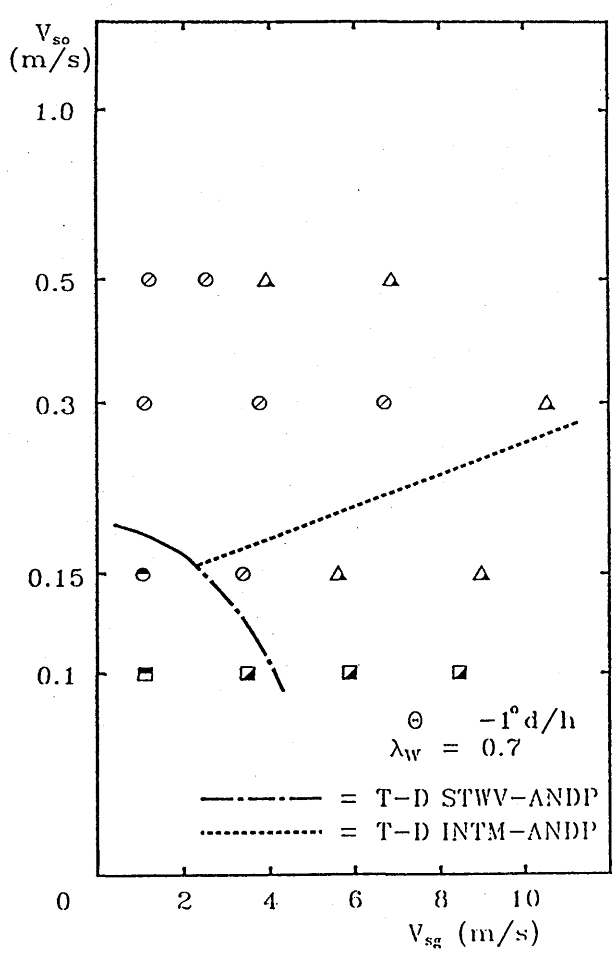


Fig 8.48 Oil/Water/Gas Flow Pattern,1 deg d/h,Oil No1

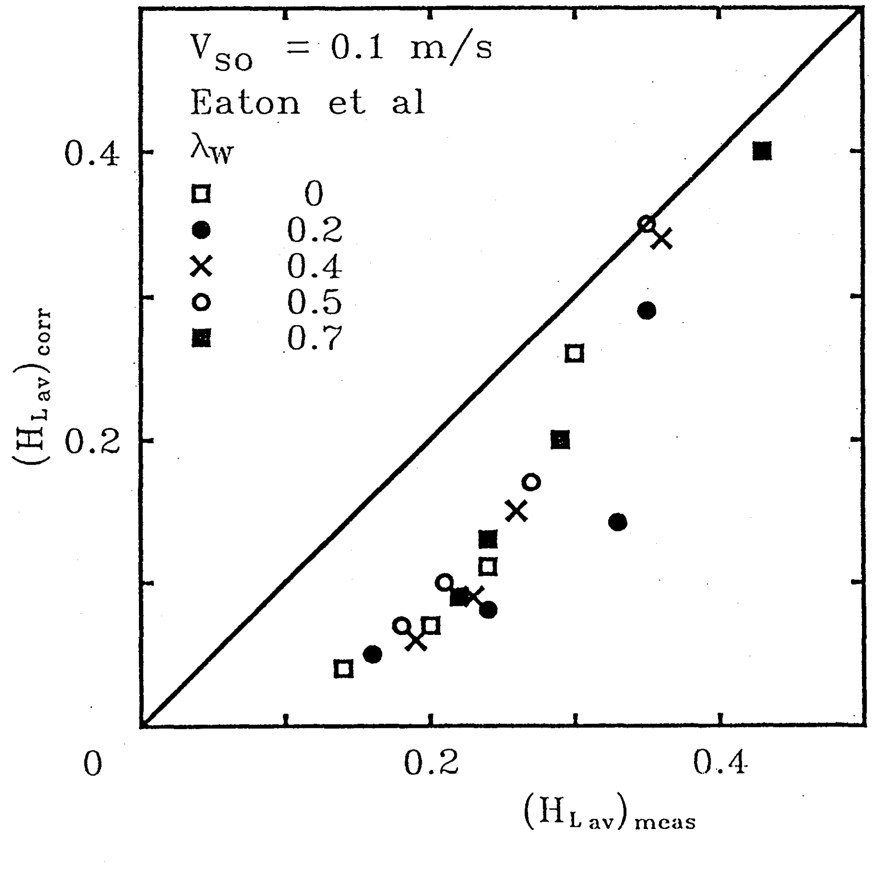


Fig 8.49 Liquid Holdup Compared to Eaton et al correlation,1 deg u/h,Oil No1

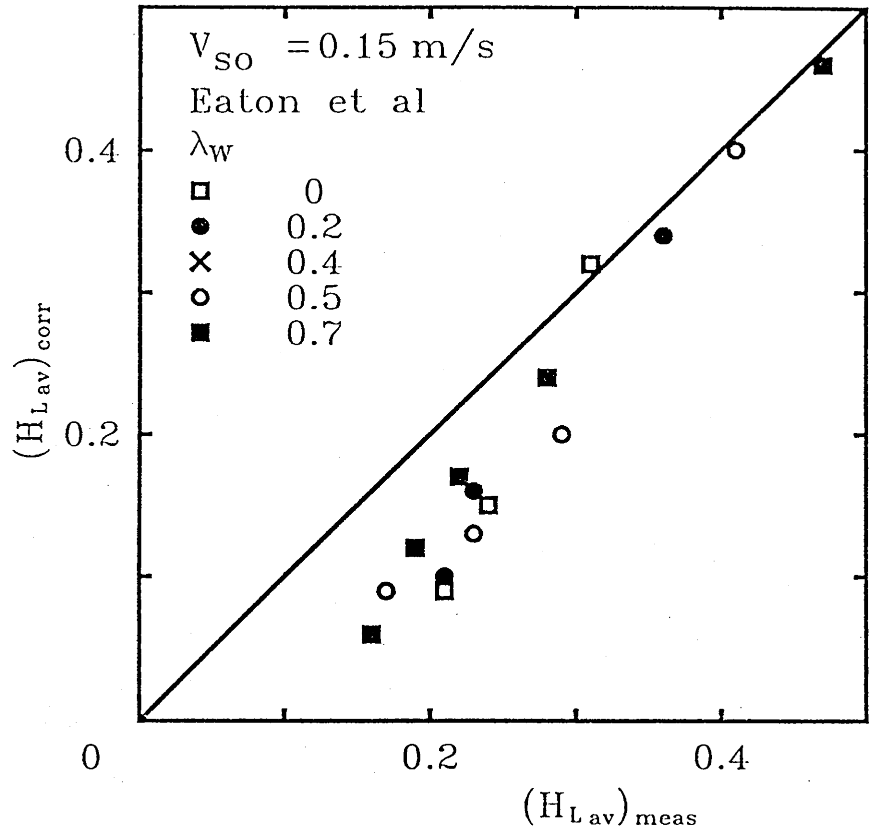


Fig 8.50 Liquid Holdup Compared to Eaton et al correlation,1 deg u/h,Oil No1

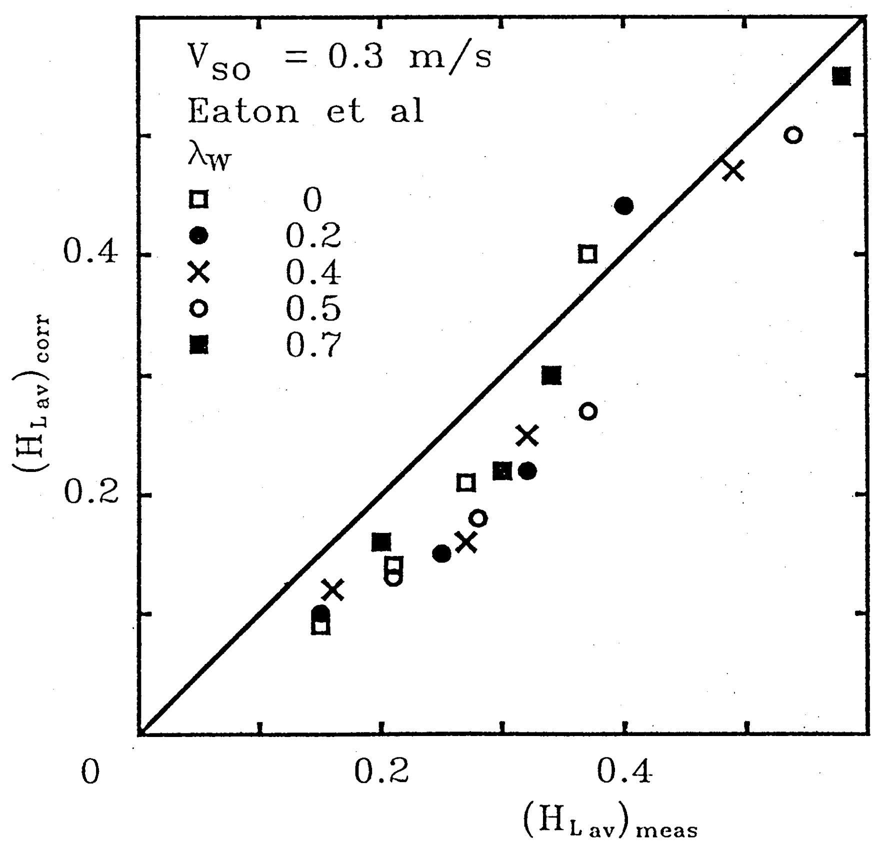


Fig 8.51 Liquid Holdup Compared to Eaton et al correlation,1 deg u/h,Oil No1

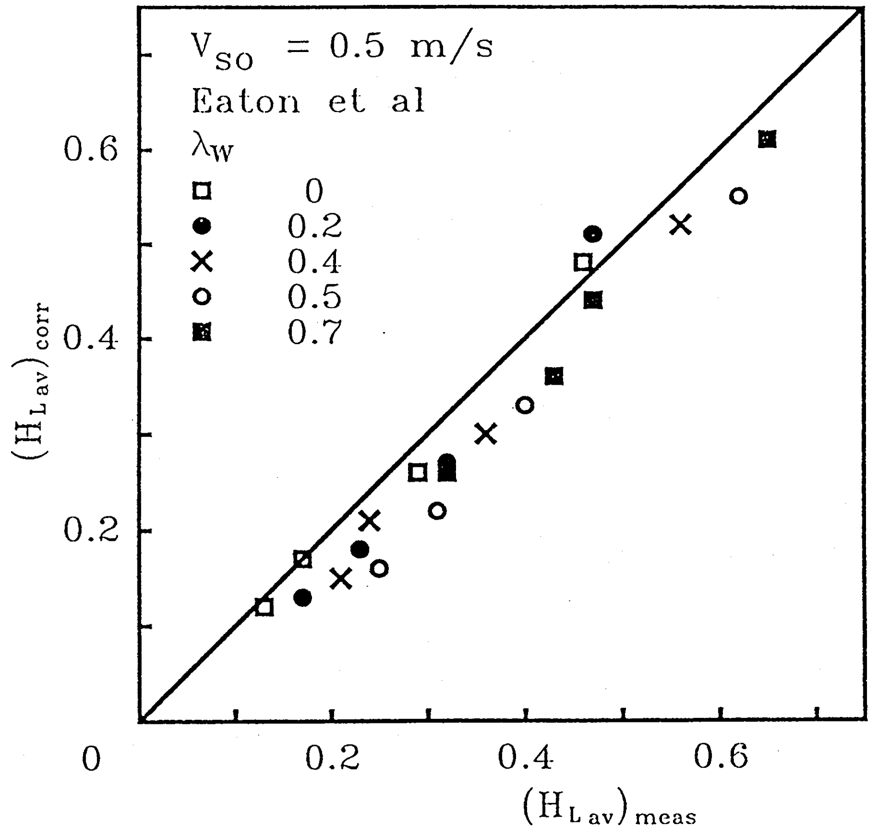


Fig 8.52 Liquid Holdup Compared to Eaton et al correlation,1 deg u/h,Oil No1

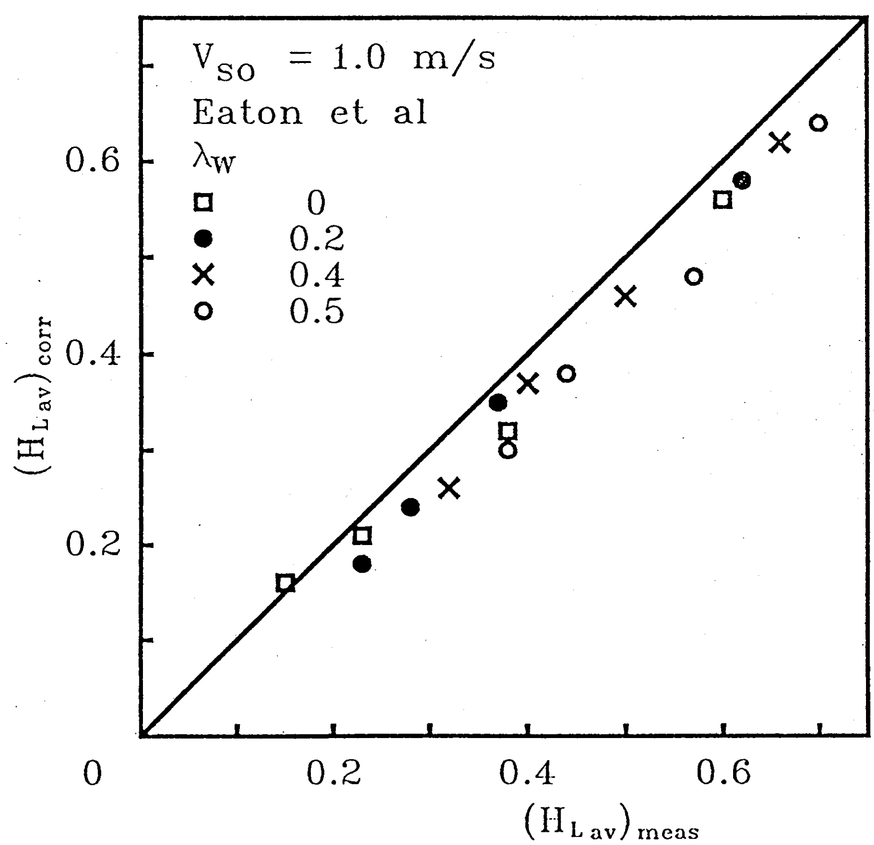


Fig 8.53 Liquid Holdup Compared to Eaton et al correlation,1 deg u/h,Oil No1

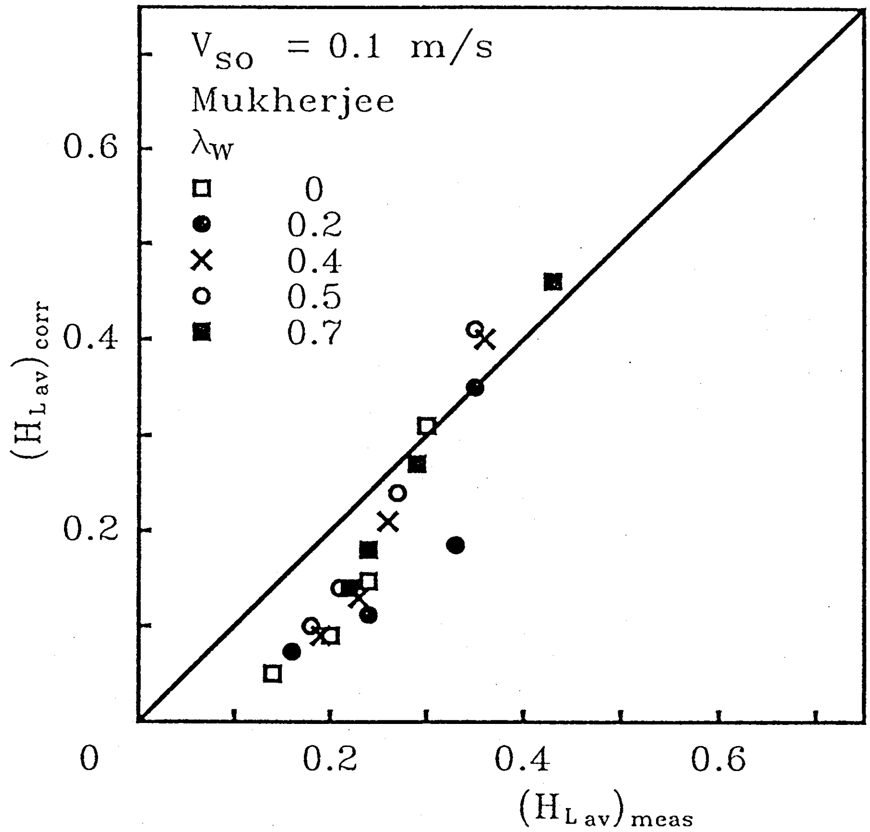


Fig 8.54 Liquid Holdup Compared to Mukherjee-Brill correlation,1 deg u/h,Oil No1

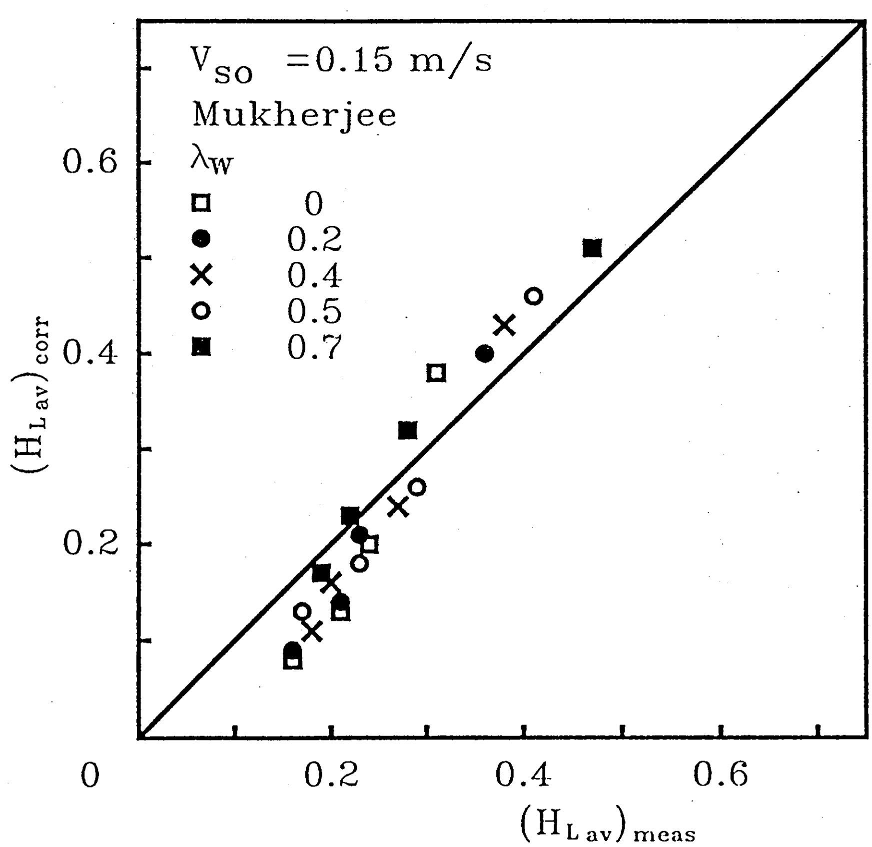


Fig 8.55 Liquid Holdup Compared to Mukherjee-Brill correlation,1 deg u/h,Oil No1

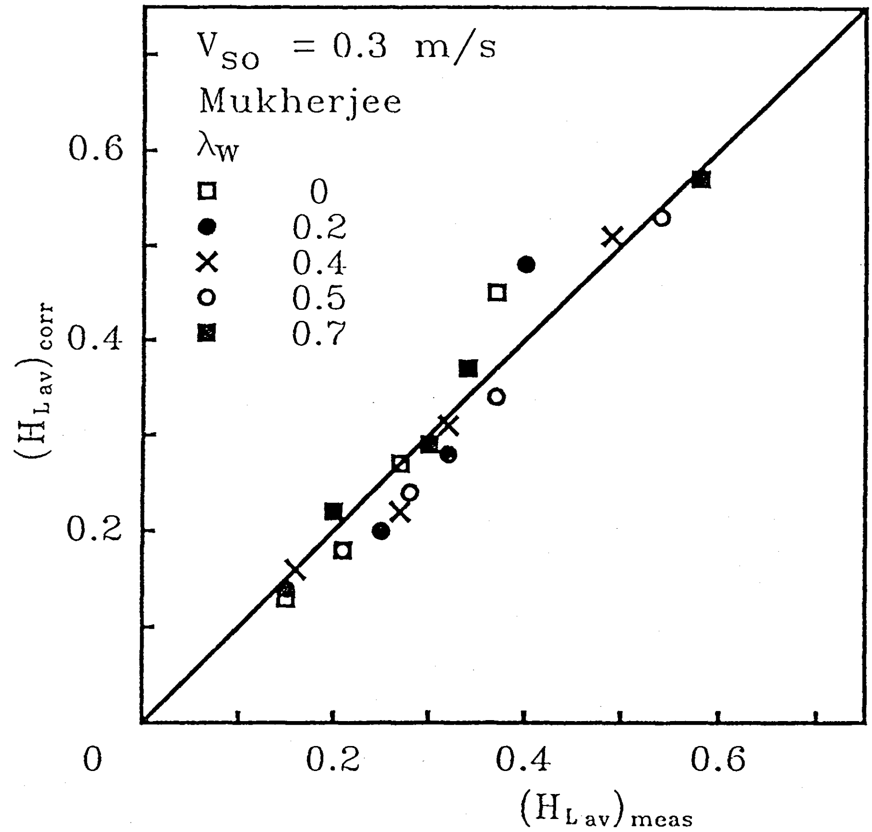


Fig 8.56 Liquid Holdup Compared to Mukherjee-Brill correlation,1 deg u/h,Oil No1

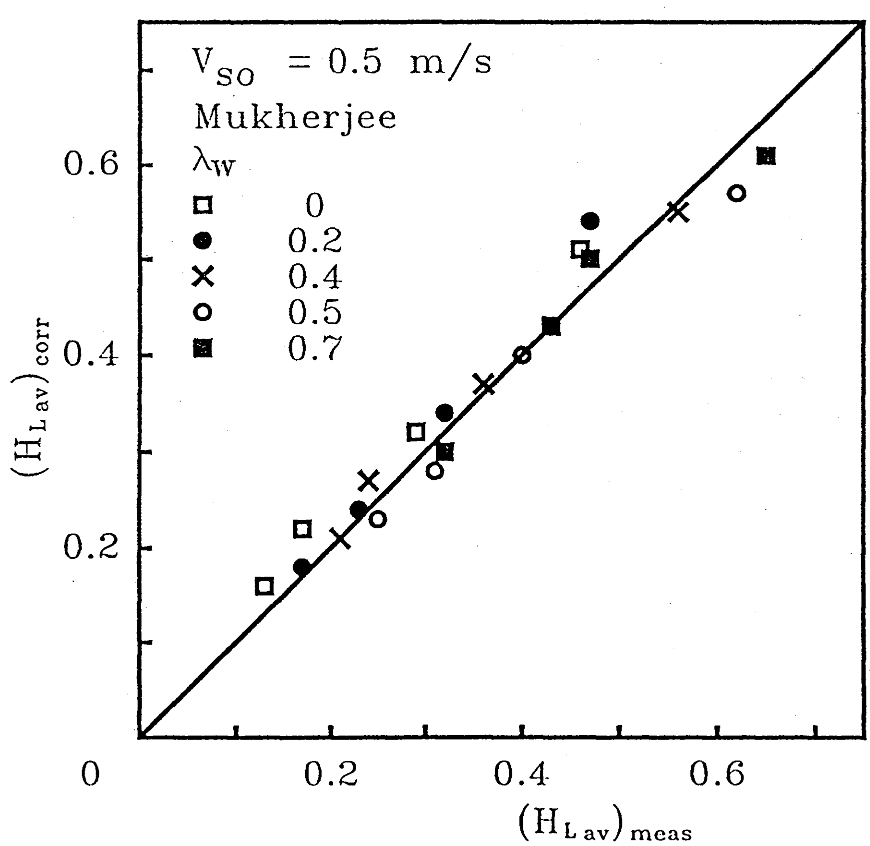


Fig 8.57 Liquid Holdup Compared to Mukherjee-Brill correlation,1 deg u/h,Oil No1

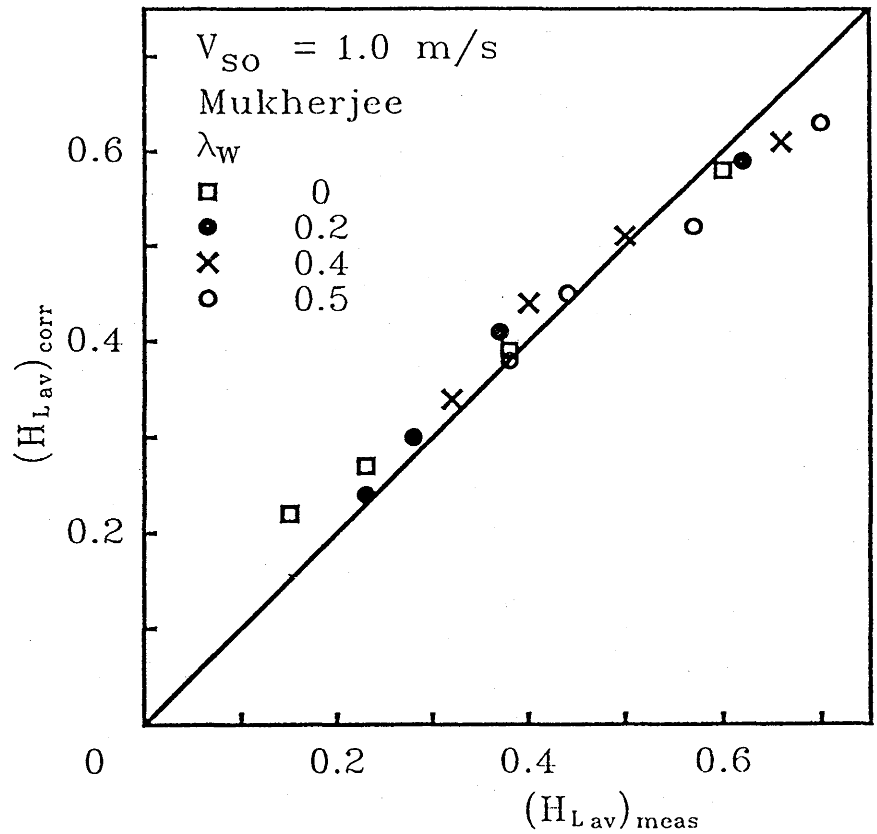


Fig 8.58 Liquid Holdup Compared to Mukherjee-Brill correlation,1 deg u/h,Oil No1

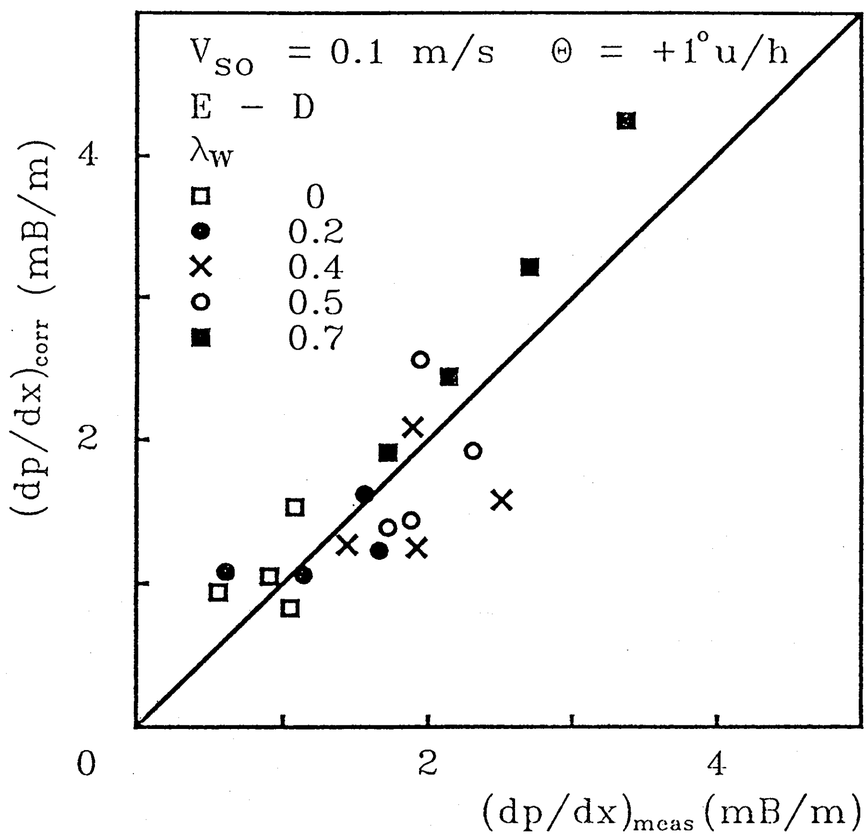


Fig 8.59 Pressure Loss Compared to Eaton-Dukler correlation,1 deg u/h,Oil No1

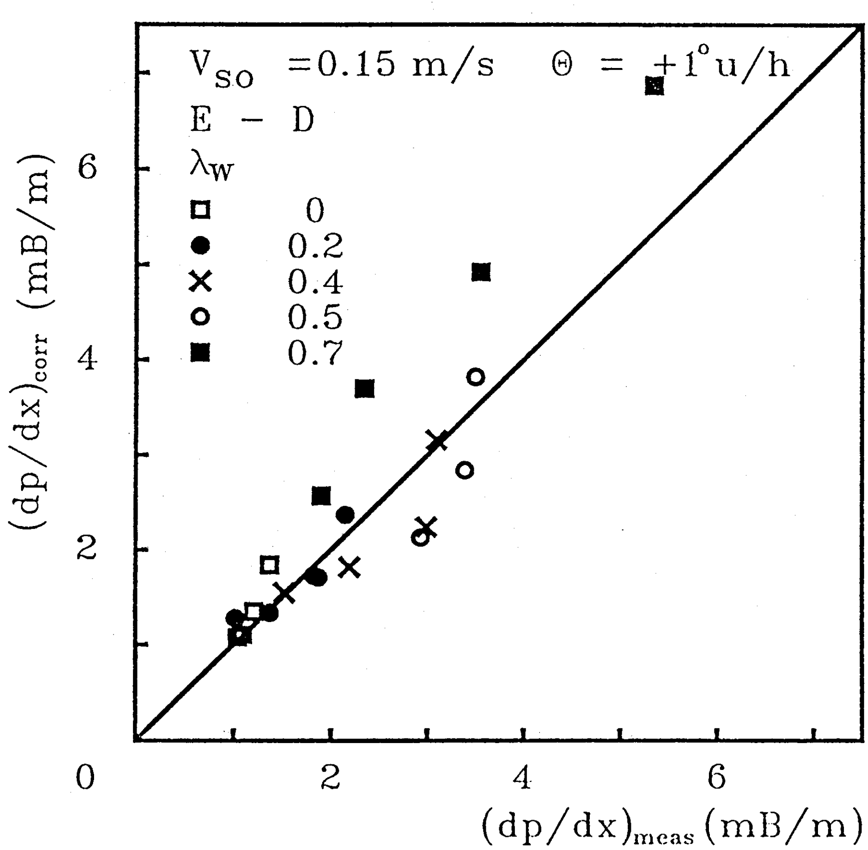


Fig 8.60 Pressure Loss Compared to Eaton-Dukler correlation,1 deg u/h,Oil No1

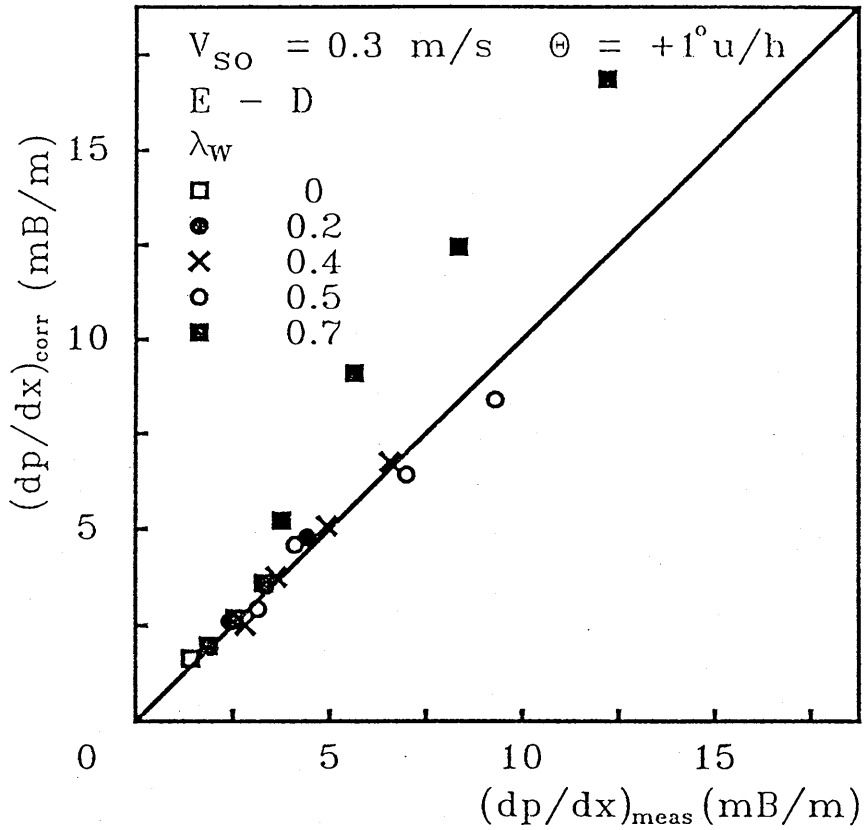


Fig 8.61 Pressure Loss Compared to Eaton-Dukler correlation,1 deg u/h,Oil No1

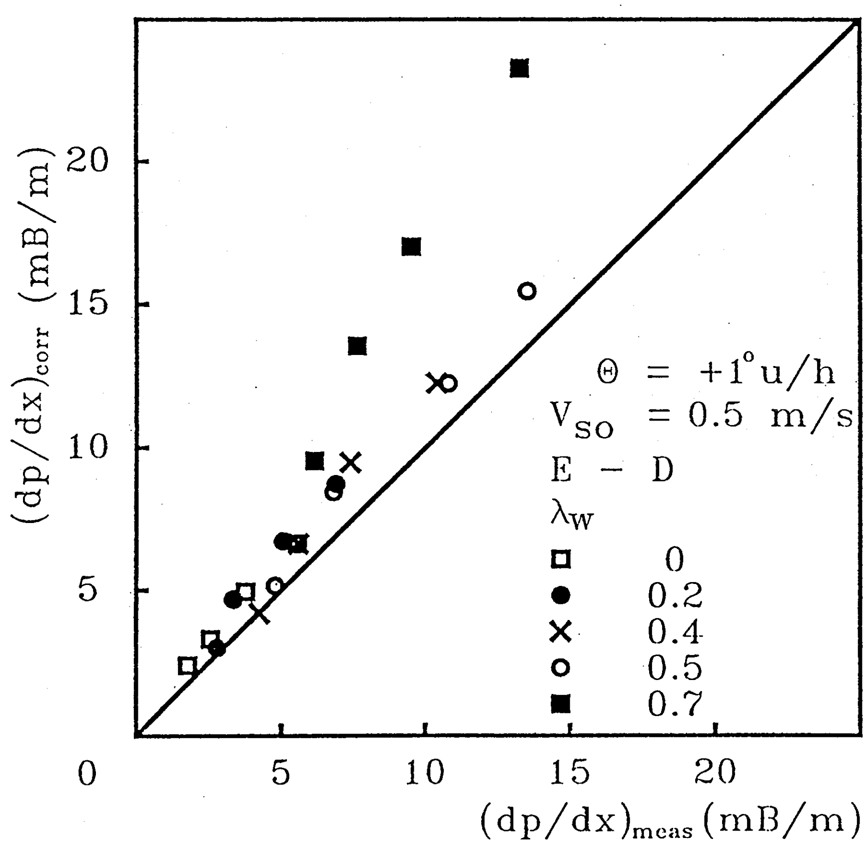


Fig 8.62 Pressure Loss Compared to Eaton-Dukler correlation,1 deg u/h,Oil No1

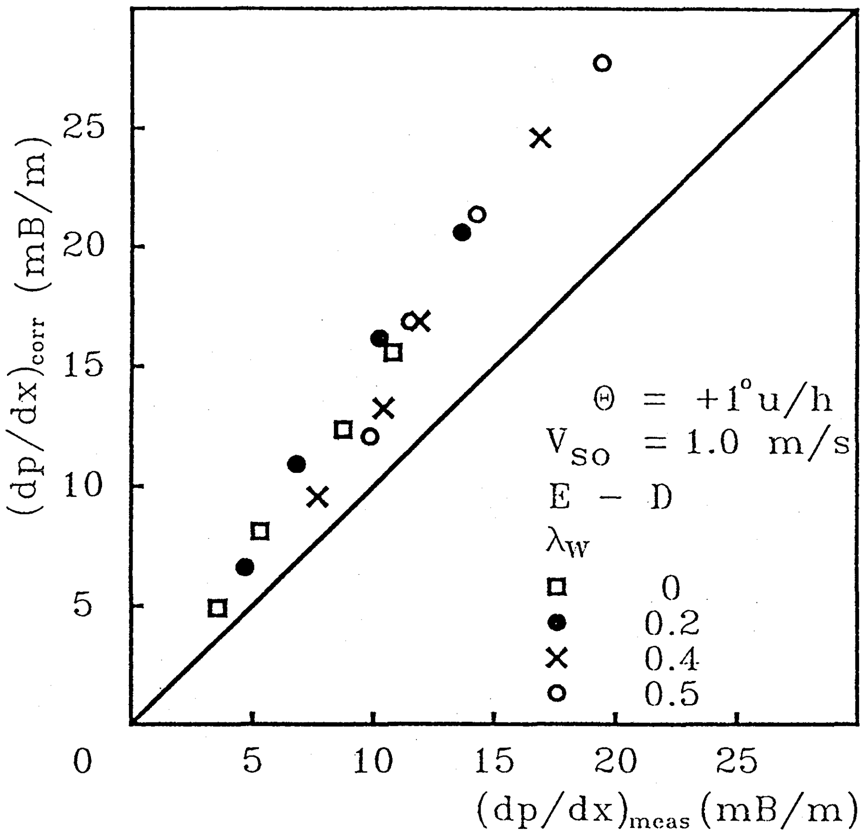


Fig 8.63 Pressure Loss Compared to Eaton-Dukler correlation,1 deg u/h,Oil No1

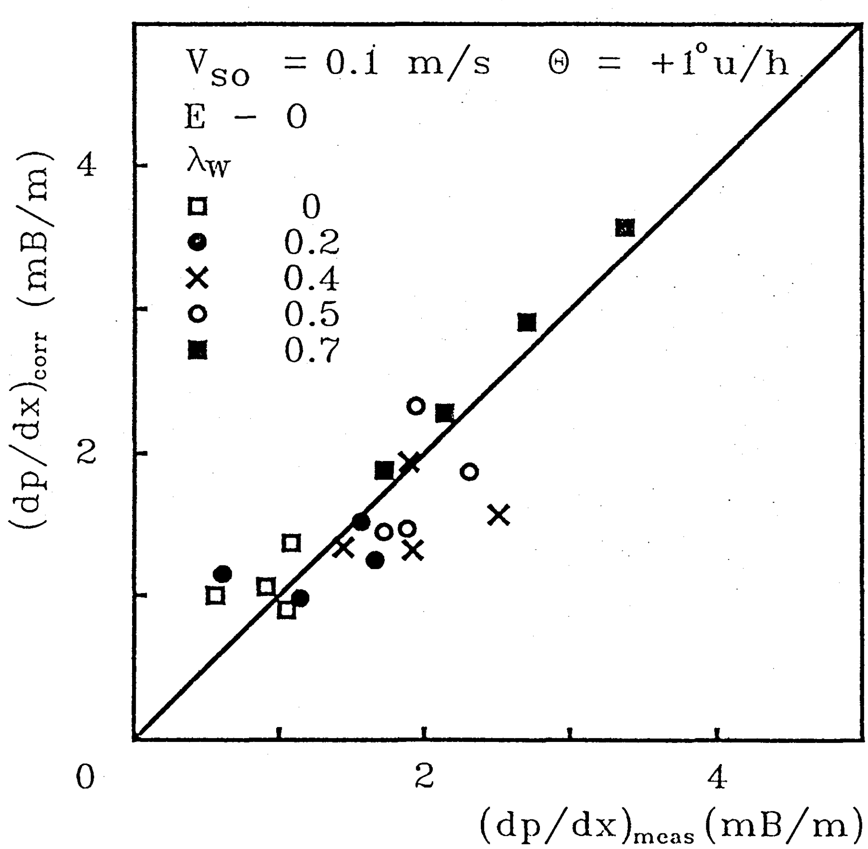


Fig 8.64 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg u/h,Oil No1

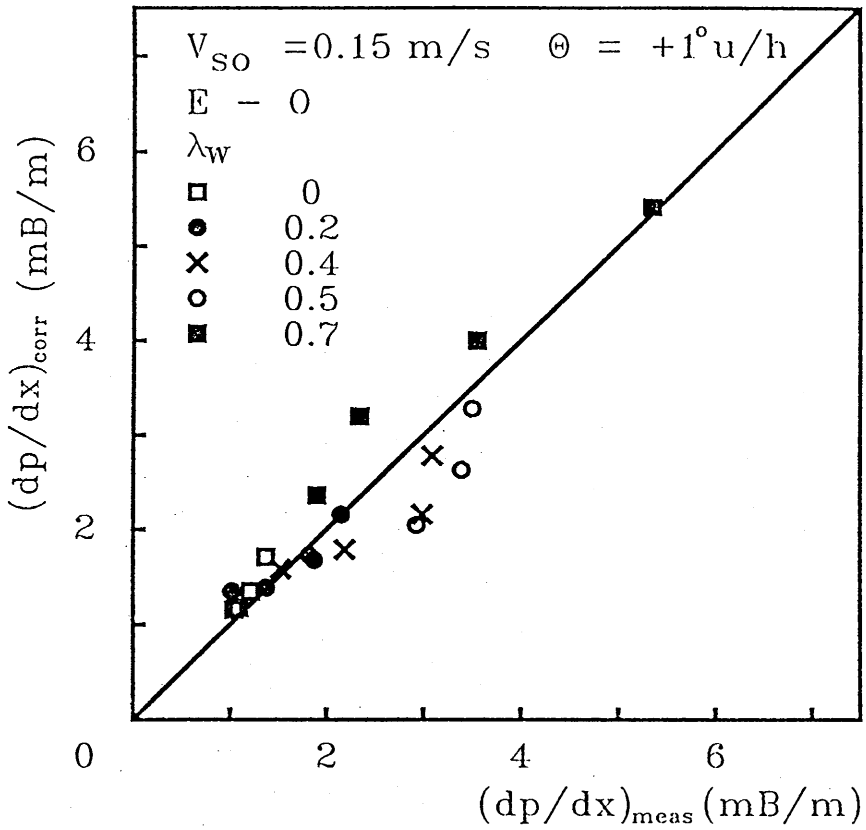


Fig 8.65 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg u/h,Oil No1

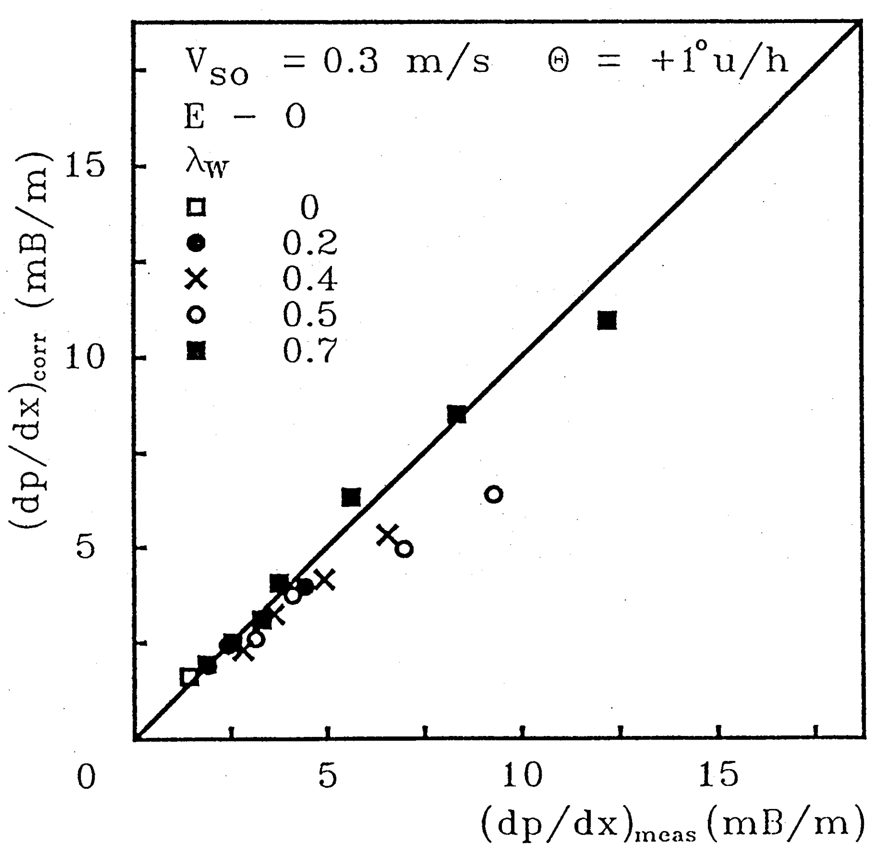


Fig 8.66 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg u/h,Oil No1

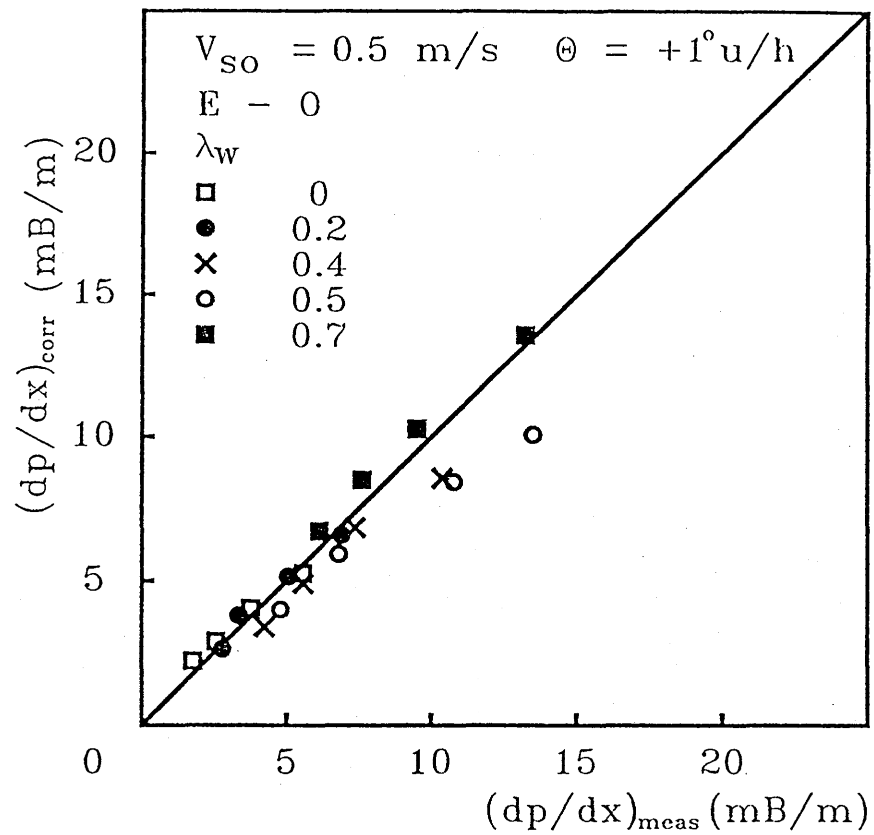


Fig 8.67 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg u/h,Oil No1

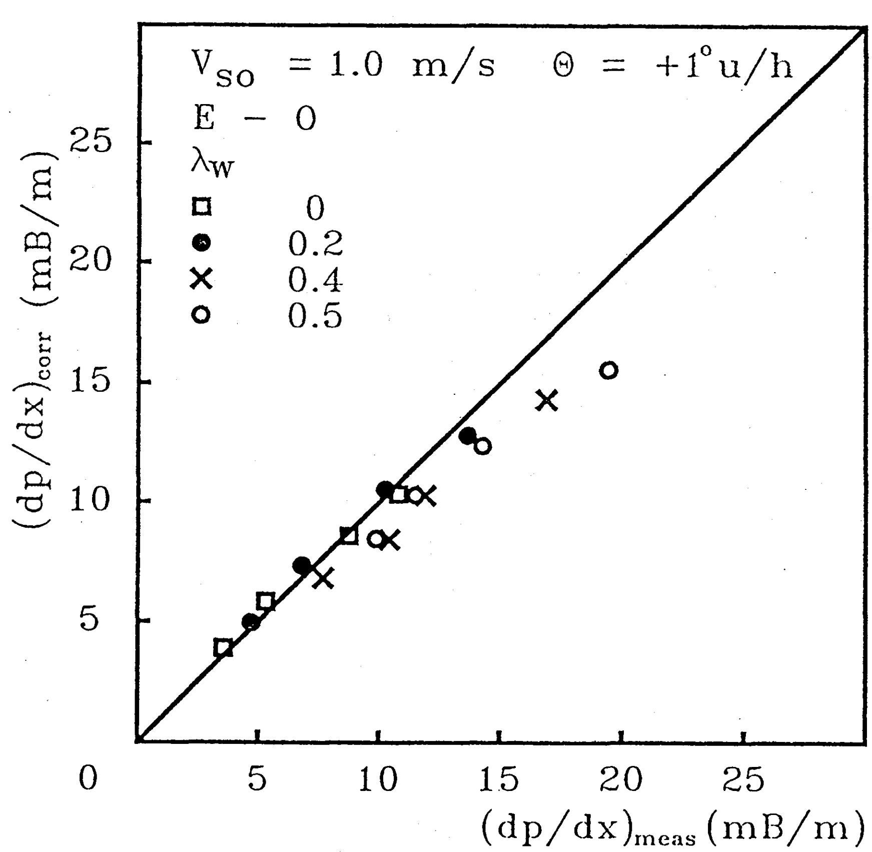


Fig 8.68 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg u/h,Oil No1

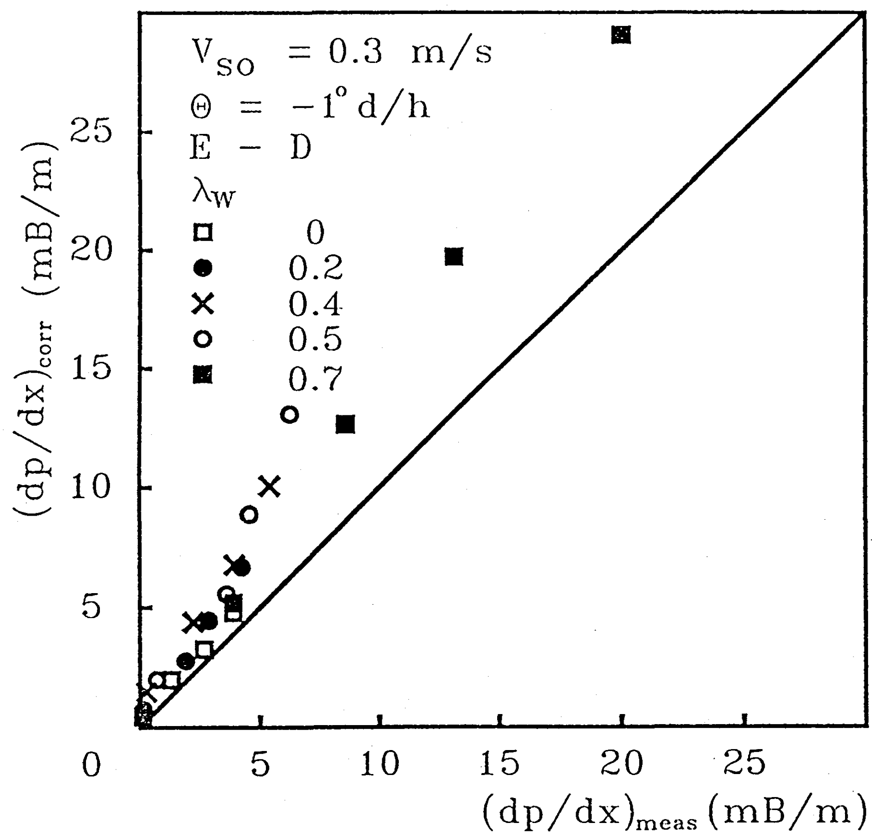


Fig 8.69 Pressure Loss Compared to Eaton-Dukler correlation,1 deg d/h,Oil No1

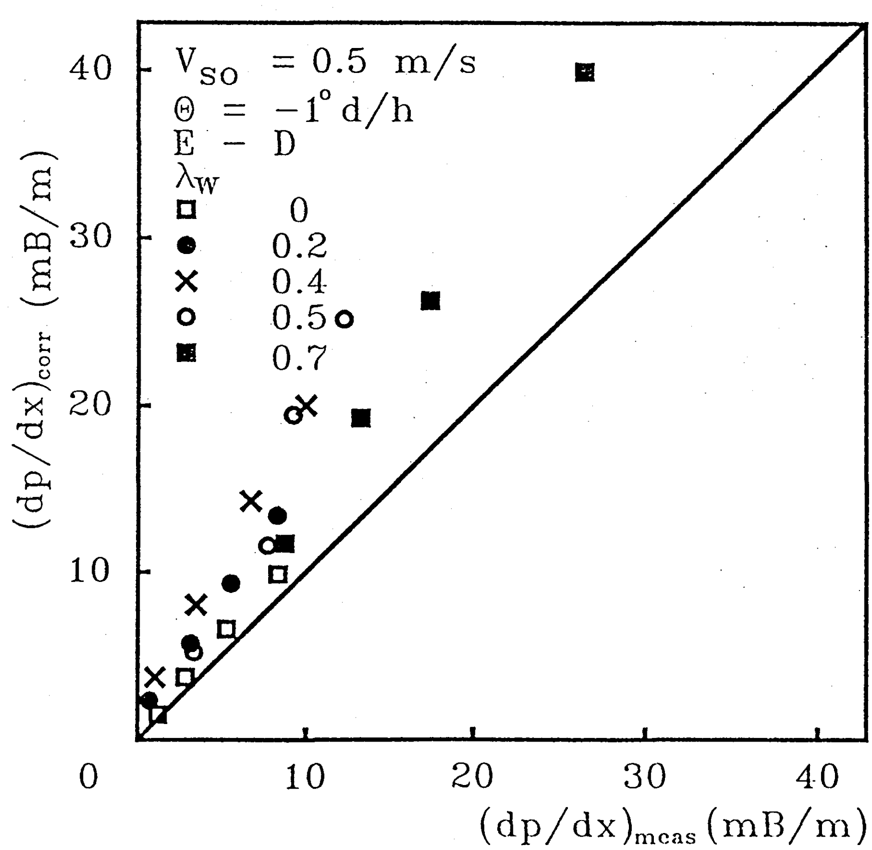


Fig 8.70 Pressure Loss Compared to Eaton-Dukler correlation,1 deg d/h,Oil No1

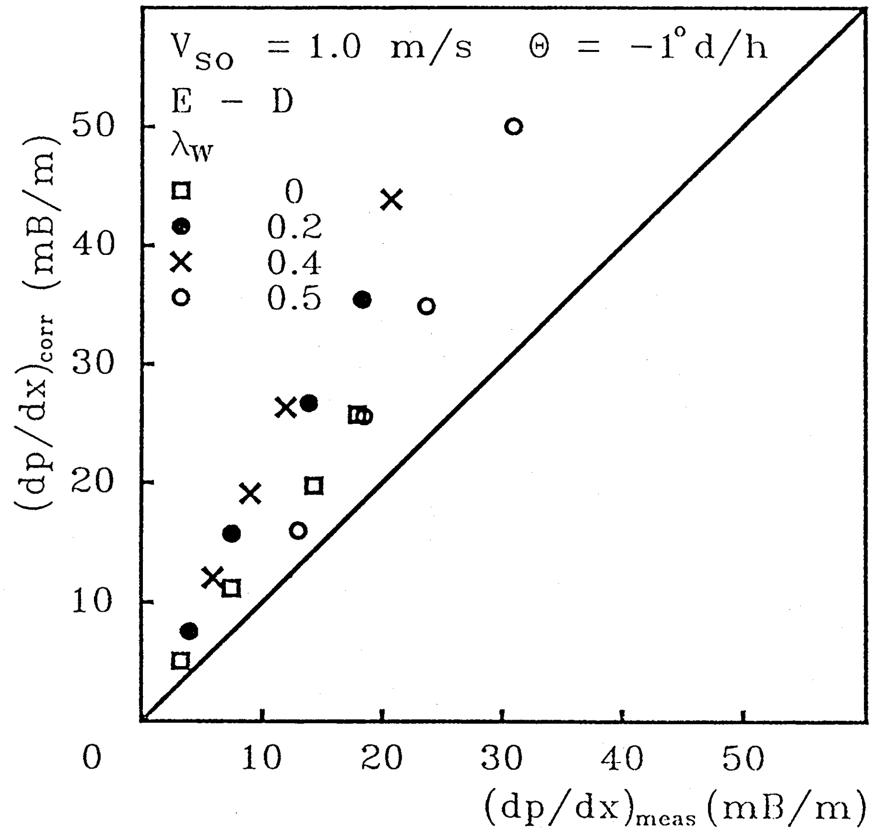


Fig 8.71 Pressure Loss Compared to Eaton-Dukler correlation,1 deg d/h,Oil No1

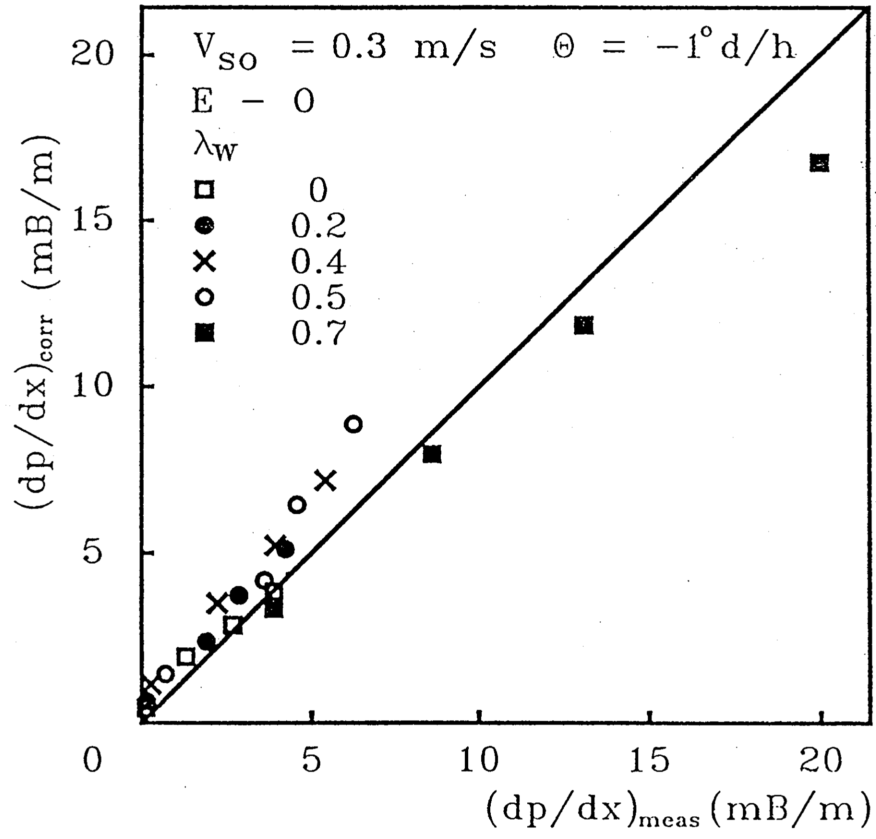


Fig 8.72 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg d/h,Oil No1

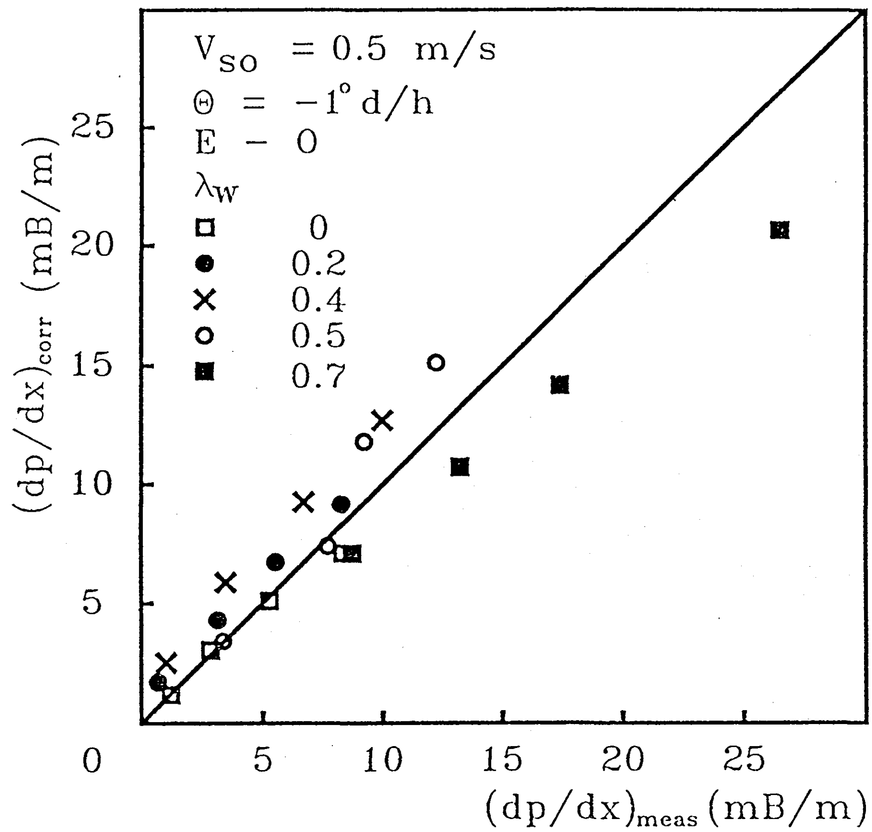


Fig 8.73 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg d/h,Oil No1

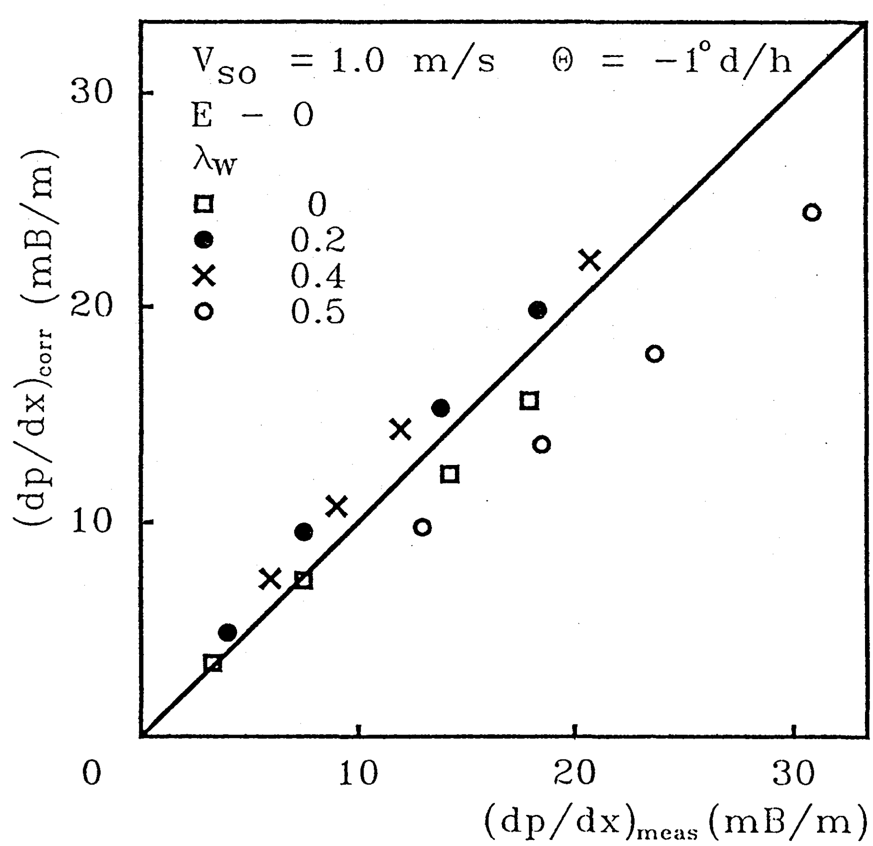


Fig 8.74 Pressure Loss Compared to Eaton-Oliemans correlation,1 deg d/h,Oil No1

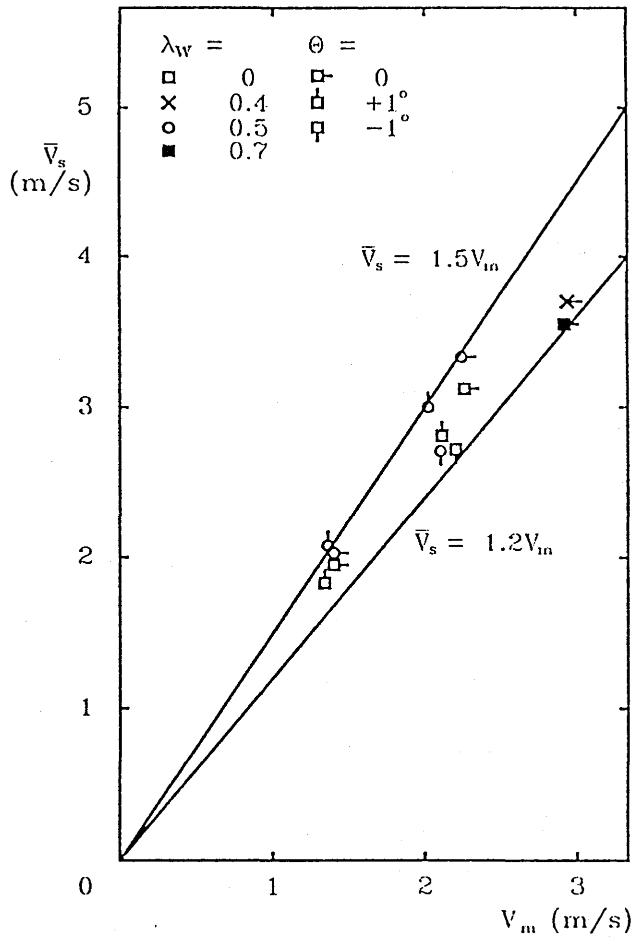


Fig 8.75 Average Slug Front Velocity,Inclined Flow,Oil No1

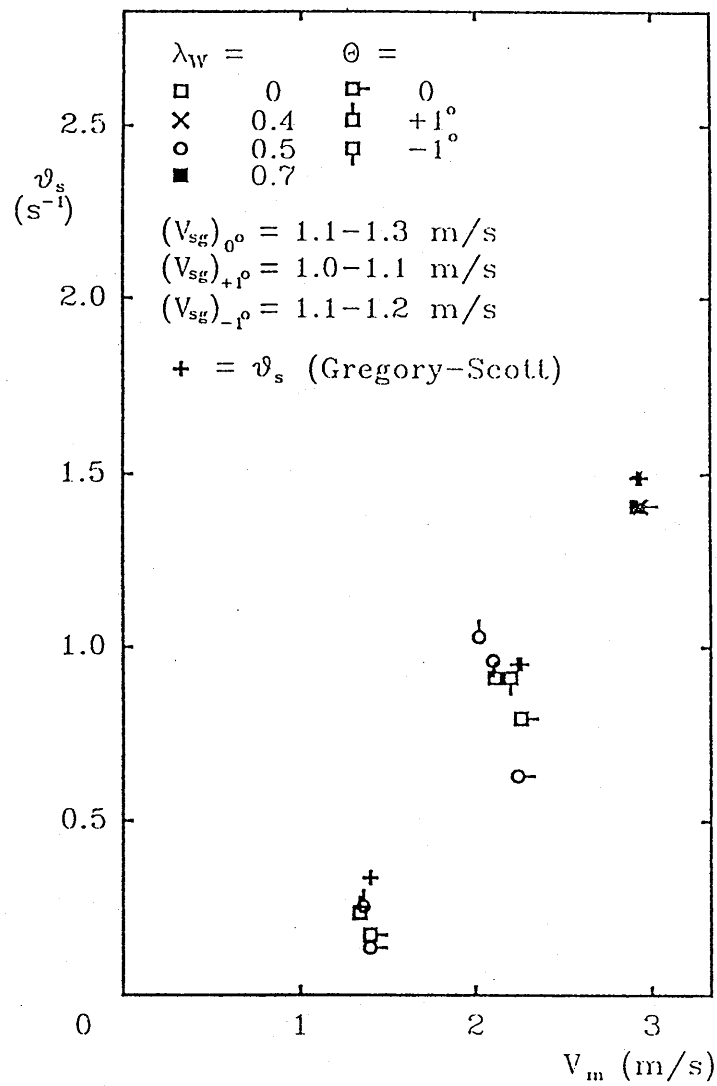


Fig 8.76 Average Slug Frequency,Inclined Flow,Oil No1

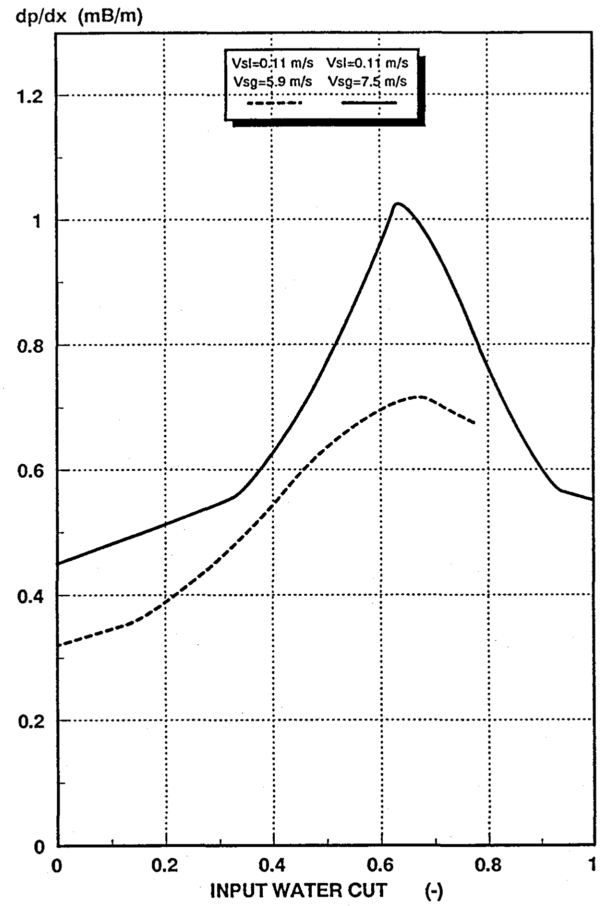


Fig 8.77 Oil/Water/Gas Pressure Drop Data of Sobocinski(1959)

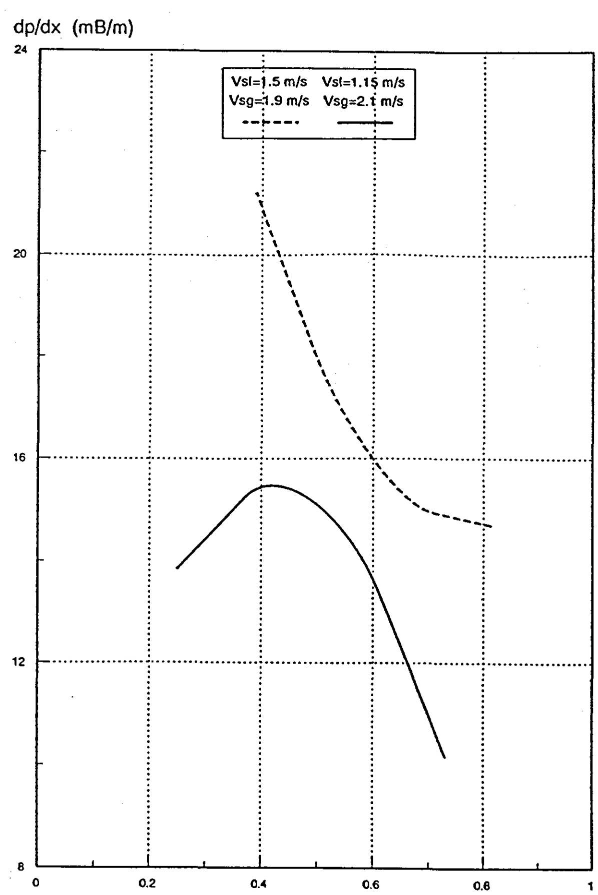


Fig 8.78 Oil/Water/Gas Pressure Drop Data of Malinowsky(1975)

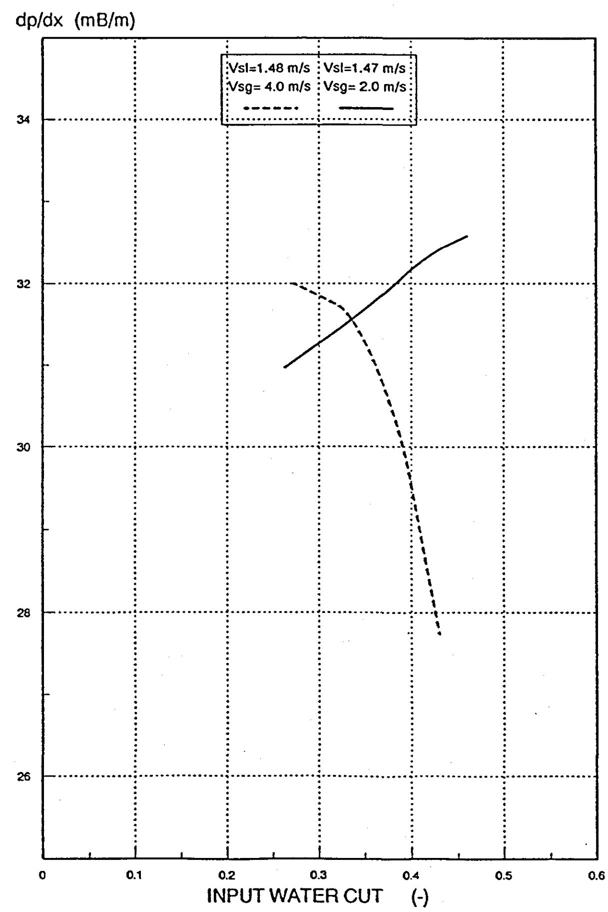


Fig 8.79 Oil/Water/Gas Pressure Drop Data of Laflin-Oglesby(1976)

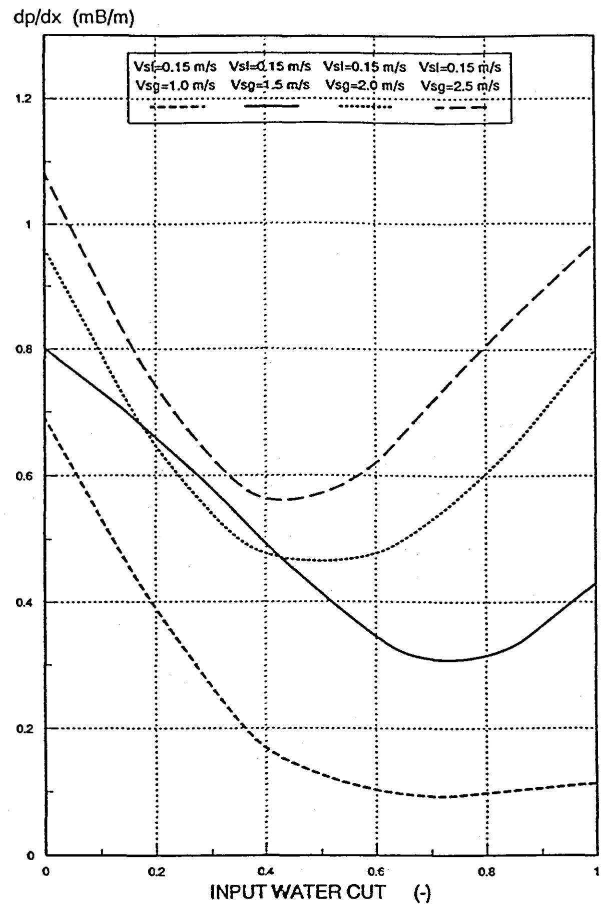


Fig 8.80 Oil/Water/Gas Pressure Drop Data of Stapelberg et al(1991)

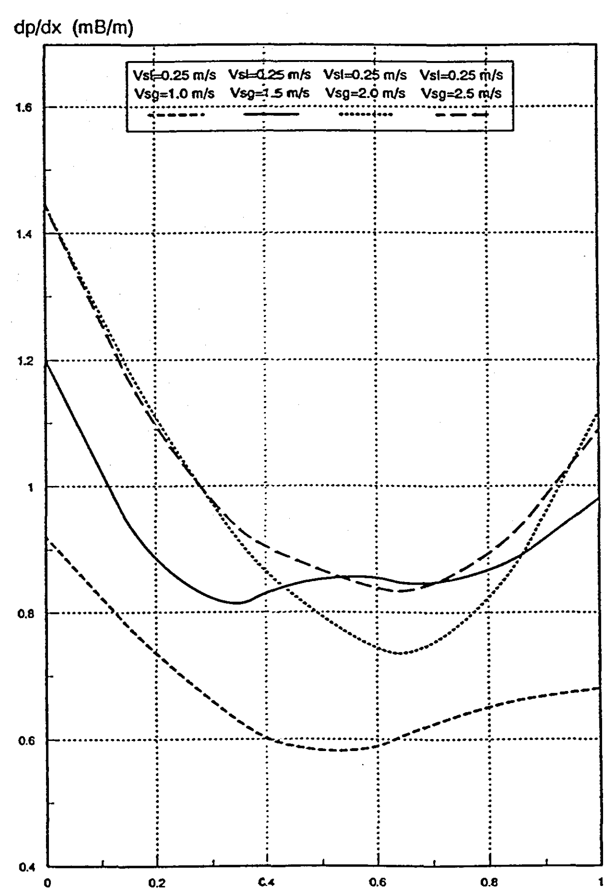


Fig 8.81 Oil/Water/Gas Pressure Drop Data of Stapelberg et al(1991)

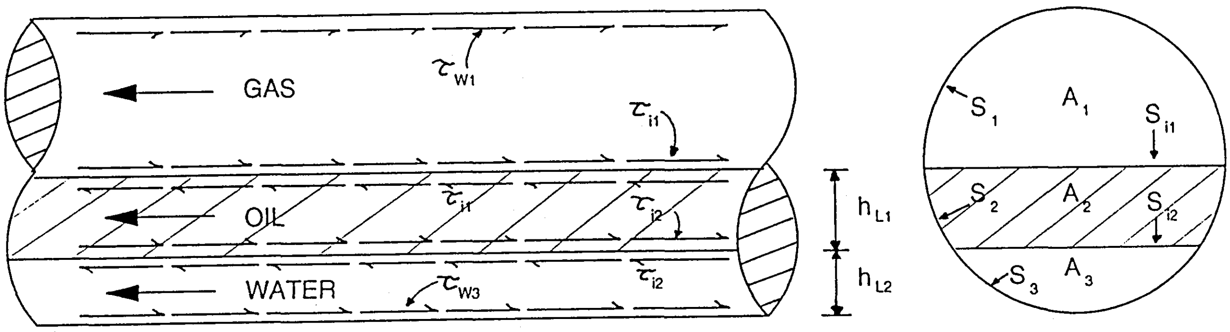
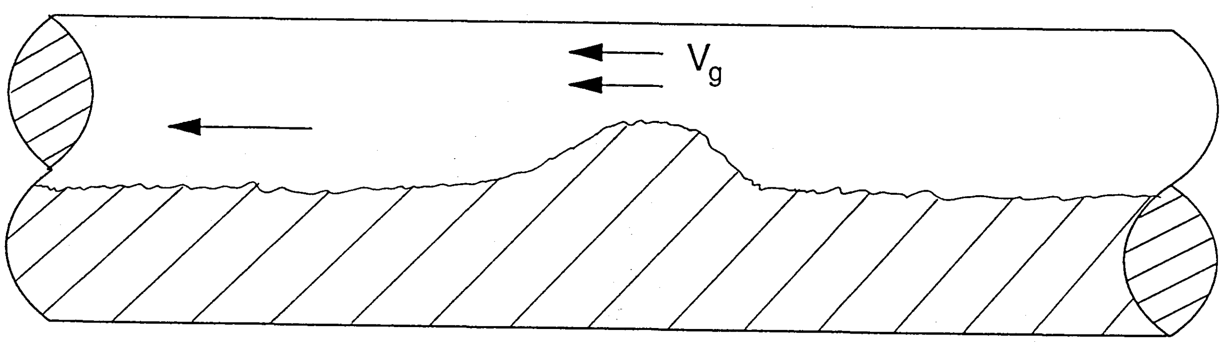
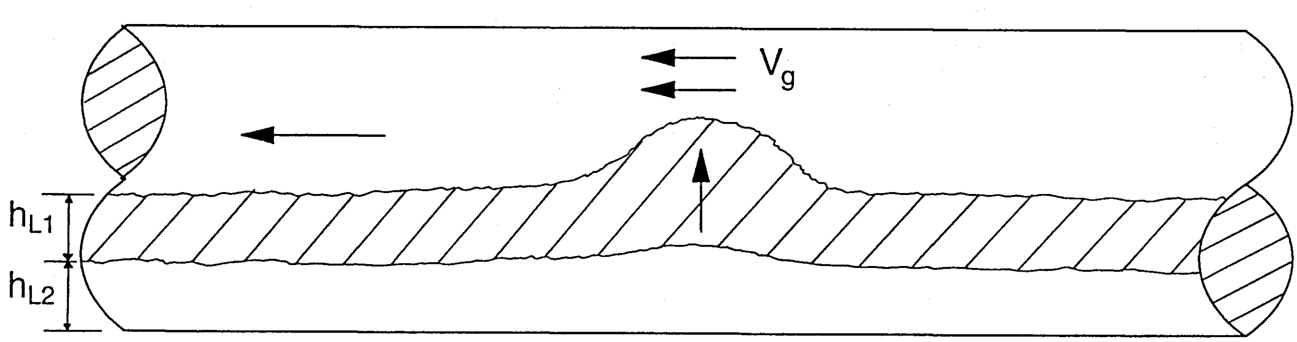


Fig 8.82 Oil/Water/Gas Three-Layer Stratified Flow

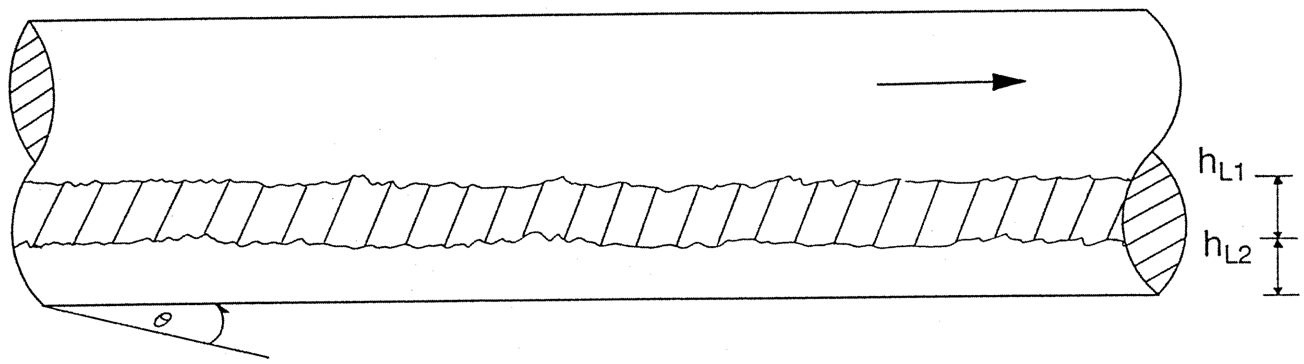


(a) Oil-only Liquid Film

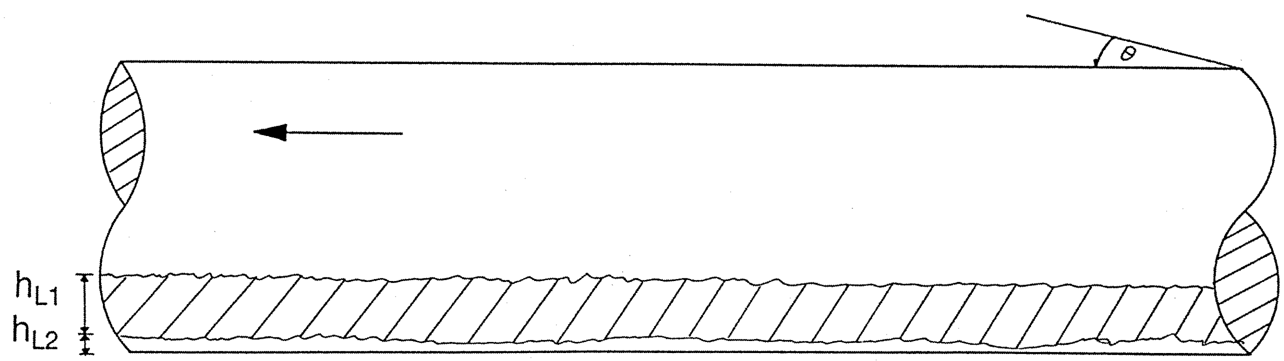


(b) Extension to Separated Liquids Film

Fig 8.83 Slug Generation Mechanism of Taitel-Dukler (1976)

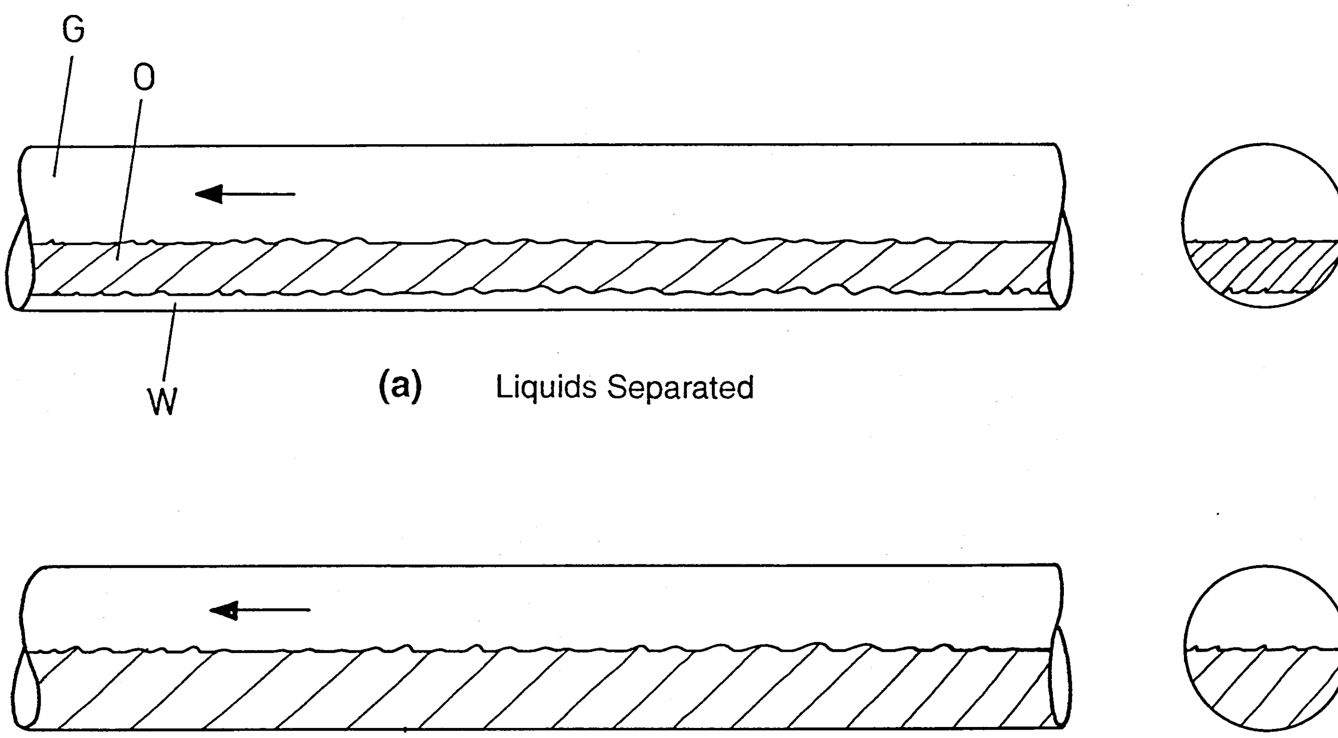


(a) Uphill Flow



(b) Downhill Flow

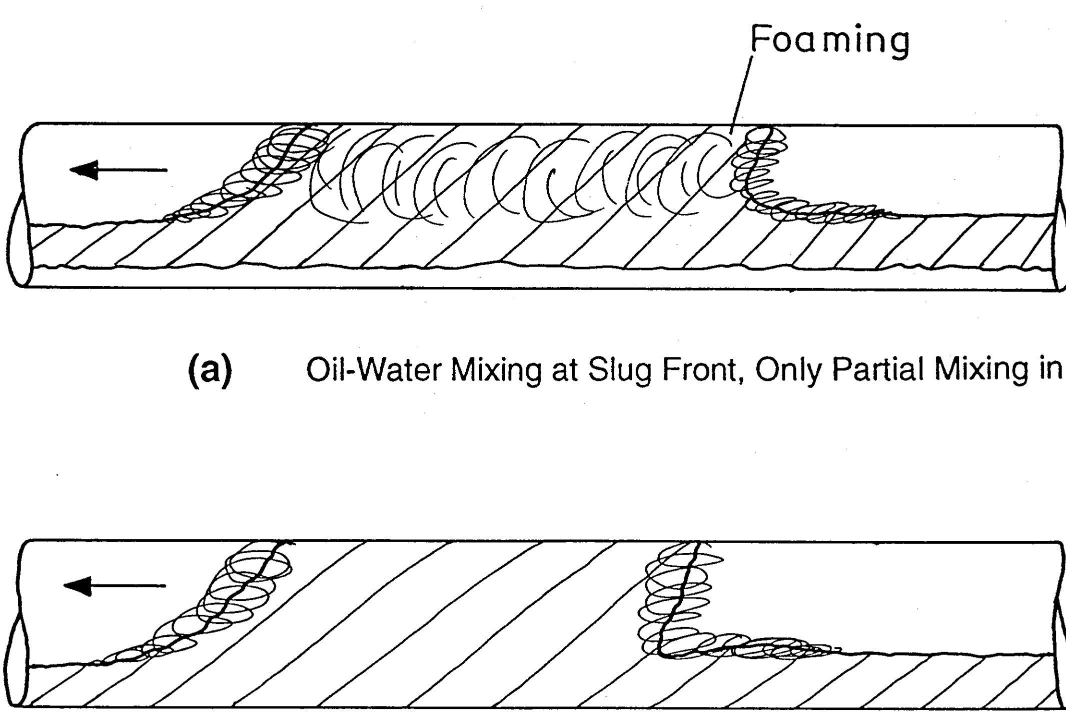
Fig 8.84 Effect of Inclination on Three-Layer Stratified Flow



(a) Liquids Separated

(b) Liquids Mixed

Fig 8.85 Three-phase Stratified Flow



(a) Oil-Water Mixing at Slug Front, Only Partial Mixing in Slug Body

(b) Oil and Water Form Homogeneous Mixture in Complete Slug Unit

Fig 8.86 Liquids Behaviour in a Three-phase Intermittent Flow

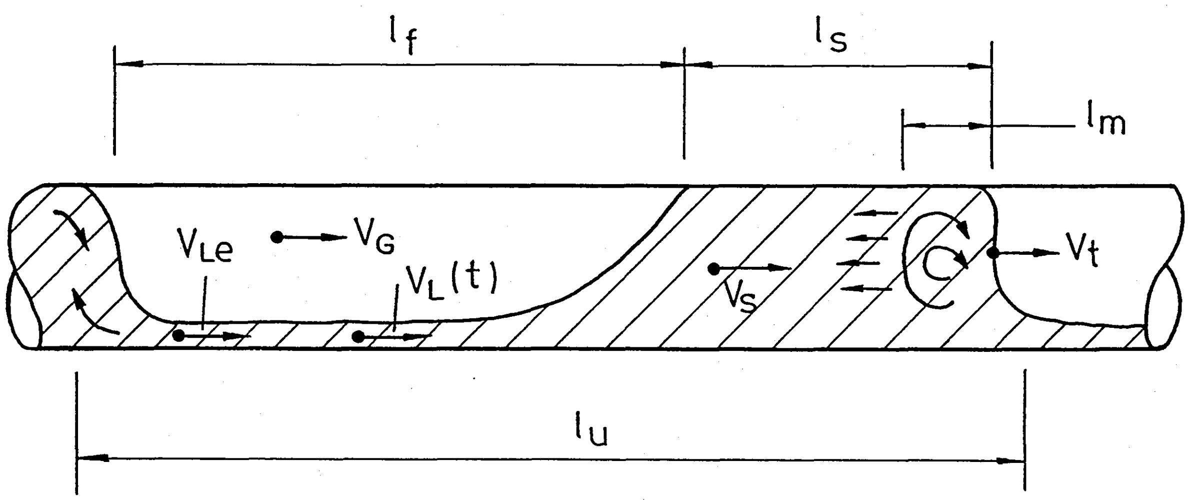


Fig 8.87 Mechanistic Slug Model of Dukler-Hubbard (1975)

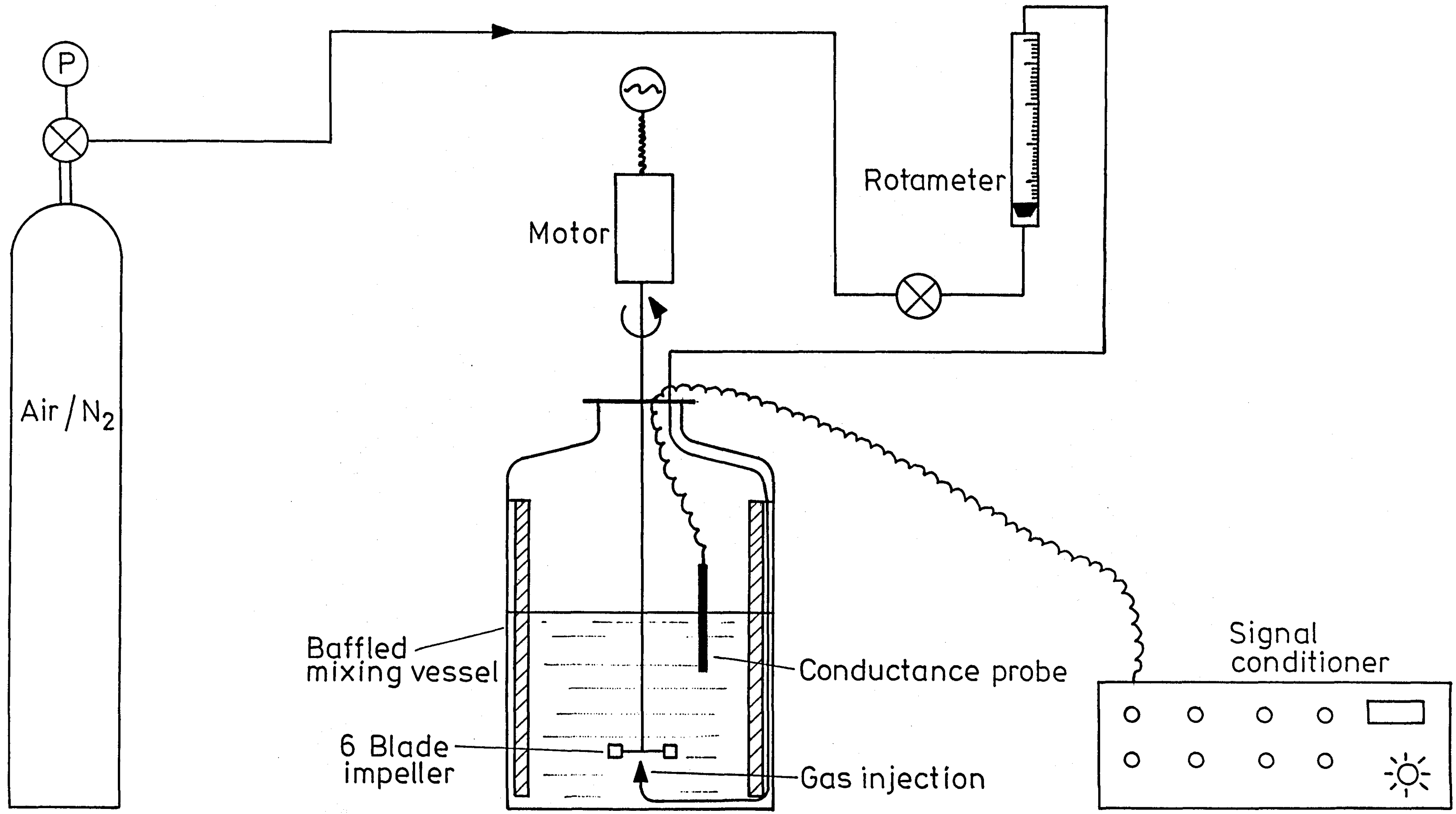


Fig A1.1 Small-scale Foaming/Dispersion Test Facility

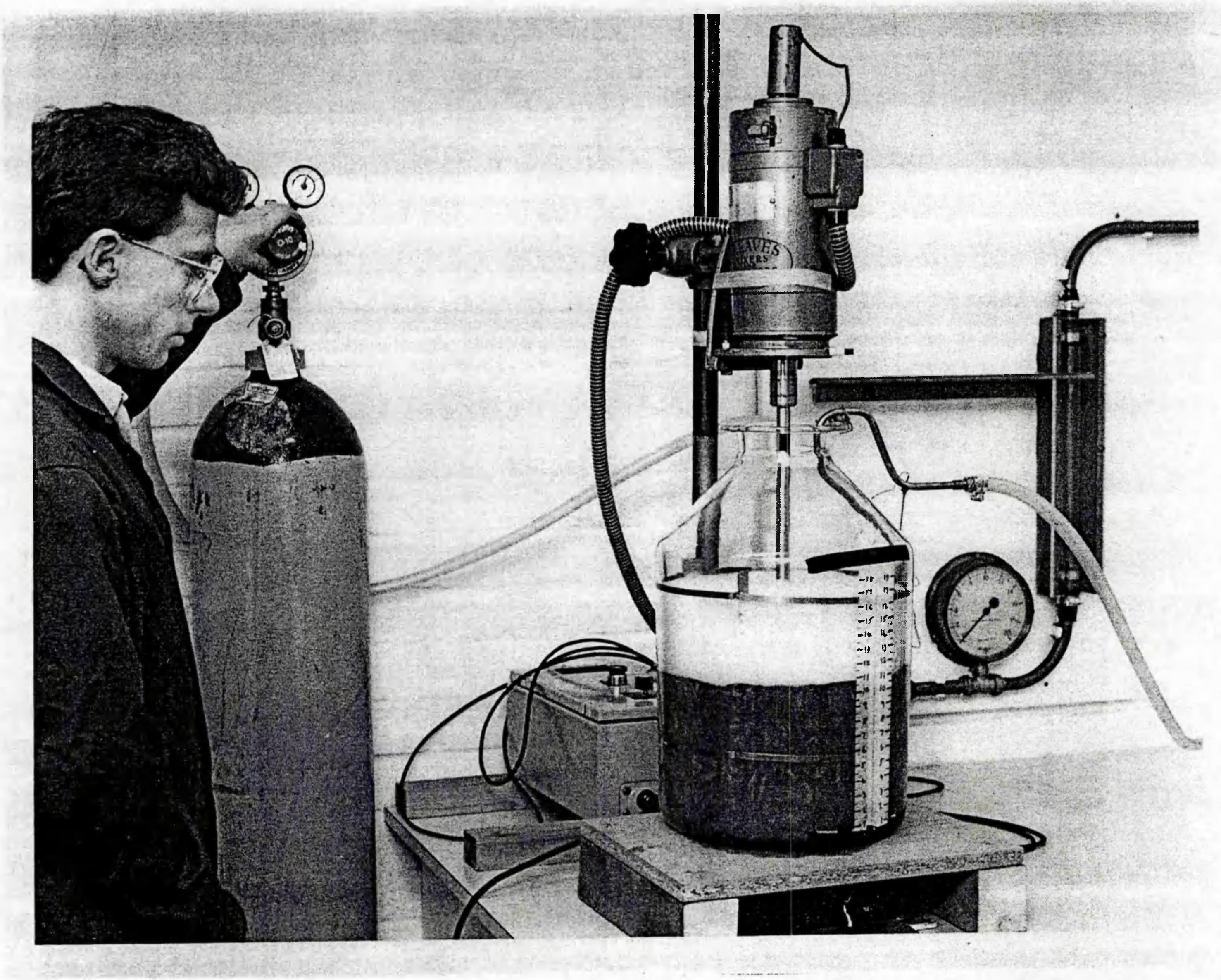


Fig A1.2 Small-scale Foaming/Dispersion Test Facility

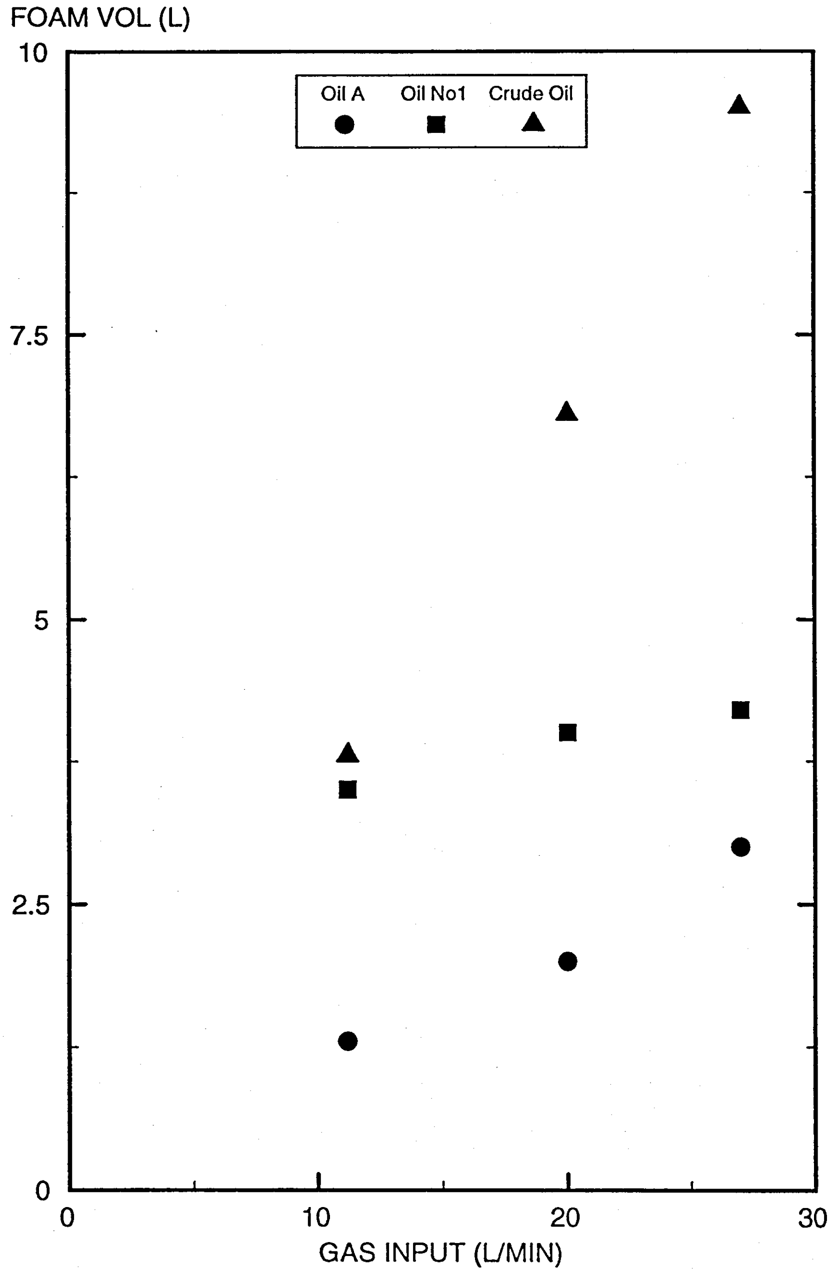


Fig A1.3 Measured Foam Data of Agitated Oil/Gas Systems