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The Flow Properties of Foams.

by

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ABSTRACT.

Work has been carried out to investigate the generation, bubble size characteristic, drainage properties and flow behaviour of high density foams in pipes.

Two distinct types of foams were investigated, the degree of dispersion of the gas phase being varied.

A very fine bubbled foam, with a mean bubble diameter of 60 microns, was generated by intense phase interaction in a centrifugal pump. The foam could be considered a homogeneous fluid, phase separation taking many minutes to occur. The flow properties of this foam were those of a Bingham fluid. The yield stress was found to be a function of the air/water volumetric ratio, the apparent viscosity being about 6 centipoise over the air/water range of 0.3 to 0.7. The flow showed the same behaviour in the pipe sizes tested, and no scaling factor depending on the bubble size was observed.

A foam with a much larger bubble size, average diameter of 300 microns, was generated from a sinter disc. This foam could not be considered homogeneous, separation into a dry foam and water layer occurring in several seconds. The foam exhibited a wide range of flow regimes, depending on the pipe size, detergent concentration and phase velocities. The foam again had a Bingham fluid type behaviour, but the flow properties were due to the presence of the dry foam and water layers in the flow system, and the effects of pipe size were important.

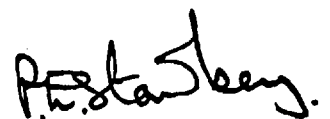
Both foams yielded friction factor plots which had the typical form associated with Bingham fluids. Use of such plots should be possible for design of foam systems.

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NOTATION.

a	:	speed of sound.
a	:	radius of tube.
A	:	area
d	:	diameter of pipe.
D	:	diameter of bubble
f	:	friction factor
g	:	acceleration due to gravity.
h	:	height of manometer.
H	:	holdup in pipe.
He	:	Hedstrom number. $\frac{\tau_0 \rho d^2}{\mu a}$
K	:	constant.
L	:	length
M	:	mass flux.
n	:	constant
P	:	pressure.
ΔP	:	pressure difference between two points.
Q	:	volumetric flowrate.
r	:	radius of bubble.
R	:	gas constant.
R_f	:	fraction of liquid in foam layer.
Re	:	Reynolds number.
s	:	standard deviation.
t	:	time
t_{50}	:	time for 50% of the liquid in a foam to drain.
T	:	temperature.
U	:	velocity.
V	:	volume.
x	:	mass fraction.
X	:	Lockhart-Martinelli factor, $(\Delta P_G / \Delta P_L)^{1/2}$
ϕ_{Lt}	:	Lockhart-Martinelli factor, $(\Delta P_{2phase} / \Delta P_L)^{1/2}$
α	:	voidage, $V_G / (V_G + V_L)$
β	:	effective slip coefficient.
μ	:	viscosity.
ρ	:	density.
τ	:	shear stress.
τ_0	:	intial shear stress, yield stress.

σ : surface tension.
 γ : volumetric ratio of gas to liquid.

Subscripts.

a : apparent.
b : bubble
c : correction
f : foam
G : gas
L : liquid.
m : mixture.
h : hydrostatic
o : initial
s : slip
s : solid
w : wall
W : water
 ∞ : infinity.

INTRODUCTION

The topic of this thesis is a relatively unexplored field, with a scarcity of pertinent papers, especially for high density foam. Most of the reported work on foams has dealt with the following aspects of the field:-

- 1) foam for use in mass transfer and stripping processes.
- 2) fire fighting foams, of high gas content and complex surfactant composition.
- 3) the properties of single films and bubbles in the foam, usually for dry foams with a polyhedral structure.
- 4) foams produced in process equipment, sieves, vats and liquid reaction vessels.

The system investigated in this case owes it's properties more to a system of small spherical bubbles, well spaced, and it is the bubble size, number and liquid properties which are controlling, rather than any surface forces due to the detergent action. The detergent was present mainly to aid in the creation and maintainance of the system in quasi-equilibrium.

Foam can be defined as a dispersion of gas within a liquid, in such a way that the liquid forms the continuous phase. The relative volumetric proportions of the gas and liquid phases can vary over a wide range, but the gas is always present as discrete bubbles, seperated by liquid layers. The system is usually meta-stable, the gas and liquid tending to separate with time. Surface active agents were added to impart foaming properties to the liquid, but some fluids with a high viscosity can form gas dispersions of sufficient stability to be considered capable

of forming a foam. The limits over which the gas/liquid ratio can vary are large, and many foams investigated are so far apart on this scale that comparison of results are inconclusive. Bikerman (1) has defined a foam as having an essentially polyhedral structure, polyederschaum, in which the liquid layers between gas bubbles are of thickness much less than the bubble diameter, with $\frac{\text{Volume Gas}}{\text{Volume Liquid}} > 10$. Cases when $\frac{V_G}{V_L} < 0.1$ he has defined as gas emulsions, the water layers being of comparable size to bubble diameter, and the bubbles spherical, Kugelschaum. He defines an intermediate region between these extremes in which the distinctions are difficult to make, mixtures exhibiting properties of the two extremes.

Foams with a greater amount of liquid in the bulk, $\frac{V_G}{V_L} < 10$, depend more on their liquid properties for their stability, albeit only transiently. Draining of liquid from between the bubbles, due to hydrostatic and surface forces, is controlled by the liquid viscosity and density while the film thickness is large, only when the film has thinned do surface forces predominate. In gas emulsions where the bubble spacing is large the stability of the system is more a function of the bubble hydrodynamics. The bubbles tend to rise up in the liquid as bubble distances are too large to create interactive forces.

The foams investigated in this work were in the range of 0.3 to 5.0, and so span the intermediate region. The liquid used, water, was of low viscosity, and a surface active agent was employed to reduce surface tension and impart a foaming ability. Foams were investigated whilst within the hydrodynamic drainage state, water layers between bubbles being of comparable magnitude to bubble diameter.

This type of system has not attracted much investigation, as most foam flows have employed the drier foams (3). However work by Grove et al (4) which give apparent viscosity measurements, and Starkey (5), have given pointers in the field. Much of the studies on gas and liquid systems have involved non-foaming systems. The dispersed flow regime can be considered a foam, even though extremely unstable and transient, and Huey et al (6, 7) have investigated some of the flow properties of such regimes.

A review of the pertinent aspects of foam and two phase flow has been carried out, and the theoretical development of the relevant equations presented. The foams have been characterised by the use of their bubble size and drainage properties, factors which can be used to classify foams on a general basis.

The flow patterns which can exist in a foam flow, for the various system conditions, are discussed in chapter 6. The presence of the detergent was found to create many differences in flow regime, when compared to pure water/air systems. The pressure drop for pipe flow of foam are given in chapter 7, the Bingham fluid properties of the foam being brought out in the yield stress and viscosity results. Friction factor plots typical of such fluids were found to results from the data.

The correlation of results by the Lockhart-Martinelli correlation, a widely used two phase flow description method, is presented in chapter 8, and indicates the use of such plots for foam correlation purposes.

The flow mechanisms for foam flow in pipes is discussed in chapter 9, and possible avenues for further study indicated.

1. Literature Review.

The foams involved in this study have a range of fluid properties similar to those of normal one phase fluids, and the methods applicable for the foam investigation are typical of any rheological study on a fluid. Foams have many properties, however, which demand a novel approach to the problem of their accurate measurement. The current literature gives a confusing picture of foam flow properties, in part due to the differing foam systems which have been investigated. The aim in this review has been to limit discussion to systems which have a large degree of similarity to the foams used.

Early foam studies have been nearly all involved with the dry foam properties, by such workers as Plateau investigating single films and gas bubbles and the forces which effect their stability. Groups of bubbles in a polyhedral foam have also been studied and the structural features determined (2). The stability of a liquid film in such a foam is a subtle combination of factors. The films are formed and maintained due to surface forces, which depend in a complex manner on the nature of the surface active agents employed and the geometry of the system.

1.1) Rheology of Foams

Foams or foaming systems have been widely used in many processes or applications and have been recognised as rheologically complex for a long time.

Sibree (8) reported work on the bulk viscosity of froths, showing that the apparent viscosity was greater than that of the bulk liquid, and varied with the shear rate. Blackman (9) discussed experiments with a concentric-cylinder viscometer,

which indicated the presence of a yield stress and apparent viscosity, on the model of a Bingham fluid.

$$\tau - \tau_0 = \mu \left(-\frac{du}{dx} \right) \quad 1.1)$$

Grove et al (4) measured the apparent viscosity of fire fighting foams by the use of friction factor/Reynolds number analogy. The viscometer consisted of an annular device, the force on the centre being measured via a previously calibrated spring. The apparent foam viscosity was found by analogy with the friction factor/Reynolds number plot obtained for one phase fluids. Results indicate that at high shear rates the apparent viscosity was constant. The results of viscosity at various foam densities are shown in Figure 1.1.

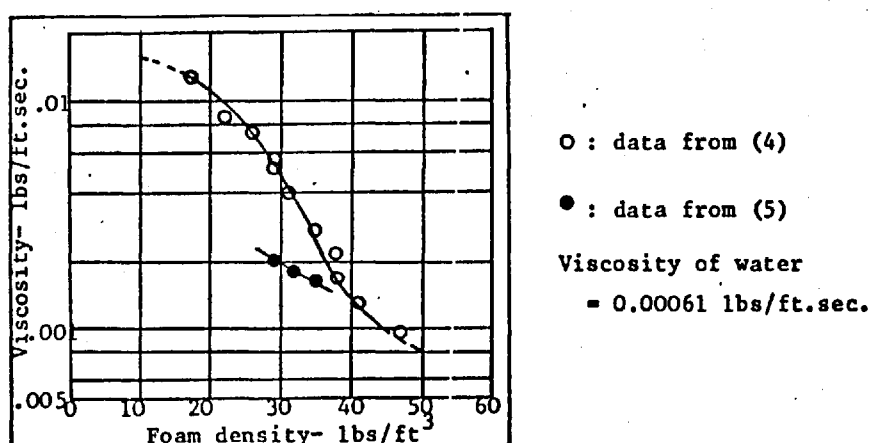


Figure 1.1: Effect of Foam density on Foam viscosity, (4) and (5).

This foam was generated by a gear pump and could be assumed to have a very small bubble size. Starkey (5) measured foam pressure drops in a foam generated from CO₂ saturated detergent solution and fitted results to equation 1.1. Values of $\tau_0 = 3.8 \text{ Nm}^{-2}$ and $\mu_a = 3$ centipoise for foams of were obtained. The viscosity values are indicated on Figure 1.1.

Wenzel et al (10) investigated the viscosity of high expansion foams with millimetre sized bubbles in both a cone and plate, and a varied concentric cylinder viscometer. The

cone and plate viscometer results gave data which could be fitted to an equation of the form

$$\tau = \tau_0 + k \left(\frac{du}{dx} \right)^{N'} \quad 1.2)$$

where the values of the constants were

$$\begin{aligned} .15 \times 10^{-3} &< \tau_0 < .85 \times 10^{-3} \text{ psi} \\ .12 \times 10^{-3} &< K < .50 \times 10^{-3} \text{ lb. ft}^{-1} \cdot \text{s}^{-1} \\ .13 &< N' < .70 \end{aligned}$$

No dependence of viscosity on bubble size was noted, 0.8 mm $< \bar{D} < 8$ mm. The yield stress was found to correlate with bubble diameter in the foam,

$$\tau_0 = 0.00435 (\bar{D})^{-1} \quad 1.3)$$

Wall slip was also noted in the experiments.

Many of the differing values and relationships which are given for τ_0 and μ_a , in the literature on foam can be attributed to the specific viscometer configurations and shear rates used in the determination of the values. The effects of the foam microstructure, bubble size and bubble separation, on the foam properties are still not clear.

The rheology of a gas dispersion, one of the extremes of a foam, can be considered analogous to that of dispersion of solids in a liquid. Einstein's theoretical investigation of such a system, with constraints on the degree of particle interactions, gives a relationship of the foam.

$$\frac{\mu_s}{\mu_L} = 1 + 2.5 \phi \quad \phi = \frac{\text{Volume solid}}{\text{Total volume}} \quad 1.4)$$

This has been found to apply only at low particle concentrations, less than 2% by volume. Modifications to account for particle size and surface forces (11) and the effects of particle interactions at higher concentrations, have led to equations of the type (12).

$$\frac{\mu_s}{\mu_L} = (1 + a_0 \phi + a_1 \phi^2 + a_2 \phi^3 + \dots) \quad 1.5)$$

These suffer from the need to determine the constants. Thomas (13) has correlated data from many sources, correcting for shear and size effects and proposes the curve shown in Figure 1.2.

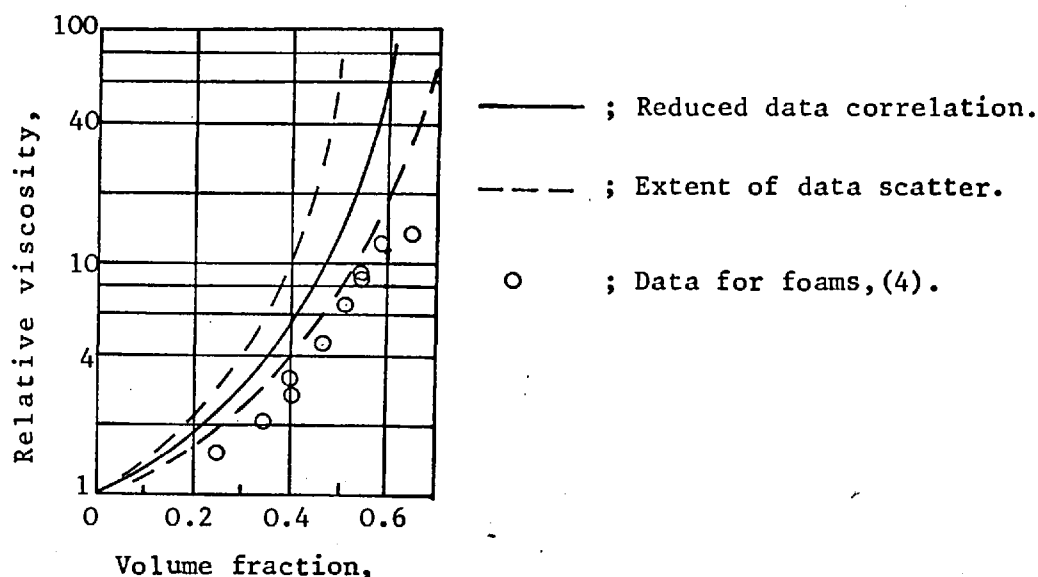


Figure 1.2: Relative viscosity of emulsions, (13).

Points included on this curve are the results of Grove et al (4). The experimental scatter about the curve is indicated by the dashed lines.

Particle shape factors have also been considered by Sibree (8, 15) who suggests an equation of the form,

$$\frac{\mu_s}{\mu_L} = \left[\frac{1}{1 + h' \phi^{\frac{2}{3}}} \right] \quad 1.6)$$

h' : shape factor

Through the previous investigations have studied foam rheological properties, and correlated results, much of the analysis was of an empirical nature, and limited to the system in question. Foam studies have indicated the complex behaviour of a foam, being a non-Newtonian fluid with a yield stress, which experiences wall slip. The aspect of the inherently time dependent properties of foams have usually been avoided by using samples at the same stage of

their history, either by homogenising so time is effectively zero (4), or by leaving for a period after a specified handling procedure (16).

The rheology of non-Newtonian fluids has been developed from studies based on Newtonian liquids, adapted to account for the non-ideality of the systems. For a Newtonian fluid there is a direct linear relationship between the shear stress and shear rate at every point in the fluid.

$$\tau = \mu \frac{dv}{dy} \quad 1.7)$$

For non-Newtonian fluids this simple relationship does not hold, modifications being necessary to account for three more types of fluid (17).

- a) time independent fluids, with or without a yield stress.
- b) time dependent fluids.
- c) viscoelastic fluids.

The internal structure of non-Newtonian fluids causes these rheological effects, and their analysis and study is by no means a trivial matter, depending as it does on the statistical properties of the micro-structure.

Additional effects due to preferential particle orientations at the fluid boundaries can give rise to an 'effective slip'. The bulk fluid properties do not extend to a solid surface, for a foam this would consist of a water layer and a region of lowered bubble density. Schofield et al (18) developed the theory to account and quantify this type of behaviour, and Oldroyd (19, 20) and Toms (21, 22) have detailed procedures to measure and correct for it's effect. Slip increases the flow rate which would exist if bulk properties persisted right up to the wall, for use in

precise rheological studies the shear stress and shear rate of the bulk fluid are required, and corrections must be made for the slip. At any pressure drop

$$Q_{\text{measured}} = Q_{\text{Fluid bulk properties}} + Q_{\text{wall slip effect}} \quad 1.8)$$

by correction for Q_s the time curve for $Q_F = f(\text{shear stress})$ can be obtained.

1.2) Two Phase Flow

The types of foams which this work investigates can at certain stages be compared to that of a gas-pure liquid system. The literature on general two-phase flow is extensive, as early as 1962 Gouse (23) lists over 2×10^3 articles on the general subject. Much additional work has been carried out since, and reviews by Kastun et al (24) and Hewitt et al (25) give more recent literature lists. Many books collating and unifying much of the available information have been published, texts by Wallis (26), Soo (27)

Govier et al (28) and Hewitt (29) have covered both the general field and the many particular aspects of two-phase flow systems. The field of investigation of gas/liquid flows is made especially difficult by the wide extremes of parameter choices which can occur, the multiplicity of flow regimes and the intractable analytical problems which can occur in the development of the theory. There is at the moment little sign that any single approach will be capable of covering all the eventualities of flow situations.

Brenner (30) and Batchelor (31) have developed part of the necessary theory to account for two phase systems, their results show the problems created by a consideration of discrete particles in the system, and the great degree of complexity introduced precluding any general analytical

solution. Drew et al (32, 33) have approached the matter similarly, and have used the results to obtain some theoretical basis for foam drainage and pressure/volume relationships.

Foams can be considered as homogeneous dispersion, and the critical flow properties of such medium show quite novel results. The speed of sound in any isotropic medium is a function of the pressure and density.

$$a^2 = \left(\frac{dP}{d\rho} \right) \quad 1.9)$$

a foam has both a high compressibility factor and density, when compared with either of the phases, and this results in an extremely low speed of sound.

The general theory has been developed for the case of homogeneous uni-directional flow, with a perfect gas by Tangren et al (35).

The relationship in terms of phase volumetric ratios is:-

$$a^2 = \frac{P}{\rho_L \alpha (1 - \alpha)} \quad 1.10)$$

This relationship has a minimum at $\alpha = .5$, for a mixture at 1 bar, $a \approx 20$ m/s which can be compared with 330 m/s for air. The speed of sound in a mixture has been found to depend on the frequency, at high frequency diffraction and bubble resonance starts to play an increasing part in the sound transmission (36). Foams can only be considered homogeneous when their characteristic dimensions, bubble size or wall film thickness, are small in comparison to pressure pulse wave length. Equation 1.10 only truly applies as the sound frequency tends to zero.

The theory to account for phase slip, and non-ideal gas dynamics has been investigated by Baum et al (37) Moody (38) and Fauske (39), where the effects due to the high

accelerational forces are considered. The results are necessarily more complex but indicate the low critical or sonic velocity. A review by Hsu (40) covers much of the available work.

Experimental investigation of choked two phase flow in a pipe flow situation has been carried out by Huey et al (6, 7). Experiments on flow in a nozzle have been presented by Eddington (41) Baum (42) and Smith (43) which show the general applicability of the one dimensional model. Herch et al (44) have investigated the action of diffusers after the high speed flow region, to obtain pressure recovery. The pressure recovery for pure liquids being poor once the flow regime passes out of the bubble stage, due to breakdown of the homogeneous aspect of the flow. Flow in nozzles was found to be essentially isothermal, the large heat capacity of the liquid swamping any gas thermodynamic effects if good thermal contact between phases exists.

The scale of the analytical problems has led to the use of many empirical procedures to correlate two phase flow system results, the effects of the phases being 'lumped' together in some manner. Lockhart and Martinelli (45) developed a correlation for the pressure drop in pipes, using the factors,

$$X = \sqrt{\frac{\Delta P_L}{\Delta P_G}} \quad \phi = \sqrt{\frac{\Delta P_{Two\ phase}}{\Delta P_{Sh}}} \quad 1.11)$$

The correlation is simple to use and gives reasonable results as a first approximation. Extensions of the basic method by Chisholm (46) and Baker (47) have improved the correlation accuracy by addition system investigation, greater range of pipe sizes, and correlation of results for

specific flow regimes.

The use of flow regimes maps to aid in the design of two-phase systems in both vertical and horizontal pipes is now well established (28). The flow pattern likely to exist is estimated from the phase properties and flow rates using a suitable map, and then the model for that regime can be employed. The accuracy of results depend to a certain degree on the similarity between the required system and the systems used in obtaining the map.

The parameters which have been used to correlate results have gone through a full cycle. One of the original flow maps by Bergelin et al (48) used the simple correlation of phase mass flux, or velocity, with flow regime and gave reasonable results if compared with simple systems in small pipes, air/water or oil/gas in $\frac{1}{4}$ " to 3" pipes. Later work by Baker (49) and Scott (5)) on a wider range of phases and pipe sizes found it necessary to include many of the fluid properties in the correlating parameters, resulting in such empirical groups as

$$\lambda = \left[\frac{\rho_w}{\rho_a} \cdot \frac{\rho_g}{\rho_l} \right]^n$$

and

$$\psi = \frac{k}{\sigma} \left[\frac{\mu_l}{\mu_w} \cdot \frac{\rho_w}{\rho_l} \right]^{n'}$$

w : water
a : air
L : liquid
G : gas

1.12)

The resultant flow maps were more difficult to use without giving any significant improvement in reliability or generality, or increasing the awareness of the factors involved in flow pattern determination.

Continuing work, based on the collection of data from many authors and systems, has amassed a large amount of information, much available for computer manipulation. Analysis of this data once again indicates that the simple

superficial velocities are the best correlating factors. Some minor changes to regime boundaries can be applied to account for the differing phase properties (51).

Experimental work on the particular flow regimes to explain flow behaviour and transition criteria is extensive (52). These show that much fine detail is of necessity lost in an overall flow map, and precise investigation of any particular flow regime is required to obtain accurate results. Recent work by Azis et al (53) on the elongated bubble regime show break-up of the bubbles under certain conditions, a factor of importance when surface active agents are present.

All flow pattern maps generally available have been prepared with non-foaming liquids, care being taken to exclude any contaminants which might affect the flow regimes. A paper based on the work of this thesis, dealing with flow regimes with foaming solution has been presented (54) and shows the marked change in flow patterns with phase mixing, pipe size and detergent concentration. A copy of this paper is given in the Appendix. The flow regimes in a polyhedral foam have been discussed by Hoffer et al (55) but no flow pattern map was constructed.

Friction factor plots for the correlation of two phase data have been presented, in the same manner as for single phase fluids. Flow regimes have been limited to the homogeneous, ie., dispersed bubble or mist, where phase-velocities can be assumed equal, $U_G = U_L$. The friction factor, $\tau / \frac{1}{2} \rho u^2$, has been plotted against Reynold Number, $\frac{D u \rho}{\mu}$, the value of viscosity used being that of the liquid, justification for this being based on the liquid film at the wall.

Results vary, Wallis (26) indicating that a value of $f = 0.005$ was a good estimate. Work by Yokota (56) and recently Davis (57) has shown a dependence of f on Re . Both these workers investigated vertical as well as horizontal flows at high velocities, in the homogeneous dispersed bubble region. The results obtained were correlated in the form

$$\frac{1}{\sqrt{f}} = A + B \ln(Re \cdot \sqrt{f}) \quad 1.13)$$

Flow direction	A	B
Vertical (57)	-10.6	2.7
Horizontal	- 3.4	1.9
Both vert. & hor (56)	- 1.14	2.16
Single phase(58)	- 0.4	1.7

Foams lend themselves to this approach, work by Grove (4) making use of such an analogy to one phase flow in a foam viscometer. Results by Starkey (5) indicated that foam friction factors correlations equivalent to those for Bingham fluids could be obtained from pipe flow of foams, on the basis of the analysis due to Hedstron (59).

1.3) Foam Systems

The foam systems which have been reported in the literature cover a wide range of the possible factors. Different pipe sizes, foaming agents, bubble generation techniques and data presentation methods have all tended to make comparisons between the various systems difficult, or even meaningless, due to the diversity of factors, some of which are unspecified or not controlled in the experimental work.

There are many methods of generating foams, each with varying degrees of sophistication and reproducibility. The early work on foams was interested in dry foams, similar to those on the top of liquids in process equipment, and the

generation methods were crude, shaking a quantity in a vessel, or using a beater (1). Work on bubble production for froth and foam mass transfer process has used gas injection via capillaries (60) or sinter discs (61). Workers investigating the flow properties of foams have employed a range of methods,

a) Sinter discs or spinnerets.

b) Blowing air and foaming solution through a series of wire gauges, a usual technique for producing large volumes of high expansion foam for fire fighting (62).

c) Passing air and water through a packed bed (63).

d) Using high speed water to entrain air, on the laboratory water pump principle, and then creating a foam by impact (64).

e) Passing the air and water through a pump which by it's mechanical design breaks up the air (4).

f) Reduction of pressure over a foam solution saturated with a soluble gas (5), or mixed with a low boiling point solvent (65).

g) Use of a venturi to entrain air, the system being employed for oxygenation purposes (66).

h) The use of a pump and air leakage into the liquid stream to create a foam (67).

This wide range of possible generation methods gives nearly as wide a range of different types of foams, gas to liquid ratios vary greatly, bubble sizes are different and foam drainage and lifetime properties differ. Even with any particular generating mode minor design or flow rate changes can alter the foam produced. The parameters of the foam to use in it's characterisation are many, and as yet there is no standard series of tests recognised as being able to

measure "foamability" of a liquid or "quality" of a foam.

The testing methods for foams have been based on a rather adhoc and empirical studies. Early work on foams were aimed at the understanding of drainage of films and and bubble strengths, Bikerman (1) reviews these aspects satisfactorily. Most foam testing methods have been based on a measure of the time for a sample of foam to:-

- a) Drain, time against liquid drainage rate
- b) Collapse, time against foam height or volume.
- c) Coalesce; time against some function of the bubble size or distribution.

The effects of bubble size, temperature, foaming agent, liquid and gas phases and measurement vessel have all been investigated, but not in any fully systematic manner. It is this lack of standardisation which makes comparison of results difficult, coupled with the general problem of reproducibility of particular foam systems.

Work at the Fire Research Establishment has aimed at standardisation of generators, foam quality, bubble size and measuring techniques. Foam parameters are defined by the time for 25% of the liquid in the foam to drain (16), and the shear force on a particular viscometer in a specified aged foam (69).

The bubble size and distribution in the foam is another important factor which has attracted much attention, without as yet leading to any unified theory on the effects of bubble size on foam properties. The usual technique of measurement is via a photographic method (10, 55, 70, 71), the foam being photographed in the pipe, or in a static condition.

De Vries (72) discussed the inherent draw-backs of

comparing foam bubble size at a wall to that of the bulk, but Chang (70) with frozen foam samples showed that bubble size and density was constant through the bulk of the sample. Typical bubble sizes range from 20 microns. to over 10 mm for the range of generators detailed above. De Vries (72) gives data for foams made with Teepol, and sodium alginate as a viscosity increasing agent, which have bubble sizes in the range 20 to 80 microns.

Bubble coalescence and size changes resulting from gas diffusion between bubbles show that the gas phase plays an important part in foam lifetimes (72, 71). The relationship between bubble size and liquid and gas phase properties has been shown to be

$$r_o^2 - r^2 = \left(\frac{4RT}{P_o} \right) \left(\frac{Ds\gamma}{\theta} \right) t \quad 1.14)$$

Experimental results uphold this relationship.

The foam can be considered a fluid in it's own right, and as such has properties which depend upon the value of pressure, temperature and foam composition. An equation of state for the gas phase, first postulated by Tait in 1867, has been presented by Ross (74) where

$$PV + 2\sigma A = nRT \quad 1.15)$$

An identical result was obtained by Drew et al (32) based on theoretical considerations. This equation relates the surface properties of the gas in the foam bubbles with the external environment properties, similar relationships could be developed for the other thermodynamic properties of the foam.

2. Theoretical Development

2.1) Rheology of Foams

The use of capillary tube viscometers for the measurement of rheological properties of non-Newtonian fluids is well established (75). The use of the pressure gradient to infer the relationship between shear stress and shear rate in this type of viscometer is by no means a trivial matter. The rate of shear is not constant over the tube radius, varying from a maximum at the wall to zero at the centre of the pipe. The generally non-analytical nature of the stress/strain relation for such fluids makes the interpretation of results both difficult and places many restrictions on the applicability of the results (17).

It can be shown that a relationship between $Q/\pi a^3$ and τ_w exists which is independent of the tube radius, however complex the relation between shear stress and shear rate-

$$\frac{Q}{\pi a^3} = \frac{1}{\tau_w^3} \int_0^{\tau_w} \tau^2 f(\tau) d\tau \quad 2.1)$$

The assumptions made in the development of the equation are:-

- 1) the fluid properties are time independent.
- 2) rectilinear laminar flow with axial symmetry exists in the tube.
- 3) the rate of shear at any point in the fluid depends only on the shear stress at that point,

$$\dot{\gamma} = \frac{du}{dy} = f(\tau_w) \quad 2.2)$$

- 4) there is no slip at the wall.

$$u_w = 0$$

Since the integral in equation 2.1 is a function only of the limits, $Q/\pi a^3$ depends only on τ_w , no matter how

complex the function. The shear stress at the pipe wall is given by

$$\tau_w = \frac{\alpha \Delta P}{2.L} \quad 2.3)$$

Hence a plot of τ_w versus $Q/\pi\alpha^3$ for any fluid will give a single curve if the assumptions above are valid. This graph can be used to derive the relationship between the shear stress and rate of shear as follows;

$$\frac{Q}{\pi\alpha^3} = F(\tau_w) = \frac{1}{\tau_w^3} \int_0^{\tau_w} \tau^2 \cdot f(\tau) \cdot d\tau \quad 2.1)$$

hence

$$f(\tau) = \frac{1}{\tau_w^2} \cdot \frac{d[\tau_w^3 \cdot F(\tau_w)]}{d\tau_w} \quad 2.4)$$

However the curve $F(\tau_w)$ is not normally sufficiently defined to allow this differentiation to be carried out in practice.

The technique often used to overcome this problem is to define some arbitrary function between $\dot{\gamma}$ and τ_w in equation 2.1, and solve the integration. Then by inspection and comparison of the data with this idealised model, with a certain amount of manipulation of the constants, ^{it is possible} to arrive at a relationship which gives agreement, within the experimental limits, of the model and experimental results.

A large number of these functional relationships have been developed, and the general theory for each case investigated. The relations most used are:-

1) For a Newtonian Fluid

$$\tau = \mu \dot{\gamma} = \mu \frac{du}{dy} \quad 2.5)$$

which on substitution gives

$$Q = \frac{\pi \cdot a^4 \cdot \Delta P}{8 \cdot \mu \cdot L} \quad 2.6)$$

the familiar Poiseuille equation for laminar Newtonian flow in a tube.

2) For a fluid with a yield stress, a Bingham fluid,

$$\dot{\gamma} = \frac{\tau - \tau_0}{\mu_a} = f(\tau_w) \quad ; \quad \tau > \tau_0 \quad 2.7)$$

which gives on substitution,

$$Q = \frac{\pi a^4 \Delta P}{8 \mu_a L} \left[1 - \frac{4}{3} \left(\frac{2L \tau_0}{a \Delta P} \right) + \frac{1}{3} \left(\frac{2L \tau_0}{a \Delta P} \right)^4 \right] \quad 2.8)$$

known as the Buckingham equation. This reduces to equation 2.6 if $\tau_0 = 0$.

3) For power law fluids,

$$\tau = k \dot{\gamma}^n \quad \text{or} \quad \dot{\gamma} = \left(\frac{\tau}{k} \right)^{\frac{1}{n}} \quad 2.9)$$

which on substitution gives

$$Q = \frac{n \pi a^3}{3n+1} \cdot \left(\frac{a \Delta P}{2Lk} \right)^{\frac{1}{n}} \quad 2.10)$$

This again reduces to equation 2.6 if $n = 1$ and $k = \mu$.

The value of the exponent in equation 2.9 is often used as a measure to define the 'degree of non-Newtonian behaviour', the closer to $n = 1$ the fitted results, the less non-Newtonian the fluid is considered to be.

4) a combination of all these functions to cover all the many fluid properties that can exist. Models used have been

$$\tau = \frac{A \dot{\gamma}}{B + \dot{\gamma}} + \mu_{\infty} \dot{\gamma} \quad 2.11)$$

and

$$\tau = A \dot{\gamma} + B \cdot \sinh^{-1}(C \dot{\gamma}) \quad 2.12)$$

A, B, C : constants.

The aim of the use of these functions has been to approximate as closely as possible the flow curve, but still to have a form capable of analytic manipulation. The use of several constants allow such a wide latitude in fitting that care must be employed if meaningful results are to be obtained, especially in cases which involve extrapolation of design.

2.1.1) Problems with Foams

The general theory so far described must be applicable to the system under investigation if it is to be used for describing the flow; or any deviation from it be accountable for, and the necessary corrections and adaptations made.

Foam is a rheologically complex material, with problems unique in it's generation, property characterisation and determination, and control of the flow system. Foam consists of two very different phases, with property values at the two extremes of their value scales, a highly variable and indefinable microstructure, and properties as yet imperfectly understood. Many of the most basic assumptions that can be made for normal one phase fluids cannot be assumed to hold, this either results in derivations which are mathematically intractable, or approximations of large orders being made in the derivation, with all the attendant inaccuracies which can arise.

The factors which must be considered in any detailed theoretical treatment of a foam are;

- 1) Normal one phase fluids in rheological terms are considered as a continuum, a factor involved in the theory

of laminar flow and the no-slip boundary condition at the most basic level. Foams, or in fact any multiphase system in which the discrete particle size is large, cannot be considered continuous on the microscale. Hence any foam system must have dimensions which are sufficiently large to create an apparent homogeneity, a size ratio between bubble and measuring device of at least 1:100 being desirable (75). Hence for the foams under consideration the minimum pipe diameter which could be used if this restriction were to apply, is of order $100 \times D$. This for the Sinter generated foams with a $D = 350$ microns, is a pipe of diameter at least 35 mm, and for the pump generated foams, $D = 60$ microns, a diameter of at least 6 mm. The use of such large pipe sizes for rheological studies bring it's own problems.

2) Foams in the long term cannot be considered a time-independent fluid, as it's properties at any instant depend to a certain extent on it's previous "history", and the forces that have acted upon it. The foams used in the experimental work have been chosen, generated and handled in such a way that they would have as constant a set of properties, over the period of any experimental determinations carried out on them, as possible. The time in which the foam takes to pass through the pipe in which the pressure drop is measured was arranged so the foam would have constant properties over that period. Experimental work on the foam drainage showed that foams may be considered to have constant properties for times ranging up to the ten's of seconds. The smaller the bubble size, the longer the foam can be assumed to have

constant properties.

3) The analysis depends on the flow being in the laminar region, there being no velocity fluctuations in the fluid on the microscale, and no radial components of velocity. The foam flow was often difficult to arrange to be laminar, due in part to the constraints mentioned in 1 and 2. The accepted definition and test for laminar flow is a fluid Reynold number below some critical value, typically 2000 for Newtonian fluids. Now

$$Re_f = \frac{\rho_f U_f D}{\mu_f} \quad 2.13)$$

The precise definition and measurement of these terms for a foam is by no means absolutely clear, there being a certain degree of ambiguity in the literature.

Constraint 1) results in a large, by capillary viscometer standards, value of D , which will give a large Reynolds number.

Constraint 2) means that time for measurement of the foam properties must be short, the time of transit for the foam between pressure measuring points must be small, and the velocity must therefore be large, increasing the value of Re .

The foam properties then tend to result in flows which have higher Reynolds number values than would be used in similar investigation of a single phase fluid.

4) The assumption of axial symmetry for a foam can also be a poor approximation under certain conditions. Foams which have drained to any extent will have a stratified structure and a marked density gradient in

the foam layer. Foams generated with the Sinter cannot be considered to have a truly even composition.

5) The rate of shear might not be only a function of the stress.

$$\dot{\gamma} = f(\tau, D, \epsilon, \dots) \quad 2.14)$$

For a foam flowing in a pipe there could be a particular bubble orientation which is favoured, such as the bubbles forming into lines, at different shear rates. Also large bubbles may have an effect on the surrounding fluid to a distance comparable to their size, variations in bubble density could cause anisotropic regions to occur in an otherwise uniform foam.

The magnitude of these effects on the foams in question is difficult to quantify, and as such, many of the possible errors have been assumed negligible. The foam produced by the pump generator was expected to conform to the assumptions made in the theoretical analysis.

2.1.2) Slip at the Wall

The assumption of no slip of fluid at the pipe wall is essential if plots of $Q/\pi\alpha^3$ against τ_w , for the different pipe sizes, are all to lie on the same line. If several curves result from the data of the different pipe sizes, then the most likely reason is that near the wall there is an anisotropic region, with different fluid properties to that of the bulk.

The presence of such a layer near to the wall is typical for multiphase media, and is widely reported for foam (10) and polymer solutions (22). The presence of the wall alters the orientation and packing characteristics of the particulate phase, resulting in a region next to the wall which is depleted in the second phase. Particles cannot have their centres nearer to the wall than the average particle radius.

The effect of this layer is to create a region of different rheological properties, and so τ_w , measured in the experiment, is not the value which would exist if the bulk properties extended right up to the wall. This situation has been investigated thoroughly (18), and methods of determining the effects of slip have been developed (19). The analysis is based on the assumption of an effective slip velocity at the wall. True slip is not believed to occur. Very dry foams with the bubbles having the wall as one side of the polyhedral bubble will undergo true slip. This analysis does not include such cases.

The effective slip is denoted by U_s at a radius $r = R$, where $U_s = f(\tau_w)$.

The volumetric flow rate through the differential annulus between r and $r + dr$ is;

$$dQ = u \cdot 2\pi r \cdot dr \quad 2.15)$$

where u is the velocity at r .

$$\therefore Q = \int_0^R dQ = \pi \int_0^R u \cdot 2r \cdot dr \quad 2.16)$$

substituting $d(r^2) = 2r.dr$. so

$$Q = \pi \int_0^{R^2} u.d(r^2) \quad 2.17)$$

Integrating by parts gives

$$Q = \pi \left[ur^2 - \int r^2 du \right]_0^{R^2} \quad 2.18)$$

or

$$Q = \pi u_s R^2 - \pi \left[\int r^2 du \right]_0^{R^2} \quad 2.19)$$

where u_s is the velocity at the wall. In cases when no slip can be assumed, $u_s = 0$, and the first term can be cancelled, and the analysis leads to equation 2.1.

The terms in equation 2.19 can be identified with the flows caused by the fluidity and the slip,

$$Q_{\text{fluid}} = \pi \left[\int r^2 du \right] \quad 2.20$$

$$Q_{\text{slip}} = \pi u_s R^2 \quad 2.21$$

hence

$$Q_{\text{total}} = Q_{\text{slip}} + Q_{\text{fluid}} \quad 2.22$$

If the effect of slip can be accounted for the relation between ΔP (or τ_w), and Q_{fluid} can be determined, and the rheological properties of the bulk foam separated from the wall effects. From assumption 3 and equation 2.2

$$\dot{\gamma}_{rx} = \frac{du}{dr} = -f(\tau_{rx}) \quad 2.2)$$

and the shear stress at any r is τ_{rx}

$$\tau_{rx} = \frac{r \Delta P}{2L} \quad 2.23)$$

at the wall

$$\tau_{rx})_{r=R} = \tau_w = \frac{R \Delta P}{2L} \quad 2.3)$$

$$\therefore \tau_{rx} = \frac{\tau_w r}{R} \quad 2.24)$$

$$\therefore r = \frac{\tau_{rx}}{\tau_w} R \quad 2.25)$$

$$\therefore dr = \frac{R}{\tau_w} d(\tau_{rx}) \quad 2.26)$$

substitution in equation 2.19 with equation 2.25 for r^2 ;
and equations 2.2 and 2.26 for dr gives

$$Q_T = \pi u_s R^2 + \pi \left[\int \frac{\tau_{rx}^2}{\tau_w^2} \cdot R^2 \cdot f(\tau_{rx}) \cdot \frac{R}{\tau_w} \cdot d\tau_{rx} \right]_0^{R^2} \quad 2.27)$$

which on re-arrangement

$$Q_T = \pi u_s R^2 + \frac{\pi R^3}{\tau_w^3} \int_0^{\tau_w} \tau_{rx}^2 \cdot f(\tau_{rx}) \cdot d\tau_{rx} \quad 2.28)$$

dividing throughout by πR^3

$$\frac{Q_T}{\pi R^3} = \frac{u_s}{R} + \frac{1}{\tau_w^3} \cdot \int_0^{\tau_w} \tau^2 \cdot f(\tau) \cdot d\tau \quad 2.29)$$

which is equivalent to equation 2.1 with the additional
term ; u_s/R

dividing by τ_w

$$\frac{Q}{\pi R^3 \tau_w} = \frac{u_s}{R \tau_w} + \frac{1}{\tau_w^4} \cdot \int_0^{\tau_w} \tau^2 \cdot f(\tau) \cdot d\tau \quad 2.30)$$

The effective slip coefficient, β , can be defined as

$$\beta = \frac{u_s}{\tau_w} \quad 2.31)$$

so β is equivalent to the effective slip velocity per unit
shear stress at the wall.

$$\frac{Q}{\pi R^3 \tau_w} = \frac{\beta}{R} + \frac{1}{\tau_w^4} \int \tau^2 \cdot f(\tau) \cdot d\tau \quad 2.32)$$

The value of β can be evaluated as a function of τ_w from experimental work on a number of pipes, with the use of equation 2.32.

1) Plot, for a range of R , but constant L , $Q/\pi R^3 \tau_w$ against τ_w . If $\beta = 0$ (no slip) all curves will coincide. If the curves are different, $\beta \neq 0$.

2) For the different curves, at a fixed value of τ_w , read off the values of $Q/\pi R^3 \tau_w$ for each R .

3) Plot $Q/\pi R^3 \tau_w$ against $1/R$. Equation 2.32 indicates that the slope of this line gives β at values of τ_w . For a fixed value of τ_w , the second term on R.H.S. is constant.

$$\therefore \frac{Q}{\pi R^3 \tau_w} = \frac{\beta}{R} + \text{Constant} \quad 2.33)$$

$$\text{i.e.} \quad y = \beta \frac{1}{R} + \text{Constant.}$$

4) Repeat steps 2 and 3 at different τ_w .

5) Plot the value of β against τ_w .

From the relationship of $\beta = f(\tau_w)$ the measured Q_T can be corrected for the effect of slip, from equation 2.22,

$$Q_{\text{fluid}} = Q_{\text{total}} - Q_{\text{slip}} = Q_{\text{total}} - \pi \beta \tau_w R^2 \quad 2.22)$$

where τ_w is given by equation 2.3.

Hence from the data of at least two pipes the relationship between shear stress and shear rate for the bulk foam can be found, separating out the effects of the wall slip.

2.2) Friction Factor Correlation

The use of the friction factor as a correlating method is common for single phase flows. Two-phase flows normally encountered do not exhibit sufficient homogeneity for the method to be applicable, only in the dispersed bubble (56, 57) and annular mist (26) regimes has this been attempted. Reynolds numbers for these flows are of necessity large, 10^5 , as high turbulence is required to impart homogeneous flow. The foams investigated have a pseudo-homogeneous form even at very low flow rates, $Re \approx 10^3$, and so would be expected to yield information over a wider range of Re .

The friction factor, f , is a dimensionless group relating shear stress to energy of the flow,

$$f = \frac{\tau_w}{\frac{1}{2}\rho u^2} \quad 2.34)$$

The value of Reynold's number is a function of the fluid and flow parameters.

$$Re = \frac{u \cdot \rho \cdot d}{\mu_a} \quad 2.35)$$

The foam flow has been shown to be typical of a Bingham fluid, and the f v's Re plot could be expected to follow the pattern exhibited by such fluids (59), as shown in Figure 2.1.

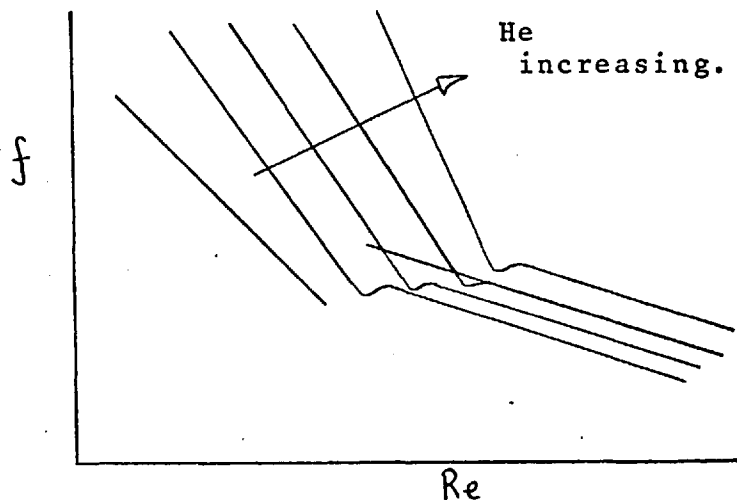


Figure 2.1: Typical f v's Re plot for a Bingham fluid.

Analysis of the equation for a Bingham fluid

$$\tau - \tau_0 = -\mu_a \frac{du}{dx} \quad (2.7)$$

shows that the friction factor is a function of two groups

$$f = f(Re, He)$$

where

$$He = \frac{\tau_0 \cdot \rho \cdot d^2}{\mu_a^2} \quad (2.36)$$

Turbulence is estimated to occur when the curve at any He value cuts the curve for turbulent single phase flow, hence $(Re)_{critical}$ is a function of He .

2.2.1) Foam Reynolds Number

The definition of the Reynolds number for two-phase flows is open to some variation of the precise values of parameters to use.

$$Re = f(d, u, \rho, \mu)$$

The value of u, ρ, μ used has depended on the system in question, and the use to which the results are to be put. The contribution of the phases, and their weighting, to these properties, has given rise to several definitions of homogeneous two-phase Reynolds numbers

$$1) \quad Re_M = \frac{d \cdot V_m \cdot \rho_m}{\mu_m} \quad (4, 5) \quad (2.37)$$

$$2) \quad Re_M = \frac{d \cdot V_m \cdot \rho_m}{\mu_L} \quad (56, 57) \quad (2.38)$$

$$3) \quad Re_M = \frac{d \cdot V_m \cdot \rho_L}{\mu_L} \quad (28) \quad (2.39)$$

where L:liquid property and M:mixture property.

The flows have been assumed homogeneous, with no phase slip, $U_G = U_L$.

V_M = Total volumetric rate/flow area.

$$\rho_M = \frac{Q_G}{Q_L} \cdot \rho_L$$

μ_M : either some weighted mean of the gas and liquid values is used.
$$\frac{1}{\mu_M} = \frac{\alpha}{\mu_G} + \frac{(1-\alpha)}{\mu_L} \quad 2.40)$$
 or some equation of the Einstein type used, or the value measured by some means.

The use of μ_L has been justified by the argument that the layer next to wall transmits the shear force, this being a liquid layer, the viscosity is best given by the liquid value.

The various definitions for Reynolds number are related to the liquid Reynolds number, for the no-slip condition as,

$$\rho_M V_M = \rho_L V_L$$

which gives on substitution in the expressions,

$$1) \quad Re_M = Re_L \frac{\mu_L}{\mu_f} \quad 2.37a)$$

$$2) \quad Re_M = Re_L \quad 2.38a)$$

$$3) \quad Re_M = Re_L \left(1 + \frac{Q_G}{Q_L} \right) \quad 2.39a)$$

The value used in this work was $Re_f = \frac{d \cdot \rho_f \cdot V_f}{\mu_a}$, the value of μ_a being obtained by application of the Bingham model to results, and was the apparent, or plastic, viscosity of the foam.

2.3) Accelerational Pressure Drop

The pressure drop in the pipe due to the effect of the fluid accelerating as it passes down the pipe can in many cases be a substantial part of the total pressure drop, especially in high speed compressible flows. The neglect of this factor in the estimation of rheological properties can lead to serious errors in the gas-liquid system, and was initially the cause of much confusion among early workers who found their pressure drops' results in disagreement.

In this work however the effect of accelerational pressure drop has been ignored in the derivation of the flow curves for the various foams investigated. Pressure drops are comparatively low, ($\Delta P_{\max} < 0.1 P_1$), unlike cases which investigate the annular flow regimes, and the errors involved in neglecting this correction are of the same magnitude as the other errors involved.

The maximum pressure drop due to the accelerational pressure drop encountered in this work is about 5% of the measured pressure drop.

The momentum equation for steady, one dimensional homogeneous flow can be written in the form,

$$M \frac{dv}{dz} = A \frac{dP}{dz} - B \tau_w - A \rho_m g \cos \theta. \quad 2.41)$$

This can be re-arranged to the form B : duct perimeter.

$$\frac{dP}{dz} = \left(\frac{dP}{dz} \right)_{\text{Fric}} + \left(\frac{dP}{dz} \right)_{\text{acc}} + \left(\frac{dP}{dz} \right)_{\text{grav}} \quad 2.42)$$

where the terms can be identified as

$$\frac{B}{A} \tau_w = - \left(\frac{dP}{dz} \right)_{\text{Fric}} : \text{frictional pressure gradient}$$

$$\frac{M}{A} \frac{dv}{dz} = - \left(\frac{dP}{dz} \right)_{\text{acc}} : \text{accelerational pressure gradient}$$

$$\rho_m g \cos \theta = - \left(\frac{dP}{dz} \right)_{\text{grav}} : \text{gravitational pressure gradient}$$

For the system in question the term $\left(\frac{dP}{dz}\right)_{\text{grav}} = 0$.

The frictional pressure gradient can be expressed in terms of the friction factor.

$$\tau_w = C_f \cdot \frac{1}{2} \rho_m \bar{u}_m^2 \quad 2.43)$$

$$\therefore \left(\frac{dP}{dz}\right)_{\text{Fric}} = 2C_f \frac{\rho_m \bar{V}_m^2}{d} \quad 2.44)$$

The acceleration pressure gradient can be expressed in terms of the kinetic energy change of the flow, where

$$\left(\frac{dP}{dz}\right)_{\text{acc}} = \rho_m V_m \cdot \frac{dV_m}{dz} \quad 2.45)$$

Consider the case of flow of mixture, $Q_G = Q_L$, at a flow rate sufficient to give a pressure drop of 100 cm of water/metre, $\simeq 0.1$ bar/metre, at average pressure of 1 bar.

$$V_{m1} = \frac{Q_{m1}}{A} = \frac{Q_L + Q_{G1}}{A} \quad 2.46)$$

$$V_{m2} = \frac{Q_{m2}}{A} = \frac{Q_L + Q_{G2}}{A} \quad 2.47)$$

but

$$Q_{G2} = \frac{P_1 Q_{G1}}{P_2} = \frac{1.1 \times Q_{G1}}{1.0} \quad 2.48)$$

Using $Q_G = Q_L$

$$V_{m1} = 2.0 V_L \quad \text{and} \quad V_{m2} = 2.1 V_L$$

$$V_m = 2.0 V_L$$

$$\left(\frac{dP}{dz}\right)_{\text{acc}} = 0.05 \rho_m \bar{V}_m^2 \quad \text{per metre.}$$

now for a 1 metre pipe, with $d \simeq 2$ cms, and using a value of C_f of 0.005. (26).

$$\begin{aligned} \left(\frac{dP}{dz}\right)_{\text{Fric}} &= 2C_f \cdot \rho_m \bar{V}_m^2 \cdot \frac{L}{d} \\ &= 0.5 \rho_m \bar{V}_m^2 \end{aligned}$$

at highest pressure drop encountered,

$$\left(\frac{dP}{dz}\right)_{\text{acc}} = 10\% \cdot \left(\frac{dP}{dz}\right)_{\text{Fric}}$$

Hence at the maximum value for pressure drop, of interest

in the foam the accelerational pressure drop was $\simeq 10\%$ of the frictional pressure drop, and was not accounted for in the estimation of pressure drop. A similar neglect of the accelerational pressure drop has been used by other works (28, 45, 56), investigating stratified, slugging and bubbly flows. Neglect of the accelerational terms in high velocity or annular flow can cause major errors in results, the acceleration pressure drops are high, and expansion of the gas contributes a large part of the total pressure drop.

2.4) Lockhart and Martinelli Correlation

The Lockhart and Martinelli correlation for horizontal two phase flow was developed to give estimates of pressure drop and hold-up ratios (45). The correlation was essentially developed with an empirical basis, but it has since been shown to have certain theoretical justifications. It is a widely used method for obtaining a first approximation to the pressure drop, what it tends to lack in accuracy it makes up for in it's ease of application (28). The results obtained are good if conditions in the system are similar to those used in it's derivation, and much additional work (46, 47), using similar analysis techniques, has been carried out on flow regimes and pipe conditions not included in the original work. The use of these alternative correlations based on the Lockhart and Martinelli model has greatly extended the accuracy and application of the basic method.

The basis of the correlation is that the pressure drop due to the liquid phase flowing in the pipe must equal the pressure drop due to the gas phase, whatever the flow pattern. This is equivalent to considering the gas and liquid to be flowing in two parallel pipes, with diameters D_G and D_L , such that

$$\left(\frac{\Delta P}{L}\right)_G = \left(\frac{\Delta P}{L}\right)_L = \left(\frac{\Delta P}{L}\right)_{\text{Two phase}} \quad 2.50)$$

rather than together in a single pipe, diameter D .

Hence the pressure drop for each phase can be expressed conventially using the friction factor concept with the relevent phase properties, as

$$\left(\frac{\Delta P}{L}\right)_L = 2 f_L \left(\frac{\rho_L V_L^2}{D_L} \right) \quad 2.51)$$

$$\left(\frac{\Delta P}{L}\right)_G = 2 f_G \left(\frac{\rho_G V_G^2}{D_G} \right) \quad 2.52)$$

where D_L and D_G is the equivalent hydraulic diameter of the region of the pipe in which the liquid and gas flow. The velocities, V_L and V_G , are those of the liquid and gas flowing in the regions D_L and D_G , where

$$V_L = \frac{M_L}{\alpha \left(\frac{\pi}{4} D_L^2 \right) \rho_L} \quad + \quad V_G = \frac{M_G}{\beta \left(\frac{\pi}{4} D_G^2 \right) \rho_G} \quad 2.53)$$

where constants α and β are to allow for the flow geometry,

$$\alpha = \frac{\text{actual cross section of liquid flow area}}{\text{cross section of pipe diameter } D} \quad 2.54)$$

and similarly for β with the gas.

The friction factors f_L and f_G may be expressed in terms of a generalized Blasius form.

$$f_L = \frac{B_L}{Re_L^n} = \frac{B_L}{\left(\frac{4}{\pi} \right)^n \left(\alpha D_L \mu_L \right)^n} \quad 2.55)$$

Hence by substituting equation 2.55 and 2.53 into 2.51, making use of the equality of equation 2.50 it can be shown that

$$\left(\frac{\Delta P}{L} \right)_{2\text{phase}} = 2 B_L \cdot \frac{M_L^{2-n}}{\rho_L} \cdot \alpha^{n-2} \cdot D_L^{n-5} \cdot \mu_L^n \cdot \left(\frac{\pi}{4} \right)^{n-2} \quad 2.56)$$

The pressure drop which would exist if the water flowed alone in the pipe can also be expressed in a similar form;

$$\left(\frac{\Delta P}{L} \right)_{SL} = 2 B_L' \cdot \frac{M_L^{2-n'}}{\rho_L} \cdot D'^{n'-5} \cdot \mu_L^{n'} \cdot \left(\frac{\pi}{4} \right)^{n'-2} \quad 2.57)$$

where B_L' and n' are the relevant constants in the Blasius friction factor correlation.

If we assume $B_L = B_L'$ and $n = n'$ then

$$\left(\frac{\Delta P}{L} \right)_{2\text{phase}} = \left(\frac{\Delta P}{L} \right)_{SL} \alpha^{n-2} \left(\frac{D}{D_L} \right)^{5-n} \quad 2.58)$$

The factor of interest in the correlation is the value of $\left(\frac{\Delta P}{L} \right)_{2\text{phase}}$, and this is combined with the liquid

pressure drop, a known and calculable value,

$$\sqrt{\frac{\left(\frac{\Delta P}{L}\right)_{2\text{phase}}}{\left(\frac{\Delta P}{L}\right)_{SL}}} = \phi_L = \alpha^{\frac{n-2}{2}} \cdot \left(\frac{D}{D_L}\right)^{\frac{5-n}{2}} \quad 2.59)$$

and similarly for the gas phase

$$\sqrt{\frac{\left(\frac{\Delta P}{L}\right)_{2\text{phase}}}{\left(\frac{\Delta P}{L}\right)_{SG}}} = \phi_G = \beta^{\frac{n-2}{2}} \cdot \left(\frac{D}{D_G}\right)^{\frac{5-n}{2}} \quad 2.60)$$

Lockhart and Martinelli postulated that ϕ_L and ϕ_G were function of X , where

$$X = \sqrt{\frac{\left(\frac{\Delta P}{L}\right)_{SL}}{\left(\frac{\Delta P}{L}\right)_{SG}}} \quad 2.61)$$

The functional relationship between ϕ_L , ϕ_G and X was determined by experiment, the initial correlation being based on data in small, 1" to $\frac{1}{10}$ " pipes, with air and a variety of liquids, with specific gravities $\approx .8$ to 1.00, and at near atmospheric pressures. The data was correlated into four groups, depending on the superficial Reynolds number of the phases. A critical Reynolds number of 1000 being taken to separate the laminar from the turbulent data.

For the purposes of comparison with this work the case of ϕ_{Ltt} , the liquid and gas turbulent cases, is used as all but a few data points give $Re \gg 1000$. The plot for the ϕ_{Ltt} case with co-ordinates is given in Section 8.1. The plot shows the typical data scatter found in such work, and is taken from the original L - M paper, and shows the ϕ_{Gtt} case. From this plot the line for ϕ_{Ltt} can be drawn, using the relationship

$$\phi_{Ltt}^2 = \frac{\phi_{Gtt}^2}{X} \quad 2.71)$$

2.5) Pressure of Gas in Foam.

If a volume of gas, V_{g0} , at a pressure P_0 , and a volume of water V_L are mixed together in such a way that an evenly distributed dispersion of small bubbles in liquid is produced there is a reduction in the total volume of the system. This volume reduction is due to the fact that the gas experiences an average pressure $\bar{P}$, where $\bar{P} > P_0$. This increase in the gas pressure comes about for several reasons:-

a) The gas in the bubbles experiences an excess pressure, ΔP_b , compared to the pressure just outside the bubble, due to the surface tension forces acting, where

$$\Delta P_b = \frac{2\sigma}{r} \quad 2.72)$$

The value of ΔP_b for the fine bubble foams used in the experiments with the pump generator is $\simeq 1 \times 10^{-2}$ bar.

b) The gas in a bubble in the bulk of the foam also experiences an additional pressure to that above the foam, due to the hydrostatic head of foam above it, where

$$\Delta P_h = h_f \rho_f g \quad 2.73)$$

For a foam in the small scale foam generating device, $h_f = 20$ cm. and $\rho_f = \frac{1}{2} \rho_L$; $\Delta P_h \simeq 1 \times 10^{-2}$ bar.

Hence the total pressure experience by the gas in a bubble in the foam is

$$P_f = P_{atm} + \Delta P_b + \Delta P_h \quad 2.74)$$

The pressure in a foam bubble is therefore a function of the bubble size, the foam density, and the position of the bubble in the foam. The average pressure in a foam $\bar{P}_f$, can be defined if the following assumptions are made,

1) The bubbles in the foam can be characterised by an average bubble radius, $\bar{r}$, which is the same value at all points in the foam, No preferential size sorting taking

place in the foam.

2) The foam density does not vary with depth in the foam, the bubbles are evenly distributed throughout the bulk of the liquid.

3) The foam sample is contained in a vessel of constant horizontal area, equal amounts of gas are present at different foam depths.

Hence the average pressure of the gas in the bubble of a foam is,

$$\bar{P}_f = P_{atm} + \frac{1}{2}\Delta P_h + \frac{2\sigma}{r} \quad 2.75)$$

In the foams in question it can be seen that the relative magnitude of these effects can be a few % of the atmospheric pressure, and hence should be measurable.

The equation for the pressure in the foam is identical with the Tait equation obtained by Ross (74) for the pressure of gas in a foam

$$PV + \frac{2\sigma A}{3} = nRT \quad 2.76)$$

which relates foam pressure, volume and surface area. This equation allows the calculation of the surface area of a foam if the pressure or volume relationship can be established.

For the case of the isothermal decay of a foam, T constant,

$$P_1 V_1 + \frac{2\sigma A_1}{3} = P_2 V_2 + \frac{2\sigma A_2}{3} \quad 2.77)$$

If case 1 is the desired foam of which the surface area is required, the case 2 is the same foam when the gas and liquid have been separated, the foam collapsed and $A_2 = 0$, then

$$P_1 V_1 + \frac{2\sigma A_1}{3} = P_2 V_2 \quad 2.78)$$

hence if this change in the foam surface area takes place at constant volume,

$$(P_2 - P_1) = \frac{2\sigma A_1}{3V_1} = \frac{2\sigma}{\bar{r}} \quad 2.79)$$

The value of the average bubble radius of the foam can be determined if the value of P_2 and P_1 can be measured, this for a very fine bubble foam should be $\cong 1\% P_{atm}$, well within the limits of accurate determination.

2.6) Speed of Sound in a Foam

It has been demonstrated that there exists a unique velocity in a pseudo-homogeneous medium at which a slowly varying pressure pulse travels (34). This is not however true of a rapidly fluctuating pressure or sound wave, which travels by a complex mechanism of attenuation, scattering and dispersion from the bubbles (36).

The velocity of a wave in an isotropic homogeneous medium is given by-

$$a^2 = \frac{dP}{d\rho} \quad 2.80)$$

A foam with small bubbles, and a comparatively low gas to liquid ratio, less than 10, can be considered to be sufficiently homogeneous for this equation to apply.

If the gas to liquid mass ratio is defined by α , and the volume ratio by δ , where

$$\alpha = \frac{\text{mass gas}}{\text{mass liquid}} \text{ per unit volume} = \frac{m_{\text{gas}}}{m_{\text{liquid}}} \quad 2.81)$$

$$\delta = \frac{\text{Volume gas}}{\text{Volume liquid}} \text{ per unit volume} = \frac{V_{\text{gas}}}{V_{\text{liquid}}} \quad 2.82)$$

For a two component foam with no mass transfer between the phases,

$$\alpha = \text{constant} \neq f(P)$$

but the value of $\delta = f(P)$

The average density of the mixture can be written as,

$$\rho_f = \frac{\text{mass of gas} + \text{mass of liquid}}{\text{volume}} = \frac{\rho_g V_g + \rho_l V_l}{V_g + V_l} \quad 2.83)$$

$$\therefore \rho_f = \rho_l \left[\frac{\alpha + 1}{\delta + 1} \right] \quad 2.84)$$

The liquid can be assumed to be incompressible, $\rho_l = \text{constant}$, and the effect of surface forces ignored. At low volume ratios the heat capacity of the liquid is very much

greater than that of the gas, and volume changes will hence occur at constant temperature to the first approximation. The isothermal equation of state can be used.

$$P\delta = \text{constant} \quad 2.85)$$

Differentiating equation 2.85 by parts w.r.t ρ gives

$$\frac{P d\delta}{d\rho} + \delta \frac{dP}{d\rho} = 0$$

From equation 2.84, $\therefore \frac{dP}{d\rho} = -\frac{P}{\delta} \frac{d\delta}{d\rho}$ 2.86)

$$\delta = \frac{P_L}{P} (\alpha + 1) - 1$$

differentiating w.r.t ρ , and substituting gives

$$\frac{d\delta}{d\rho} = -\frac{(1+\delta)^2}{P_L(1+\alpha)} \quad 2.87)$$

substituting in equation 2.80, with 2.86 and 2.87, gives

$$a = (1+\delta) \left[\frac{P}{P_L(1+\alpha)} \right]^{1/2} \quad 2.88)$$

For the foams in question, $\delta = 0[1]$, and so $\alpha \simeq 10^{-3}$

Rearranging equation 2.88 using α to replace δ , where

$$\alpha = \text{voidage} = \frac{\text{Vol of gas}}{\text{Volume of foam}}$$

and assuming $(1+\alpha) \simeq 1$

$$a = \left[\frac{P}{P_L \alpha (1-\alpha)} \right]^{1/2} \quad 2.89)$$

The speed of sound given by equation 2.89 at normal pressures is low, and has a minimum when the term $\alpha(1-\alpha)$ is a maximum, at $\alpha=0.5$. The speed of sound calculated using equation 2.89 is shown in Figure 2.2. Of note are the low values of a , and the flatness of the curves between $.3 < \alpha < .7$, the speed of sound only changing by 10% over this large range.

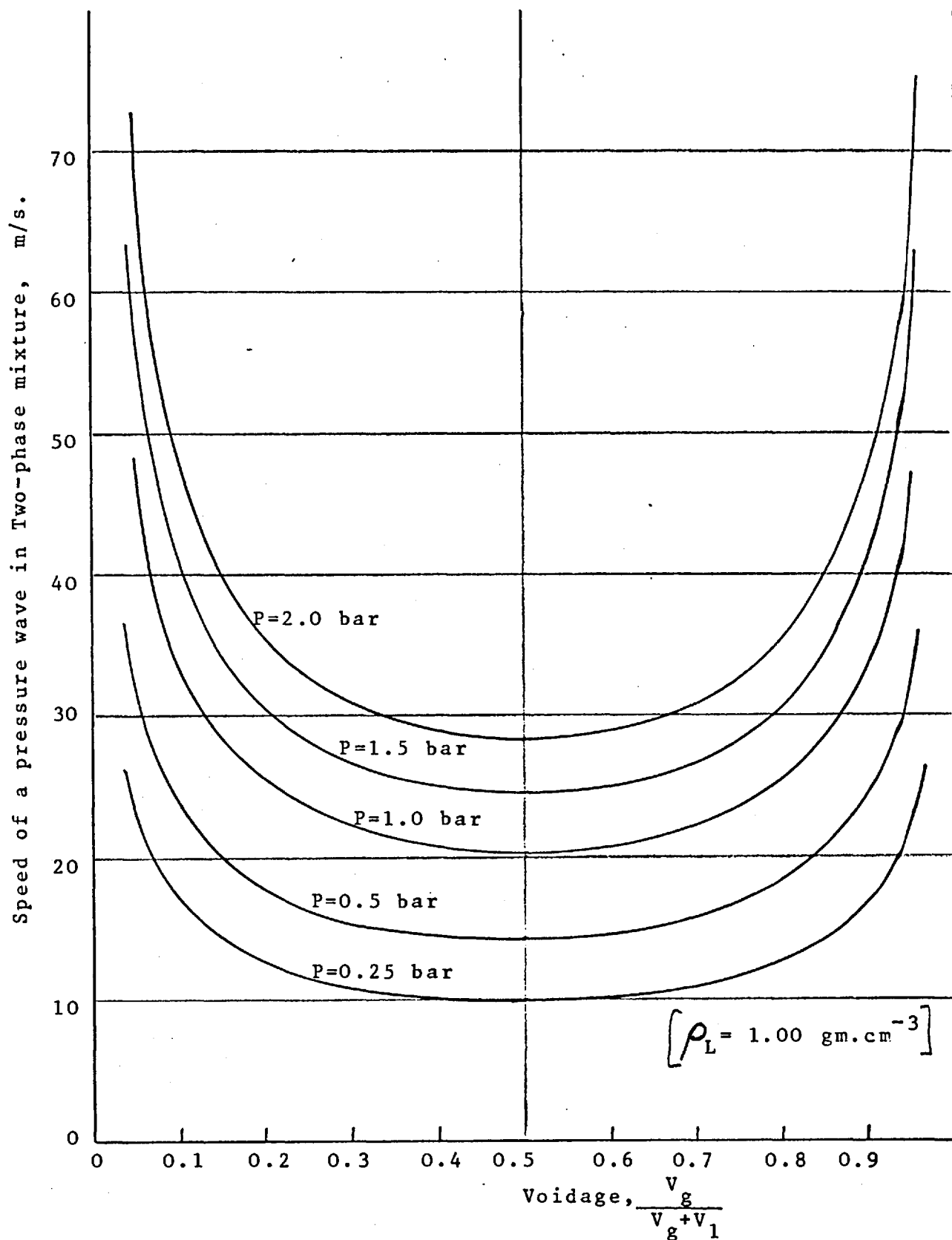


Figure 2.2: Variation in the speed of sound with Absolute pressure and voidage. Calculated by Equation 2.89.

3) Equipment and Experimental Details

3.1) Main Equipment

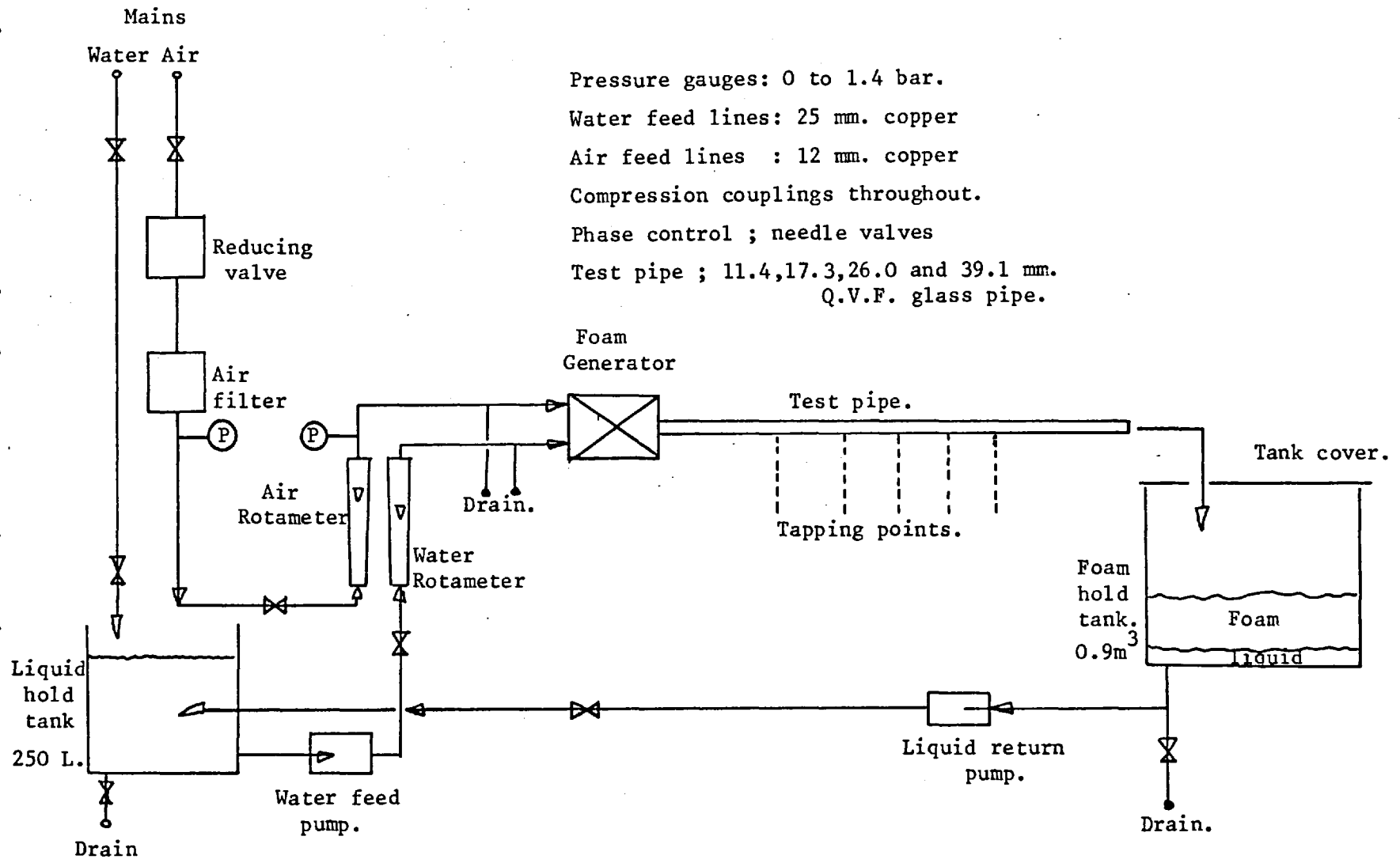
The main equipment was designed and built as shown schematically in Figure 3.1. The equipment can be considered to consist of 5 main sections, the air/water metering system, the foam and liquid holding tanks, the pipework used in the tests, the pressure drop measurement devices and the foam generation equipment.

3.1.1) Air/Water Supply System

The air used in all experiments was taken from the compressed air supply available in the laboratory, at a pressure of 90 psig. The air was passed through a reducing valve and filter before use. Flow rate measurement was carried out using a range of rotameters, to give flow rates from 1.0 to 60.0 litres/min. The air pressure was measured after the air filter to indicate supply pressure, and immediately after the rotameter, to allow the air flow rates to be calculated, and to give the pressure of air into the foam generator. The air flow rates were calculated by the use of prepared calibration charts for each rotameter, these were checked, at selected rotameter readings and outlet pressures, by comparison with a displacement gas meter, and were found to agree within the rated errors for these rotameters, $\pm 2\%$. Tables of flow rate versus rotameter reading over a range of pressures, from 1 to 20 psig intervals, were drawn up and used with interpretation.

The water used in the experiments was taken from the mains, and held in a 250 litre galvanised iron tank. All water used for experiments was held for at least $\frac{1}{2}$ hour in the tank, and pumped round the system at least once, to ensure knockout of

Figure 3.1: Schematic diagram of Main Equipment.



any dissolved air, mains supply being 40 psig. The amount of air which would come out of solution was appreciable, if mains water was used to flush out manometer tapplings the very small bubbles formed caused problems as they were very difficult to dislodge off the tube walls. The mains water quality was not checked chemically in any way, but periodic readings of surface tension were made to ensure pure supply, values of 71 ± 1 dynes/cm were always recorded. See Section 3.1.5 for details on system cleaning to remove detergent traces in rig.

A charge of 192 litres of water was used for each experimental run, this quantity being measured by a depth marker in the liquid hold tank. The desired detergent solution was made by adding Teepol to this charge, to give the required concentration.

a) 96 mls for a 0.05% solution

b) 960 mls for a 0.5% solution

These volumes being measured with a graduated cylinder. The detergent solution was changed regularly during the course of the experiments, 3 days being the longest time any solution was kept. The action of the close mixing of the air with the detergent solution causing a cloudiness to occur as the surfactant degrades, Teepol being a bio-degradable detergent.

The water temperature was measured by mercury thermometer, and was periodically monitored during experimental runs, as prolonged running would cause a temperature rise due to heat gain from the pumps. Temperature rises of $\approx 2^{\circ}$ C/hr were noted with constant running.

Flow rate measurements were carried out with a suitable range of rotameters, giving flow rates from 1 to 60 l/min. Charts were constructed and checked by sample collection,

errors being within $\pm 1.5\%$.

The metered air and water flows were then piped to the foam generation equipment.

3.1.2) Foam Holding Tank

After the foam has passed through the test pipe it had then to be destroyed, the air allowed to escape, and the water re-cycled. During the course of experimental runs any foam samples taken for external measurements were returned to the foam holding tank, so that the total water stock remained constant.

The air and water were separated by allowing the foam to drain and collapse in a large tank of volume 0.9 M^3 . The water which drained from the foam was drawn off at the bottom of the tank, the air escaping from the top surface of the collapsing foam. This method of separation was chosen after trials with a mechanical foam breaker, of the basket or disc type (82), had been found to be ineffective for breaking low voidage foams with small bubble sizes, and incapable of handling the flow rates used in the experiments. The tank was of a reinforced glass fibre construction, of the type used for water storage. The size of the tank, $.9\text{M}^3$, was sufficient to allow continuous return of the liquid at most water flow rates, problems only being encountered with the extremely fine bubbles produced by the pump generator, see Section 4 on foam drainage. These bubbles were removed in the liquid hold tank, and never re-entered the cycle. The liquid from the collapsed foam was returned via a pump and a 1" pipe.

The tank was covered with a wooden top, with suitable holes to allow foam entry. This was used to stop the

dispersal of the tiny detergent droplets which formed a mist above the breaking foam. As the foam bubbles broke they would release droplets into the air, these droplets containing a higher percentage of detergent due to the stripping action (60), and could cause health problems, if breathed in any quantity. The cover on the tank stopped them being dispersed into the laboratory, most hit the tank walls or top and drained back into the foam.

The water drainage from the foam was satisfactory, but as Teepol forms a very dry and long-lived foam on prolonged drainage, this very dry foam would slowly fill the tank if a high air ^{rate} was being used. This would happen after about 1 hour's running at values of air above 30 l/min, the foam flow had then to be stopped to allow this dry foam to break, or it would overspill the tank.

3.1.3) Test Pipe and Framework

The pipe, in which experimental runs were carried out, was supported at a height of about 5 feet, in a series of pipe rings mounted on a metal framework. Supports were placed to hold the pipe depending on the particular configuration being employed, but never more than 1 metre apart. The pipe was firmly held in each ring using packing, and the position of each ring then fixed by use of an adjustable clamp on the top of the framework. See Figure 3.2.

Two phase flow pattern can depend to a large extent on the pipe orientation, differing flow patterns being obtained in vertical and horizontal pipes (28). Initial work on the stratified foam regime showed that the effect of a non-horizontal pipe caused major changes to the interface, creating waves.

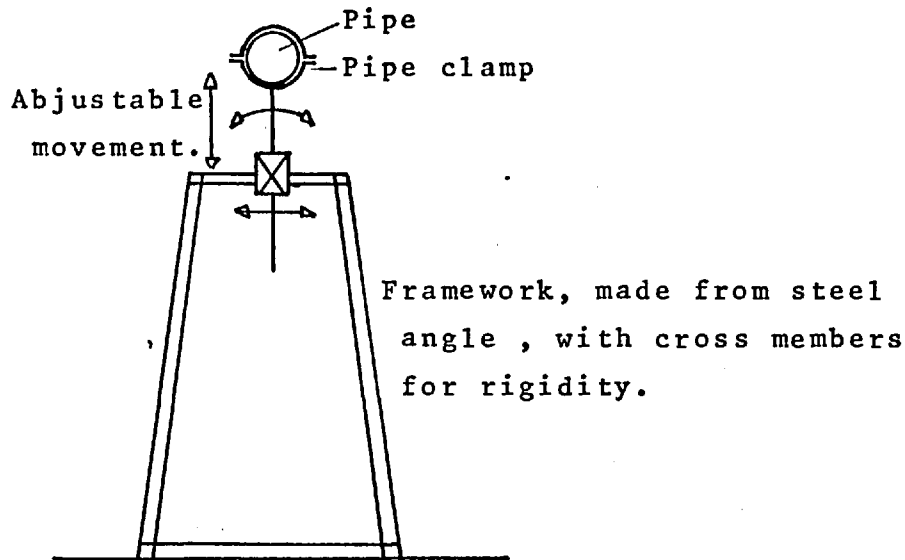


Figure 3.2: Pipe arrangement on frame-work.

The length of pipe was levelled horizontally using a spirit level, and a length of transparent tubing containing water, setting the two water levels to the bottom of the pipe at different positions. The pipe was kept to within ± 2 mm over the usual 6 metres of run, problems being encountered due to the slight bends inherent in the glass pipes, although some slight adjustment could be made at the joints. The end of the pipe was fed into the holding tank, via a 90° bend and a short vertical length of pipe to minimise splashing and aid foam drainage.

The sealing rings between pipe sections were made of P.T.F.E., and were designed to be the same internal diameter as the pipe, to minimise flow disturbances at the joins in the test pipe. A slight bellling of the pipe was present at the end, tapping points were always placed at least 20 cm. from any pipe end.

3.1.4) Pressure Measurement System

The pressure measurements taken on the pipes were carried out using a manometer system, using water as the working fluid. Provision for up to six tapping points to be read simultaneously was made. A schematic of a pair of lines is shown in Figure 3.3 the complete system is given in Figure B1 in the Appendix.

3.1.4.1) Operation of System

Before use all the manometer lines were flushed out with water to remove any air bubbles trapped in the lines. Water from the liquid hold tank, supplied via the main pump was used, so that any flow from the lines into the test pipe would not alter liquid compositions. Initial trials with mains water showed problems with dissolved air in the mains water, which formed tiny bubbles in the pipes.

With the manometer lines filled and the foam flowing in the pipe, the levels in manometers, A for point A etc., were allowed to reach their own levels, adjustment to the air pressure at the top of the manometers being carried out to keep the levels within the range of the manometers. All lines were again purged to remove any foam which might have entered the lines at the tapping points as the manometer levels changed. This air adjustment and line purge was repeated until the manometers gave constant readings after subsequent cycles.

The pressure difference between points A and B, and the absolute pressure could then be found.

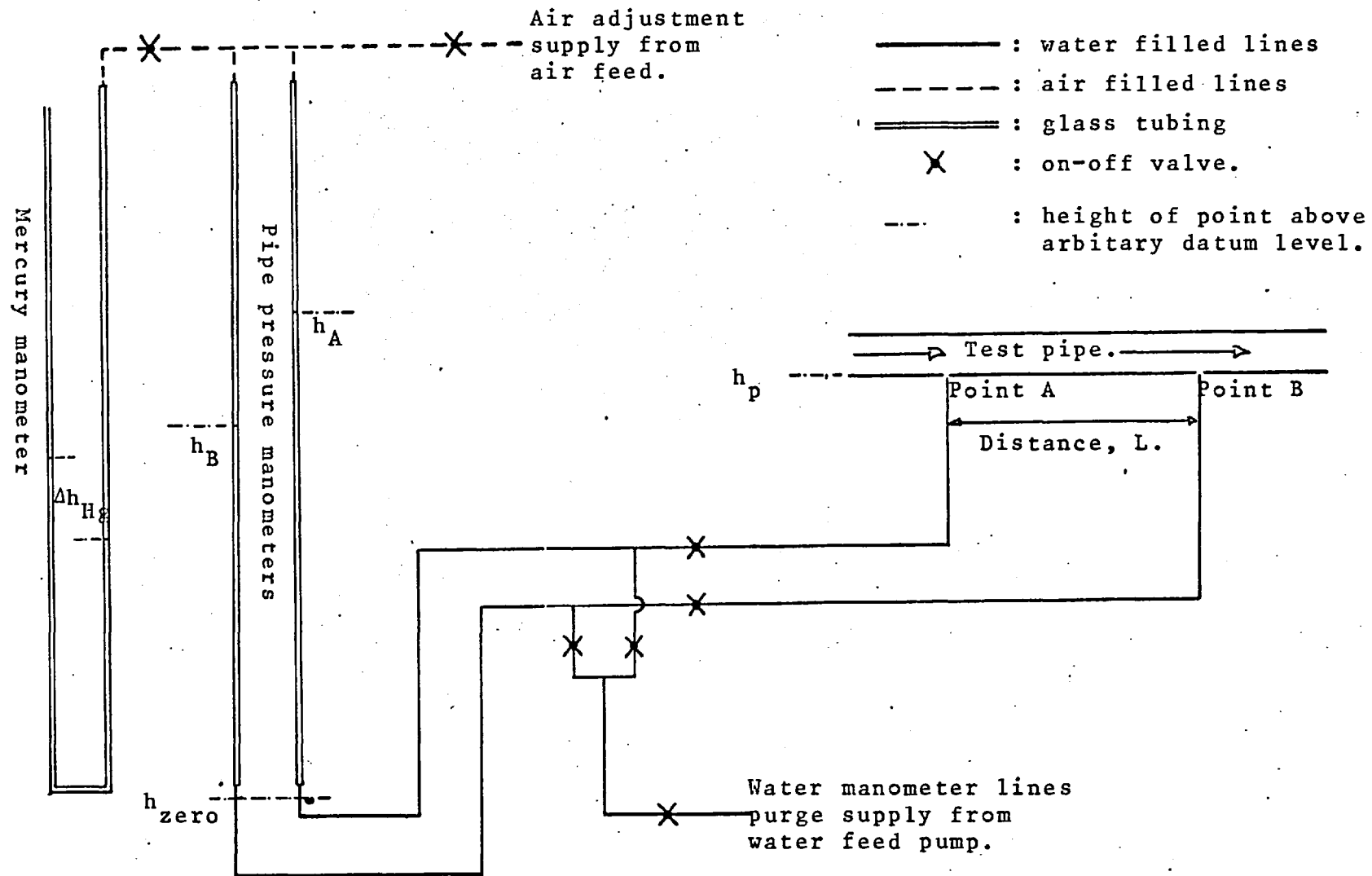
$$(P_A - P_B) = \rho_w \cdot (h_A - h_B) \cdot g \quad 3.1)$$

and

$$P_A = P_{atm} + \rho_{Hg} \cdot \Delta h_{Hg} \cdot g + \rho_w \cdot (h_A - h_B) \quad 3.2)$$

where Hg ; mercury, P_{atm} : atmospheric pressure.

Figure 3.3: Schematic of a pair of manometer tapping lines.



and

$$P_{\text{mid}} = \frac{1}{2}(P_A + P_B) \quad 3.3.)$$

The valves of h_A etc., were measured from the datum with a metre rule.

3.1.4.2) Tapping Points on Pipe

The tapping points were drilled in the glass pipe using a $\frac{1}{8}$ " high speed diamond impregnated glass drill. The holes were drilled various distances apart on the pipe, the distance between hole centres being subsequentially measured. Connections from the pipe to the manometer lines were made with a short length of $\frac{1}{8}$ " copper tubing, set in a perspex rod, which was glued to the pipe. The end of the perspex rod was machined to be a close fit to the outside of the pipe, and glued using Araldite. A gap of $\frac{1}{2}$ cm was left between the copper and glass tubes, so the presence of any bubbles could be noted.

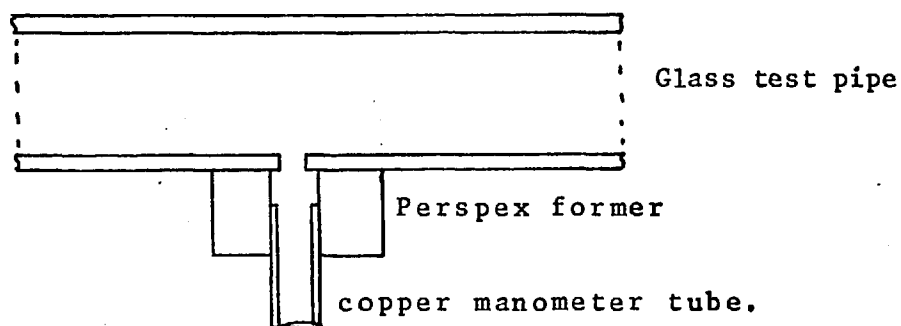


Figure 3.4: Tapping point on test pipe.

The copper tube was connected to the manometer lines with a length of clear P.V.C. tubing, so that checks on bubbles in the manometer lines could also be made.

3.1.4.3) Behaviour of System in Use

The water manometer system depicted was easy to use once the order of manipulation of the valves, for purging and manometer level control, was mastered. The large number of valves to operate when six tapping points were in use took great care, as the correct sequencing was necessary. The final adjustments of manometer levels and line purgings had normally to be done twice for each particular foam flow rate, the time taken to complete this for six tapping being about five minutes.

The results obtained in the experiments gave a high degree of repeatability, and little difficulty was experienced in taking the readings. The pressure drops in foam flow are remarkably steady, only in the slug flow cases was any oscillation of manometer levels noticeable.

The only main problem encountered was found to occur at the extremes of gas or liquid flow rates, when the generating devices were operating at their limits. The slight drift in flow rates which occurred caused problems in the manometer set up cycle, as it took so long to effect in relation to the flow rate changes. Under these conditions the number of correction cycles was reduced to two to speed up manometer adjustment.

3.1.5) Tank Cleaning for Pure water experiments.

The experiments in which pure water-air flows were investigated had first to be prepared for by a thorough cleaning of the complete rig to remove all traces of the detergent. After long use the equipment had detergent traces which were only removed after a long and very careful systematic cleaning procedure.

The old solution was drained away, fresh water added and pumped through all the water rotameters, pipework and liquid return line, while being run to drain. While this was continuing both the tanks were washed and scrubbed internally and all manometers and lines flushed with mains water. A water volume equivalent to about five water loads was used for washing, and the tanks scrubbed over twice, in rotation from the clean water inlet end. At the end of this cleaning the surface tension of a solution held for an hour in the rig was around 68 dynes/cm. and would give a transient surface foam on brisk agitation. The system was completely filled with water and left overnight. The system was then drained, and washed and scrubbed with another load of water. Most of the final detergent traces were removed by the long overnight soak, as adsorbed detergent leached out. The system was then flushed with clean water until no foam would form, and the surface tension of solution passing round the system was greater than 70 dyne/cm.

If the sinter generator was to be used in the experimental runs it was back-washed during the cleaning, to remove traces of detergent in the pores of the disc, and then dried.

The water used in these tests was changed continuously during the duration of the runs, as if left in the equipment for any length of time the detergent levels would slowly rise beyond a tolerable limit. The equipment was clean enough so

that no tendency to form any foam, however transiently was apparent and the surface tension was always exactly similar to that of the pure mains supply.

3.2) Foam Generation Equipment

A range of devices were investigated and employed for the generation of foams. The degree of phase interaction was altered to show the effects of bubble size, on the flow regimes and pressure drops of the air-detergent solutions.

The degree of phase mixing varied greatly, from effectively none with the T-piece mixer, via good mixing with a Sinter disc to give a fine air dispersion, to intense agitation in a centrifugal pump. No dimensional dependence on the performance of any generator was investigated, as only foam of these general types was required. For any generating device there will be an optimum dimension for a particular foam flow rate. Due to the wide range of flow rates employed, and the lack of any meaningful definition of the 'best' foam, the optimisation of the generators was not of importance in this work.

3.2.1) T-piece Mixer

The T-piece mixer was used to introduce the air and detergent solution into the pipe with the minimum of phase interaction. The mixer was devised to separate the effect of generator from that of pipe in creating foams and changes in flow regime.

The mixer consisted of a standard QVF glass T-piece, used such that the arms were all in the horizontal plane. The air flow entered through the side arm, and flow in the T-piece was stratified, as shown in the Figure 3.5. The surface of the lower water layer was not disturbed in any way, no bubbles or waves being generated by the entry of

either phase. This situation changed rapidly within several pipe diameters of entering the pipe, as a range of flow regimes were established.

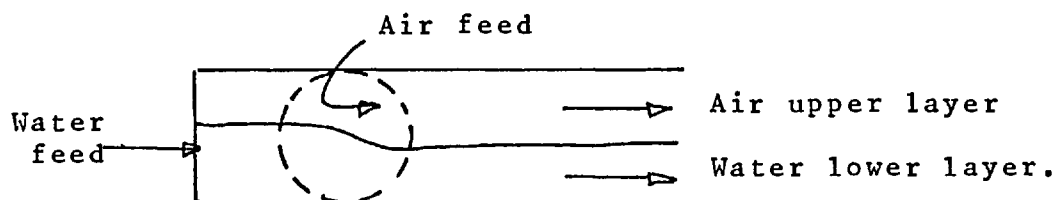


Figure 3.5 : Arrangement of the T-piece Mixer.

3.2.2) Sinter Generator

The air was injected into the liquid via a Sintered disc positioned in the liquid, see Figure 3.6.

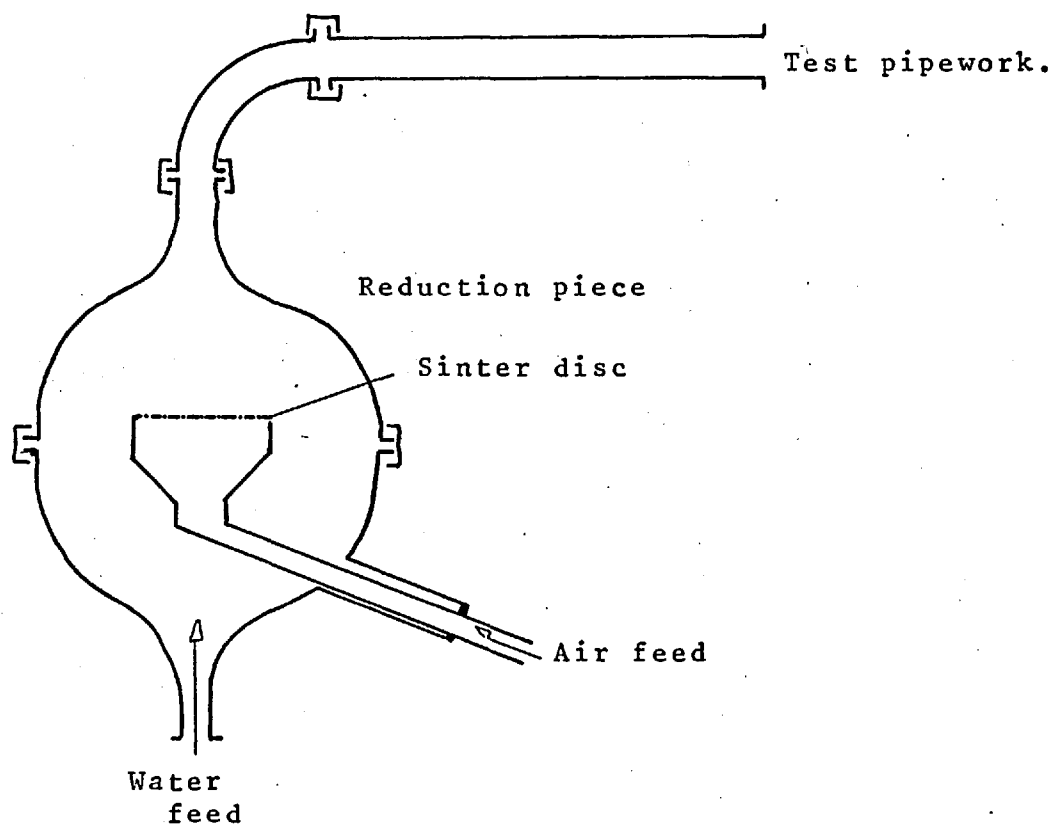


Figure 3.6 : Sinter generator, and pipe to test pipe.

The main vessel consisted of two 6" to 1" QVF reduction pieces, one with an arm fixed to take the Sinter. The sintered disc was sealed into a glass tube, and fed from below with air. The surface of the Sinter was kept level and was just above the joint between the reduction pieces.

A Sinter of porosity 1, pore size 100 - 120 microns, was used for the gas distribution. The diameter of the disc was such that oscillations or blanketing of the disc would not occur at the highest air flow rates. This was to limit the effect of the air injection method on the flow-behaviour as any slugs of air being formed would set up disturbances in the flow.

It has been shown (68) that the foam column produced from a Sinter in a large water mass, free edges to the Sinter and a reasonable depth of liquid, will start to pulsate if

$$Q_p > 1.3r_s^{\frac{5}{2}}g^{\frac{1}{2}} \quad 3.4)$$

and the disc blanket if

$$Q_b > 7.3r_s^{\frac{5}{2}}g^{\frac{1}{2}} \quad 3.5)$$

The maximum air flow rate expected was 60 litres/min and the size of disc required to stop

a) pulsation : $r_s = 3.60$ cms.

b) blanketing : $r_s = 1.81$ cms.

A disc of radius 4.5 cms was used, and sealed into a tube, giving a slightly smaller effective size of disc.

At no time in the experimental work did the disc blanket, even when no surfact^{ant} was present. Pulsation only occurred when water flow rates were quickly changed when using the pump in line, this being caused by a hunting effect of the pump and was easily controlled.

No visual evidence suggested that slugs formed due to the action of the Sinter, slugs of air forming in the pipe in flow regime experiments did not originate from air flow variations from the Sinter.

The Sinter was always operated so that liquid did not run back through below the Sinter. Whenever this happened the water was drained off, as it caused air flow fluctuations due to the liquid foaming below the Sinter.

The bubbles coming off the Sinter varied slightly in size, depending on the air flow rate, pressure above Sinter and detergent concentration. Low air flow rates gave smaller bubble sizes, as did higher pressures and detergent concentrations. As the air rate increased the bubble size would also increase and more sites on the Sinter start bubbling. The number of sites working was increased at higher detergent concentrations. Photographs of the foam produced at constant water rate and increasing air rate are given in the Appendix.

3.2.3) Valve Generator

The valve generator was used to produce a fine bubbled foam by the intense mixing of the phases. The valve generator consisted of a standard QVF 1" glass valve, positioned at the start of the test pipe. The design of the valve was of the simple stop-cock pattern, as in Figure 3.7.

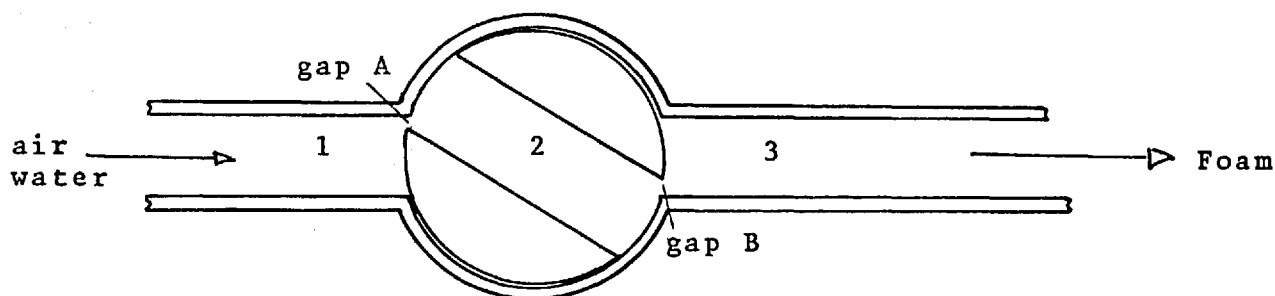


Figure 3.7: Valve generator, with tap in the operating position, giving small gaps for flow.

The generation of the foam occurred due to the high shear forces acting on the air and water mixture passing through the valve. The tap of the valve was adjusted so that the gaps A and B between section 1 and 2, 2 and 3 were sufficiently small that a high velocity occurred. The shear forces acting were sufficient to break the air into many fine bubbles. The region 2 in the tap acted as a smoothing chamber, the jet of air and water coming into this section was broken into foam and the phases mixed in this region, this foam would pass out through gap B into the pipe.

The valve depended for its smooth operation on a constant supply of air and water, slugs of air could be turned into foam, but once the size of the slugs was such that the centre section could not smooth out the phase fluctuations, the quality of the foam deteriorated. To ensure an even supply of air and water the Sinter generator was used to inject the phases into the pipe leading to the valve, but this high degree of phase pre-mixing was not essential to the operation of the device.

The pressure drop over the valve was large, as high speed flow had to be attained in the valve gaps if good foams were to be produced. A pressure drop of approximately 11 psi (.75 bar) over the valve was found to be adequate for uniform foam production. At any phase flow rate the valve tap was adjusted to give this value. The foam produced by this simple valve was found to have the same pressure drop V'S flow rate behaviour, and exhibit the same flow regimes, as that produced by the pump, and so was not extensively used.

3.2.4) Pump Generator

The pump used as a foam generator was a No. 12 Stuart Turner centrifugal pump. This was positioned in the line with suitable copper pipework, such that the outlet from the pump passed directly into the test piping.

The pump was a high shear device, and produced a fine bubbled foam, a photograph of which can be seen in Figure E2 in the Appendix. The action of the impeller was to break up the air going into the pump and mix it intimately with the water. The pump speed changed as both the amounts and relative proportions of phases introduced varied, but the structure of foam produced appeared constant. The flow limitations of the pump were

- a) Minimum water rate of ≈ 3 L/min, below this value the pump speed would start to hunt with rapid fluctuations, causing sudden changes in foam flow.
- b) Maximum air rates depended on the water rate as the pressure drop in the test pipe was a limiting factor at the high flow rates. However air to water volumes ratios in the pump of up to 8 X were reached, even with relatively low water flow rates.
- c) Minimum air rate was limited by the rotameters, which were inaccurate below 1 L/min.
- d) Maximum water rate was purely a function of the pressure drop needed for the test pipe. A larger water feed pump, and air supply pressure would have been required for higher values.

The air/detergent solution feed to the pump was generated either by the valve or Sinter generator. An even mixture gave better speed control of the pump, but was not essential for the production of a fine bubbled foam.

The pump was used with an air/pure water feed, and did not produce a fine dispersion, giving the normal regime flows for air-water mixtures. The presence of detergent was necessary for the operation of the pump as a foam generator.

3.3) Foam Sample Production

The provision of small quantities of a constant and uniform quality, of the foam, was required for initial testing and exploratory work. The use of foams in rheological studies creates several problems of unusual nature, the fluid in the required state being of such an extremely transient nature that the fluid sample cannot be kept while experimental equipment is adjusted or allowed to settle into a steady state. The main test equipment was designed for the production of large quantities of foam, and was not suited for the generation of small samples. Techniques which would give a sample of foam, of variable gas-liquid ratio with an even bubble size, and having the same properties as that used in the flow experiments were investigated.

3.3.1) Foam Condensation Method

Samples of foam were produced by releasing from pressure a sample of carbon dioxide saturated detergent solution. The reduction of pressure causes the CO₂ present to condense out at nucleation sites in the bulk of the liquid. This method is akin to the methods used for aerosol can foams, where a low boiling point solvent is used to provide a gas phase. The foam was made with the use of a standard sparklets soda-water syphon, the required detergent solution being used in place of water (5). The CO₂ was provided by the use of the CO₂ cartridges provided for the syphon, or by filling off a large CO₂ bottle with reducing valve. Control of the foam quality, air-water ratio, was possible by either varying

the amount of solution in the syphon, when cartridges were employed, or by changing the pressure in the syphon. A pressure of 45 psig was adequate to give a gas-liquid ratio of 2:1.

The foam produced by the syphon was of an extremely even and uniform consistency, with an even bubble size distribution. The gas bubbles formed were created in the bulk of the fluid, the bubble nucleation and growth taking place at the high velocity/low pressure region in the controlling valve. The use of CO₂ as the gas phase had very little effect on the initial properties of the foam, and so gave equivalent results to an air expanded foam.

3.3.2) Small scale Air-foam Generator

A method of producing a small quantity of uniform foam using air, or any other low solubility gas, as the gas phase was also developed. Not all gas/detergent solutions combinations will be compatible or stable, and so an alternative method giving fluid choice flexibility is required to cover all possible phase combinations. One of the main factors governed by the gas phase is the diffusion between bubbles, giving bubble growth and coalescence (1,72). Many of the small scale methods employed by other workers have used combinations of heating or agitation to mix the phases, these methods tending to give comparatively large bubble-sizes.

The basis of the foam generation technique was a scaled down version of the large pump/valve generator used in the pipe tests. The use of a similar process resulted in closer properties of foams made by the large and small scale

equipment. The equipment is shown schematically in Figure 3.8 and consists essentially of a closed pumping loop. The main parts of the equipment are the holding vessel, pump, cooler and valve.

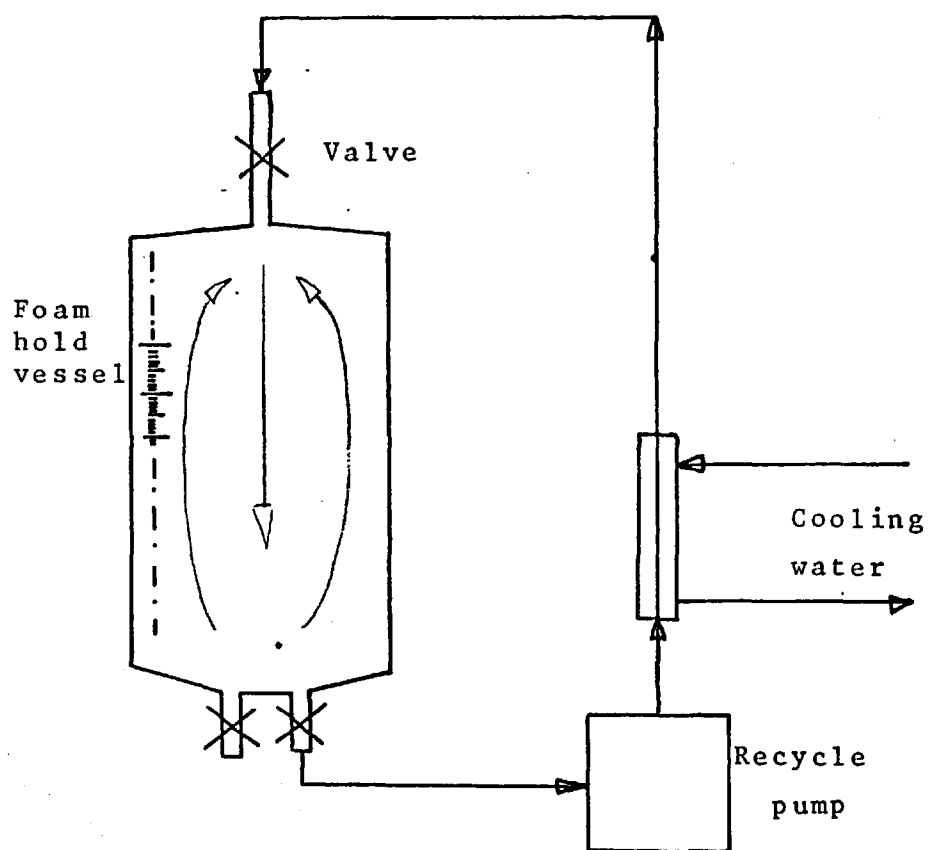


Figure 3.8: Schematic of the small scale air Foam generator

Foams of different air content can be generated by varying the amount of liquid in the vessel, then running until a uniform foam is produced. The foam was generated as the liquid, and after a short time the air as well, was pumped round the system. The action of the pump and valve break up any large bubbles that are carried round in the liquid, see Section 3.2 for details on valve and pump generators. After about 10 minutes the foam in the vessel reaches a uniform consistency. The foam was cooled to remove the heat generated by the pump motor, which would cause temperature rises of about 3 °C in ten minutes.

Turbulence in the vessel was high, the flow pattern being indicated on the figure. Soon after start up the cycle time for bubbles in the vessel was about 10 seconds, when the foam sample had reached equilibrium the cycle time was about 30 seconds, far faster than any bubble rise or drainage flows which occur in a static foam. The turbulence caused by the foam entering the vessel mixes the phases together, and then they are intensely mixing to give fine bubbles by the action of the pump and valve, the valve was adjusted to create high velocity flow through it.

Tests on foam drainage of air-detergent solution foams were carried out in the holding vessel. The vessel was constructed to have similar dimension as that devised by the Fire Research establishment for foam drainage tests (16).

Vessel diameter = 9.5 cms.

height = 20.0 cms.

Total volume between valves = 1445.6 cm³

A centimetre scale was fused into the glass to give a permanent marking and reference system.

3.4) Foam Expansion Measurement

The measurement of the relative amounts of air-liquid in a foam sample was carried out by the use of a measuring cylinder.

A sample of foam was run into a graduated cylinder and the foam volume noted. The amount of liquid in the foam was measured by either a) allowing the foam to collapse and reading the liquid volume off the scale. b) Weighing the foam sample in a tared cylinder. For dry long half-life foams the time taken to drain completely was so great that the measurement of liquid volume by drainage was impracticable in a series. A density of 1.00 gm/cm^3 was assumed for the liquid, no correction being necessary for the mass of the gas present in the foam

3.5) Foam Drainage Experiments

Foam drainage experiments were carried out in a range of vessels of varying diameter and height. Vessels used being

a) The small scale foam generator vessel

Diameter = 9.6 cms

Height = 20.0 cms

b) A series of measuring cylinders

1) 1000 mls ; ht = 34.8 cms diameter = 6.05 cms

2) 500 mls ; ht = 28.4 cms " = 4.74 cms

3) 250 mls ; ht = 22.5 cms " = 3.76 cms

Tests in vessel a) were carried out using the full height of the vessel, tests in vessels b) were carried out using a range of different initial foam heights in the cylinders.

3.6) Photographic Details

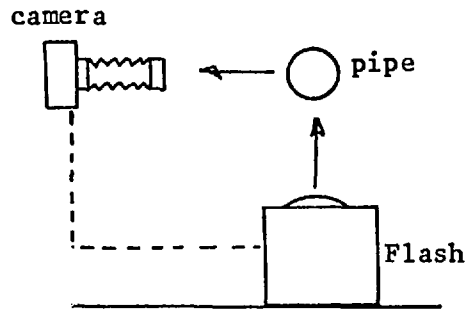
Photographs were taken of the various types of foams flow in the pipe, to investigate bubble size and distribution, and also to show phase distribution and flow pattern existing in the pipe. All foam photographs were taken of the foam flowing in the pipe under the conditions of interest, so as to obtain information on the foam as it existed in the pipe, rather than taking photographs of static samples which were extracted from the pipe.

3.6.1) Foam Appearance

The foam appears visually to the naked eye as a white opaque fluid, and except for very dry foams only the bubbles in the outermost layer, next to the wall, can be distinguished. The close packing of this outside layer, and the reflections from it, mean that any view of bubbles in the interior of the foam is impossible. However with the dryer foams at low velocities, the movement of the next few layers into the foam could be seen under certain conditions, when it is obvious that the layers further away from the wall are moving more rapidly than the layers adjacent to the wall. The presence of air slugs in the foam, essential for regime identification, was easily noticeable by eye as the change in optical density under back lighting as the slugs moved was very distinct. This effect however was not recorded by the photographic technique, which used the light reflected from the outer foam layer, which was not destroyed by the passage of the slug.

3.6.2) Lighting Method

The mode of lighting employed was a high speed electronic flash, synchronised with the camera shutter. The light source



orientation was as shown in Figure 3.9, the light direction being 90° to that of the camera view. This position of side lighting of the foam was arrived at after tests

Figure 3.9: Light orientation.

on the effect of light orientation on the photographic record of the bubble. This arrangement gave the best bubble definition, the shadows cast by the bubbles permitted good bubble size measurements, without causing excessive glare from reflections off the pipe and bubble surfaces. The difficulty in attaining adequate illumination was severe, and basically consisted of overcoming the problems of reflections and the effect of the high bubble density. Transmitted light, through the foam, gave very poor results due to the opacity of the foam, and the effect of multiple reflections in the bubbles. Front lighting at angles less than 90° reduced the shadow cast by the individual bubbles, and caused problems with reflections off the glass surfaces.

3.6.3) Cameras Used

Various cameras were used for taking the photographs, depending on the magnification necessary.

a) A 4' x 5' plate camera, with extension bellows.

This was used for taking photographs of the whole pipe for flow regime analysis. Polaroid High Speed film, 51, with an ASA of 3000 was used, and the image size on the print

varied from life size to 2 x magnification. Because of film speed the aperture size was small enough that the whole pipe was in focus.

b) A 35 mm Leica - coupled to extension bellows. This gave a magnification of 4 x on to the negative and was used for photographing sinter generated foams. Film used was Kodak Tri-X, ASA 400.

c) A 35 mm microscope camera, coupled to microscope. This gave a magnification of 10X on to the negative and was used for the valve/pump generated foams.

The plate and 35 mm camera viewed the pipe in the horizontal plane, with light source either vertically above or below the pipe. The microscope camera was used to view vertically downwards, with the light in the horizontal plane, this being the only mode of employment possible with the particular equipment. All the photographs were taken under a black-out facility, consisting of a frame-work built around the pipe and camera system, and covered with heavy black cloth, to exclude extraneous ambient light.

3.6.4) Pipes Used

Photographs of the foam were taken through the walls of either a square sided tube, of the same cross-section as the pipe, for 17 and 11 mm pipes, or through the wall of the pipe, 25 mm, when the curvature of the pipe wall was such not to cause excessive distortion. The photography of regime patterns were all taken of the whole pipe, with an optical system of adequate depth of focus to ensure the whole pipe was in focus. The square section tube for the 17 mm pipe was made up out of plate glass, cemented together and joined with smooth transition pieces to the round pipe to minimise flow

disturbance. The square section for the 11 mm pipe was constructed using a square drawn tube made with a suitably sized graphite former.

3.6.5) Focusing Arrangement

The focusing of the plate camera caused no problems, the extremely small aperture used, $f64$, gave a very deep field of focus, and once the enlargement required had been set with the bellows expansion adjustment, no other adjustment was necessary.

The other two camera systems did however require careful setting up, if a clear picture of the bubble layer next to the wall was to be obtained. The use of the systems with such large magnification, coupled with the low apertures used. $f4$ to $f3.5$, gave an extremely small depth of focus, usually less than 0.5 mm, but at least smaller than the size of the larger bubbles present in the foam being photographed. Initial focusing was carried out by using a static dry foam in the pipe, and focusing on the layer of bubbles against the inside wall of the pipe. With a large bubble dry foam, bubbles greater than 3 mm, it was possible to focus on internal bubble walls, up to about a distance of two bubble diameters into the foam. The foam when flowing had a layer of water between the outer bubble layer and the wall, and this bubble layer was no longer in focus, as set up in the static test. The optical system was moved inwards in stages, using the micrometer adjustments on the microscope or bellows stand, and a series of photographs taken at each position.

The best focused set being used for further analysis. The foam flow was set at the start of each run and kept constant over the period of photographing the foam.

An accurate scale was photographed with each camera set up to allow the scaling on the prints to be determined.

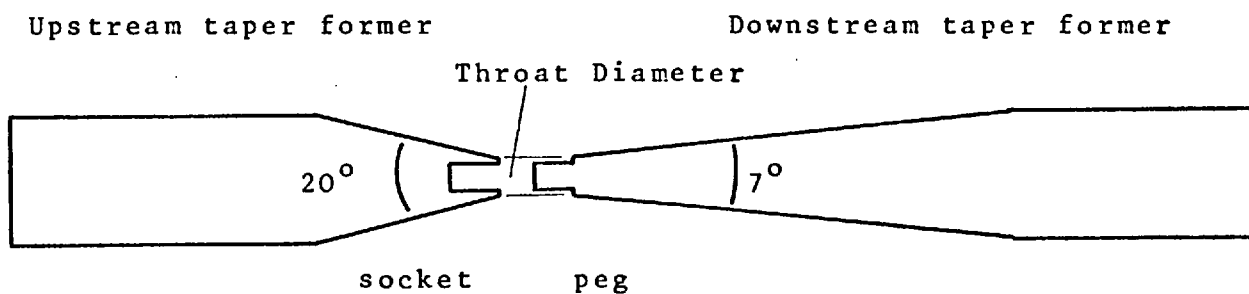
3.7) Accurate Glass Venturi Nozzle Manufacture

The venturi nozzles used for the initial foam nozzle experiments were made by the free-hand manipulation of the glass pipe, the pipe being heated and drawn until an acceptable shape with the approximate throat dimension was obtained. The resulting nozzle was adequate for qualitative tests, but due to its non-uniformity and uncontrolled shape was not suitable for experimental work for the investigation of foam flow through a venturi nozzle.

Hence the manufacture of a more precise venturi nozzle was undertaken, and the process finally devised was as follows:-

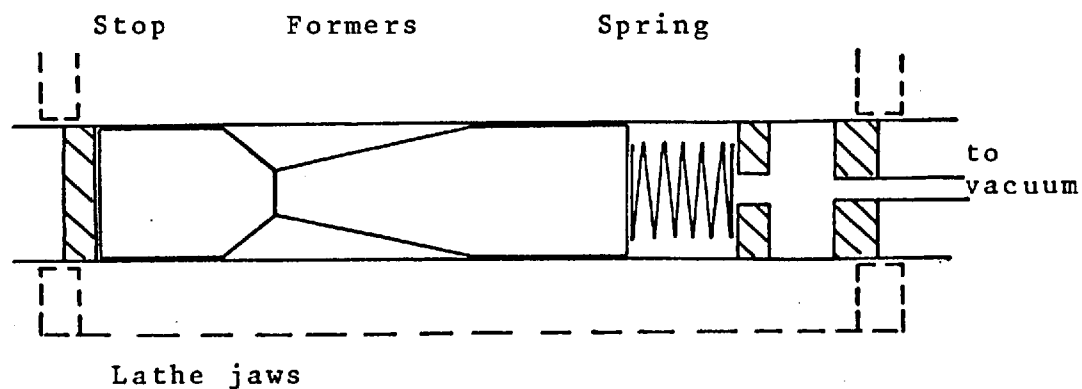
i) The shape and size of the nozzle for the experimental work was first calculated, and the dimensions determined. Tapers of 20° on inlet, and 7° on outlet were used, and a throat diameter so as to give a velocity of over 20 m/s at the throat for a flow rate of 30 litres/minute, 0.45 cms.

ii) Metal formers of the inside shape of the venturi were then fabricated out of stainless steel. The formers and dimensions were as shown in Figure 3.10. The peg and socket was used to locate the formers accurately together so that no lip was formed at the throat and the nozzle axis be kept true, the peg dimension being dictated by the throat diameter. The length of constant diameter former after the tapers was required to stop the formers 'slopping' about as the pipe is turned, about 3 pipe diameters are needed. The surface finish was good, but was not polished as a graphite release medium was used to coat the formers to allow easier removal once the glass had been shrunk on, this coating also gives the formers a smooth finish.



Pipe diameter, 17.0 mm. Throat diameter, 4.5 mm. Tapers, $20^\circ, 7^\circ$.
[not to scale]

Figure 3.10: Former used to make Venturi Nozzle.



[not to scale]

Figure 3.11: Formers in pipe ready for the start of the glass working process.

iii) The formers were then set up in the pipe, of slightly greater diameter than the formers so they would just slip in, as shown in Figure 3.11. The upstream former is butted against a stop, the down stream former positioned, taking care that the throat joint as a good fit. The two formers were held together by the spring behind the second former, used so good contact is made when heating causes expansion to take place. Facilities for applying a vacuum were arranged and the whole pipe and formers held in a glass working lathe.

iv) The pipe was then heated while being turned in the lathe, heating starting from the far end to the vacuum tube. The glass melted and was sucked on to the former shape. This operation needed care and skill to obtain a smooth collapse of the glass on to the former so as not to leave any air pockets between the glass and former. The heat must not be excessive, or the metal formers would distort and the glass become opaque.

v) When cold the metal formers were removed and the nozzle annealed to remove stresses.

vi) Tapping points for pressure measurements were then drilled as required using an ultra-sonic diamond tipped drill piece, great care being taken to avoid any spalling as the drill broke through the nozzle wall.

The nozzle so formed was reasonably strong due to the thickness of glass at the throat, and was placed in the test line without additional support, after the necessary end pieces for Q.V.F. piping had been fitted to the nozzle. The thickness of glass around the throat was several times that of the normal pipe wall, but observations of the foam flow could be made.

4) Foam Drainage

The mixture of air and detergent solution in the foam will separate after a time, and the initial homogeneous structure of the foam lost. The foam drainage can therefore be used as a measure to define foam stability.

The foams investigated had air/liquid volume ratios which did not give close packing of the bubbles. The liquid drainage in this first stage will be caused by hydrodynamic forces, the surface tension forces only coming into play when the water layers between bubbles is small. Many of the foams of interest to other workers have been comparatively dry foams, with drainage characteristics which depend on the surfactant and bubble size (1, 72, 74, 65). The characterization of high density fire fighting foams by the time taken for 25% of the liquid to drain is well established (16). The drainage characteristics, and parameters which can effect the 50% drainage time of CO₂ expanded high density foams have been investigated (5).

4.1) Foams Generated with Intense Phase Interaction.

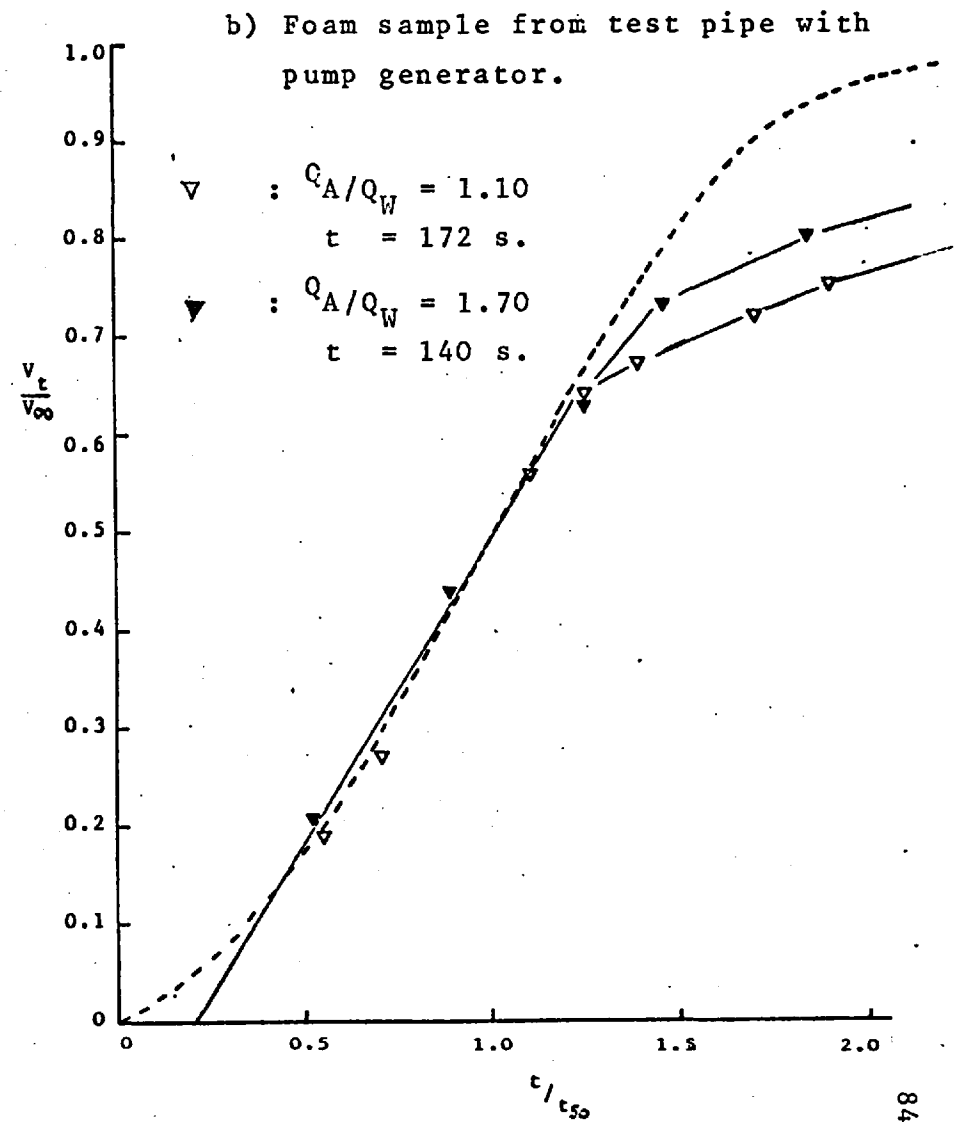
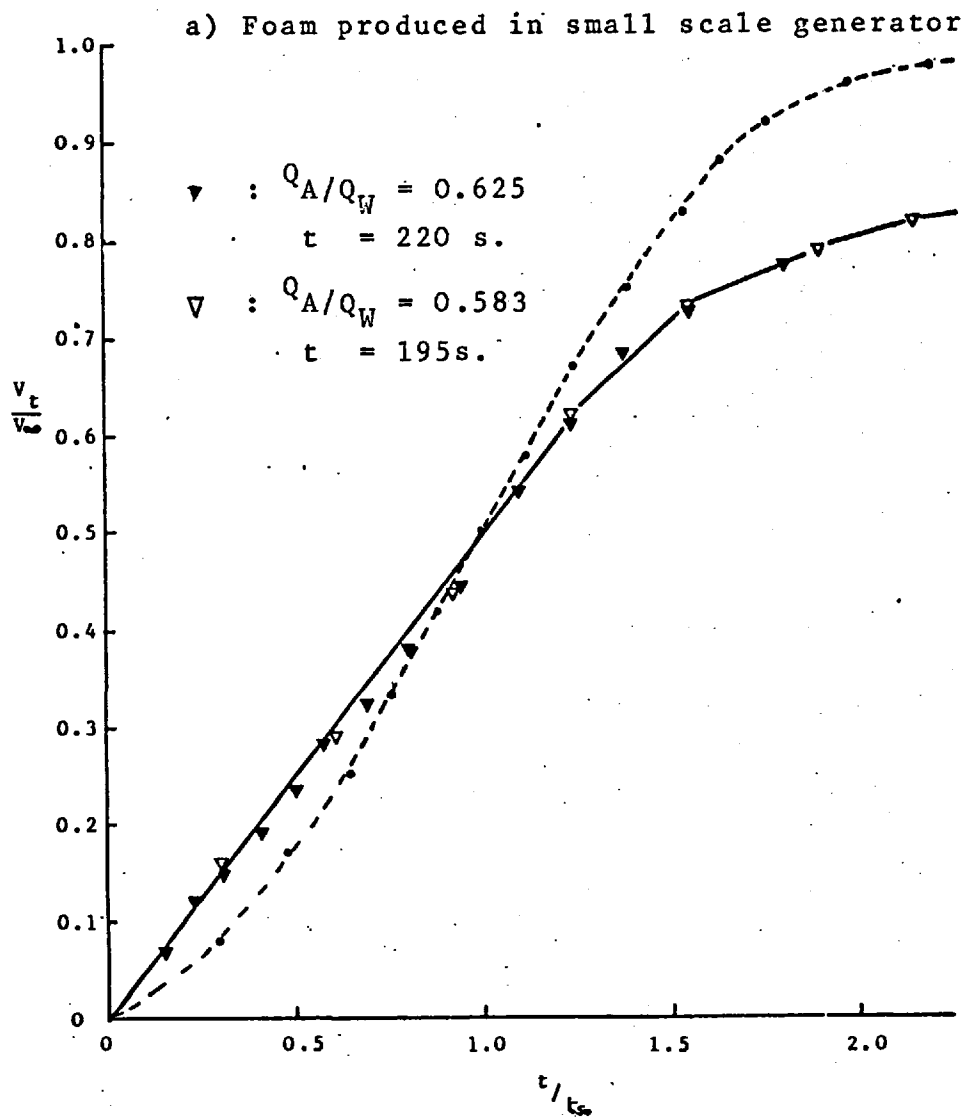
The drainage characteristics of foams produced with the valve/pump generators were investigated, using foams produced in the small scale air-foam generator and samples from the test pipe.

Results of foam drainage for foams produced in the small scale generator are shown in Figure 4.1. The results have been normalised so that the general shape of the drainage curve can be seen.

$$\frac{V}{V_{\infty}} = \frac{\text{Volume drained at time } t}{\text{Total volume of liquid}} \quad 4.1$$

$$\frac{t}{t_{50}} = \frac{\text{time}}{\text{time for 50\% of liquid to drain}} \quad 4.2$$

Figure 4.1: Drainage curves for the Foams produced with intense phase interaction, in the small scale and in test pipe.



Of note is the constant drainage rate, being constant until 60% of the liquid has drained from the foam sample. Data plotted was obtained by measurement in the holding vessel, and also by removing a sample and draining in a measuring cylinder. The bottom interface between liquid and foam was cloudy over this initial drainage period, the small bubbles washed out of the foam having a rise velocity less than the interface rise velocity. Once drainage rate slowed these small bubbles could then catch up with the interface and rejoin the foam.

Also indicated is the drainage curve for CO_2 expanded foam, an initial slow rate of increase, followed by a constant rate until 70% of the liquid drained. No very small bubbles were present in this foam, and the interface was always sharply defined (5).

Results for foam drainage of samples collected from the test pipe are also given in Figure 4.1. The same general shape of curves result, with a period of time elapsing before drainage starts to occur. These curves match those obtained with the CO_2 expanded foam, the discrepancy at short times being due to the inability to define any interface, as the small bubbles obscure the division between liquid and foam. The marked difference in amounts of liquid drained at long times, $2 \times t_{50}$, for the air and CO_2 expanded foams was attributable to the difference in gas diffusion properties. CO_2 diffuses between the bubbles more rapidly than air, and so bubbles will coalesce more rapidly, liquid films break sooner, and liquid drain faster in the surface tension controlled regime (72). This

difference of long term drainage characteristics was further noted when the dry air foam overspilled the foam holding tank, the design of the tank being based on the behaviour of CO_2 expanded foam.

The results shown on Figure 4.1 were obtained when foam samples with an initial height of approximately 20 cms were measured. As the foam drainage was in the hydrodynamic state the height and cross section area of a sample could be expected to have some effect on the drainage time for the foam (16). A range of vessels were used to measure t_{50} for a foam, the CO_2 foam being employed as a more exact control of air/liquid ratio and foam sample volume was possible (5). The results are shown in Figure 4.2, and this particular type of behaviour could be expected to apply to any similar foam.

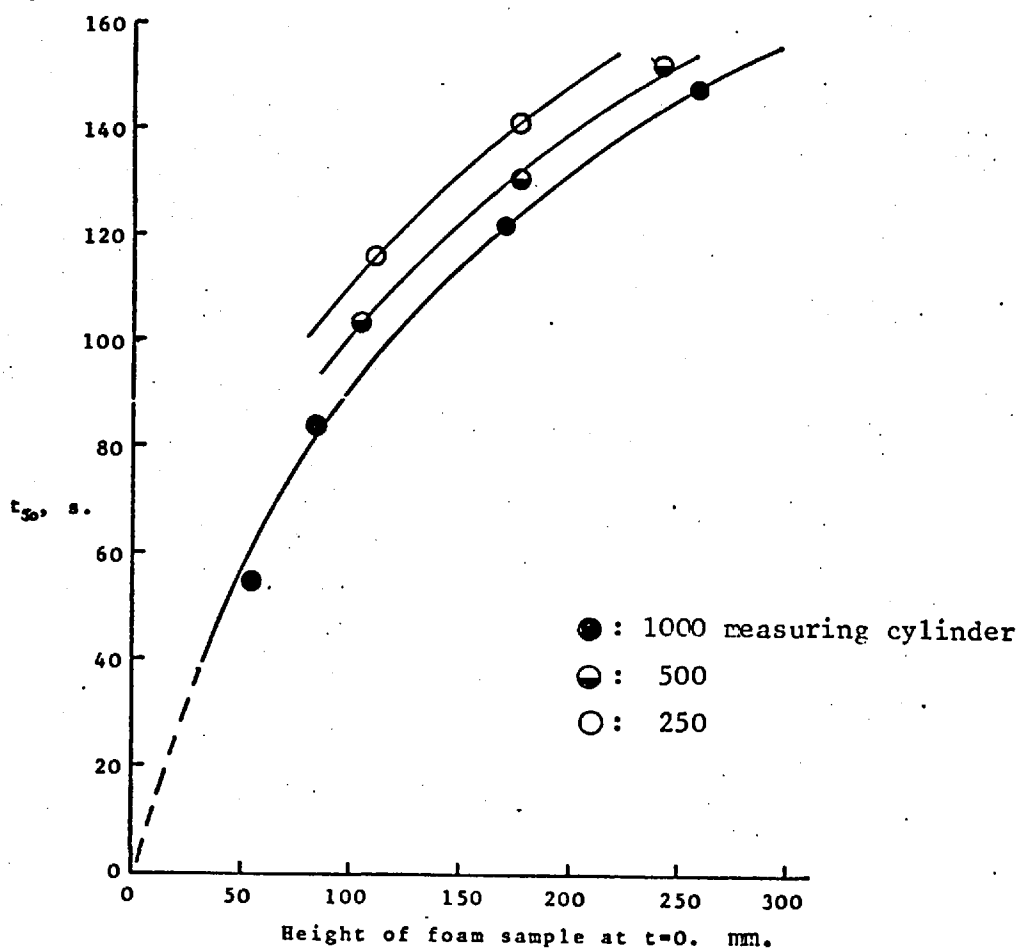


Figure 4.2: Variation in Half-life with Foam sample height.

These results would suggest that a foam made with the pump generator would have a t_{50} of 0(10 secs) in the test pipes.

The hazy interface between the liquid and foam layers was caused by the presence of small bubbles washed out of the foam being unable to rise as fast as the interface. The interface became sharp once the rise velocity of the interface was less than 0.03 cm/sec, this value being measured in the graduated cylinder. The rise velocity of these small bubbles can be approximated by the use of Stokes Law,

$$v = \frac{g \cdot (\rho_f - \rho_l) \cdot d^2}{18 \cdot \mu_l} \quad 4.3$$

which gives a value of d of ≈ 23 microns. Bubbles of this size were measured off the photographs, and the drainage experiments confirm the presence of these small bubbles.

4.2) Foams Generated with the Sinter Generator

The foams generated with the Sinter cannot be considered to be homogeneous. Flow regime investigation showed that stratified foam flow or slugging was the most likely regime, the dispersed and foam flow regimes being at the extremes of flow rate investigation. To measure the drainage characteristics of the upper foam layer the effects of the water layer must be accounted for. The flow regime in stratified foam will consist of two phases

- a) A foam phase, containing all the gas and some of the water
- b) A liquid phase, containing the remainder of the water.

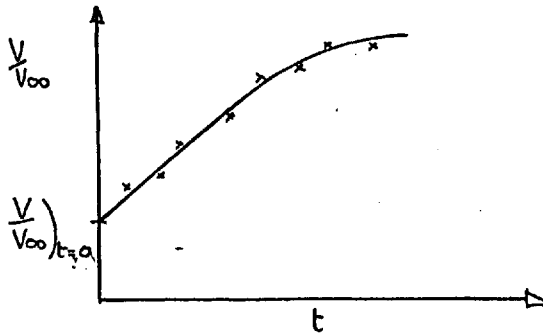
If a sample of this foam and liquid were collected from the pipe, ideally it would consist of a volume of foam, Q_f and a volume of liquid, Q_L . As the sample ages, liquid from the foam drains, and the water volume increases.

Let R_f be the fraction of water feed in the foam, and R_L the fraction in the water layer

$$Q_f = Q_A + R_f \cdot Q_W \quad 4.4$$

$$1 = R_f + R_L \quad 4.5$$

If the amount of liquid drained from the sample is measured with time, and $\frac{V}{V_\infty}$ vs time plotted, the result would be given by the diagram below.



The value of $(V/V_\infty)_{t=0}$ can be found by extrapolating the volume measured back to $t = 0$. The amount of liquid present in the foam, and liquid layer can then be deduced.

$$\left(\frac{V}{V_\infty}\right)_{t=0} = R_L \quad 4.6$$

The drainage rate for the foam can also be found by plotting $(V - R_L \cdot V_\infty) / R_f \cdot V_\infty$ vs t . A line passing through the origin, similar to that obtained with the pump generated foams results.

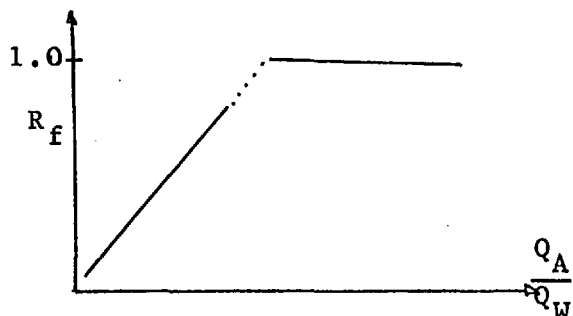
The amount of liquid contained in the foam will depend on the relative gas/liquid volume ratios. From the regime flow maps it is obvious that at any fixed water rate there is an air rate which is sufficient to include all the water in the foam, the stratified foam to foam flow boundary indicating the required ratio.

When all the liquid is included in the foam the void fraction of the foam will equal that of feed to pipe,

$$H_f = H_{\text{feed}} \quad \text{at } R_f = 1 \quad 4.7$$

So at any fixed water rate, with an increasing air rate,

there comes a point when $R_f = 1$, and $H_f = H_{\text{feed}}$. A plot of R_f v's Q_A/Q_W will have the form shown in the figure.



- 1) As $Q_A \rightarrow 0$, foam volume goes to zero, $R_f \rightarrow 0$.
- 2) Q_A increases, some liquid in foam, rest in stratified layer.
- 3) When critical value reached, the foam contains all liquid, $R_f = 1$.
- 4) At greater Q_A/Q_W , foam starts to break down and slug.

The value of Q_A/Q_W needed to give $R_f = 1$ would intuitively be the volume ratio needed for close packing of the bubbles.

Results of sample drainage from the 39.1 mm pipe are shown in Figure 4.3. A fixed water rate, with increasing air flows was used, values of R_L read off the plot, and drainage curves for the foams constructed. Tabulated data is given in the Appendix.

The foam produced with a Sinter generator has a $t_{50} \approx 15$ secs. Errors in measurement will be about 10% due to methods employed. This t_{50} was for a foam height of approximately 200 mm, drainage time in the pipe will be of order 1 or 2 seconds.

The stability of these foams was low, and the rapid drainage indicates the reason for stratified flow being the primary flow regime.

The fraction of liquid in the feed which is included in the undrained foam can be seen in Figure 4.4.

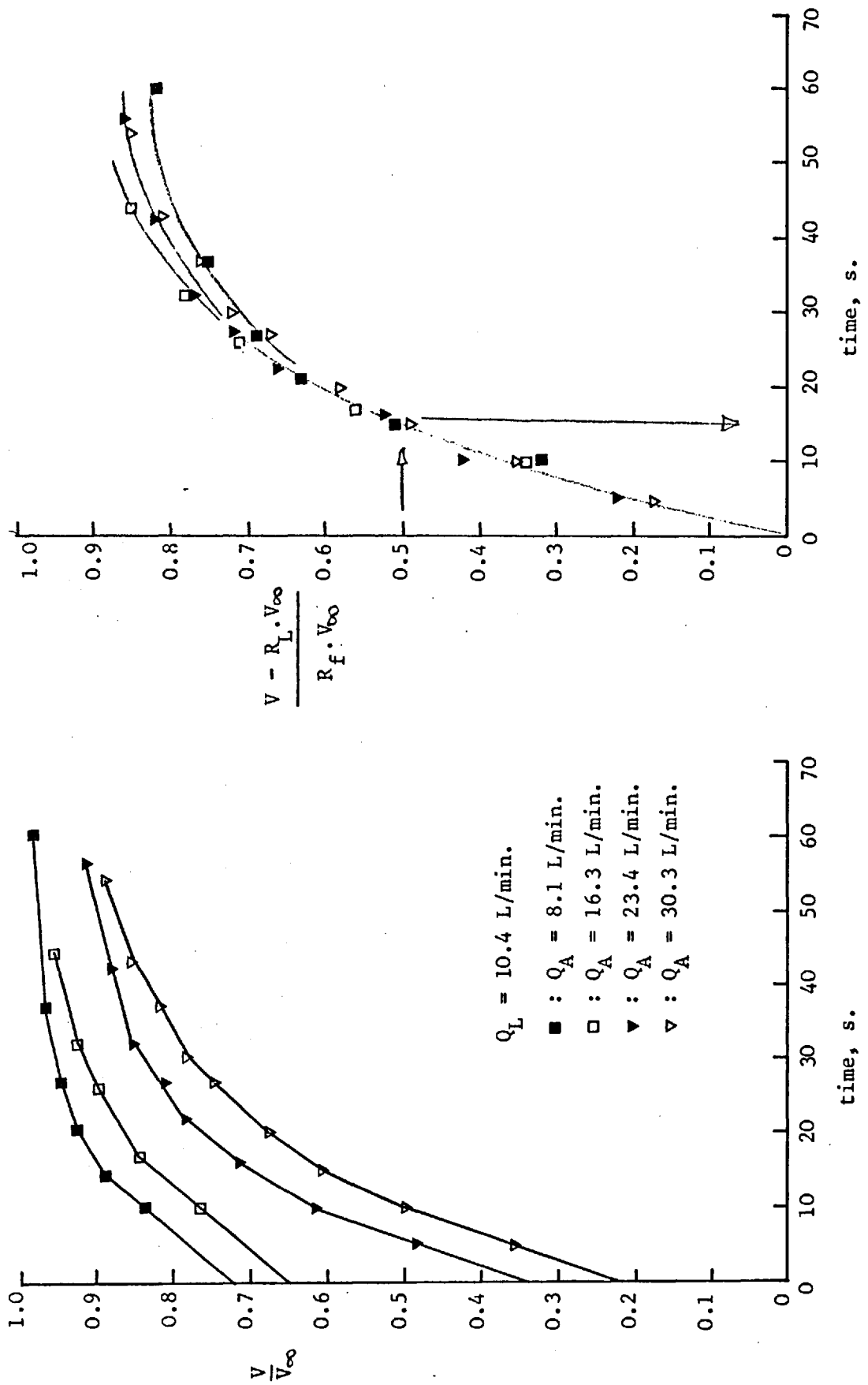


Figure 4.3 : Drainage of sinter generated foams, from 39.1 mm. showing effect of liquid layer, and drainage curves for the foam alone.

A Q_A/Q_W ratio of 3.6 was required to form a uniform mixture, this compares with the value of ≈ 3.0 which would result if the foam consisted of mono-sized bubbles in close packing.

The stratified foam to foam transition for flow in the 39.1 mm pipe, at a water rate of 10.4 l/min, can be estimated to occur at ($V_L = 0.15$ m/s; $V_G = 0.62$ m/s) where $Q_A/Q_W = 4.0$.

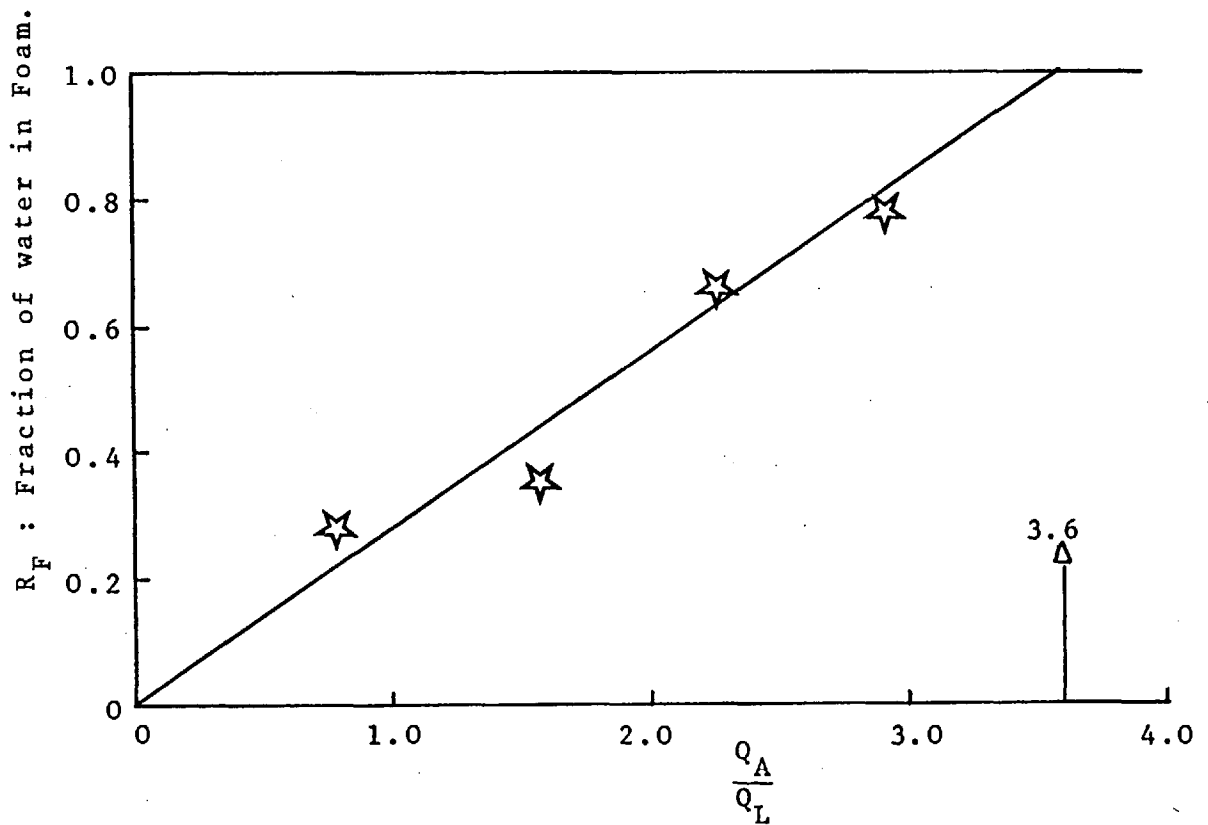


Figure 4.4: Variation of R_F with phase ratio, in 39.1 mm. pipe.

5. Bubble Size Measurement Techniques

The bubble size in a foam is an important parameter, and is one of the major factors in determining the foam properties and behaviour. The rheological properties of a foam could be expected to depend upon the average bubble size, and the degree of dispersion of the bubbles in the liquid, as well as the total number of bubbles present. This factor of the bubble size, and it's effects on the various foam properties, has been studied by many workers (10,15,55,70), some of which have used the average bubble size as a correlating factor to explain their results.

The effects of bubble size on the flow properties of foams are not clear, but what information is available seems to indicate that at high flow rates the bubble size is not an important factor (62). Many workers have carried out foam flow experiments in pipes without attempting to measure and correlate bubble size (4, 63), this in fact is the norm for pure water/air in the dispersed region, where the degree of subdivision of the gas phase is not used in the correlation of results, attempts however being made to get good subdivision so an homogeneous approximation can be made (7, 56). Hence it would appear that little work on effects of bubble size on flow properties has been carried out.

The foams mainly of interest in this work were high density, where the air/water ratio was low, and in many cases the bubbles in the foam are separated by thick lamellae, the bubble density often not being sufficient for close packing to occur. The pipe diameter to average bubble diameter ratio was always large, a value of ≈ 50 for the sinter generator and 250 for pump generator, in the smallest pipe size.

Hence the effect of bubble diameter could be expected to be small, as the foam can be assumed homogeneous when considered on the scale of the pipe.

This work involved looking at two main types of foams, produced with medium and intense interaction of the phases, which gave two distinct bubble size distributions. The generators used did not give a controllable bubble size, but did produce an approximately constant average size and distribution over the operating ranges. The bubble sizes in the foam were not therefore an independent variable, to be adjusted at will over an exact range, only gross changes being possible by generator choice. Hence to have used the bubble size as a correlating factor would not have been justified.

It is necessary, however, to have some idea of the typical bubble sizes which were present in the foam, in order to characterise and differentiate the action of generators, the basic reasons for the foam properties and flow behaviour. To this end a few spot values of foam bubble size, and size distribution, were made in order to give this information, and allow comparisons with other workers to be made.

5.1) Methods of Bubble Size Determination

The method used for most bubble size determination has been a photographic method. The foam or gas dispersion is photographed, the prints or negatives then analysed for bubble size by direct measurement of the bubble images. The resultant data can then be handled to give the required information about the foam. This process suffers from many draw-backs.

a) Foams are difficult subjects to photograph satisfactorily. The gas-liquid interface gives

rise to much light distortion, and clear and well-defined prints for analysis are difficult to obtain. The magnification needed in taking plots of very small bubbles introduces it's own problems of depth of focus.

b) The analysis of many prints is an extremely laborious manual task, requiring patience and skill if accurate and reproduceable results are to be obtained. Large numbers of bubbles have to be measured individually to determine the size distribution. The possible number of foams samples which can be produced, in even a small experiment run, means a check on bubble size at more than a token number of points is beyond the capacity of any such investigation as was involved in this work.

c) It is by no means clear that the bubble size observed at the wall, or even at the surface of any probe inserted into the flow stream, is a true representation of the bulk size distribution. Work on the freezing of foam, and sampling layers through the frozen bulk, has indicated that such an approximation can be expected to hold for a static foam (70), the case for a moving foam is not conclusive however. The wall will impart some preferential phase movement to bubbles near the wall, notwithstanding the inherent effects which must be present due to the non-symetry of the situation (72).

Information on the range of bubble sizes, which the sinter and pump generators produced, was obtained by using a photographic and counting technique. Additional work and investigation was carried out to develop a method of bubble size

estimation by methods which would avoid the use of any manual counting, and could be used to give readings of bubble sizes on a more readily available basis. The use of the excess pressure in the bubbles due to surface tension was investigated and found applicable.

Methods which would involve the automatic counting of the foam sample were also suggested, a system using the coulter counter principle is discussed. The adaptation of the bubble in tube measuring method is also discussed.

5.2) Bubble Sizing by photographic Method

Photographs of the flowing foam in the 17.3 mm pipe were taken, and prints prepared. Each particular flow condition was photographed several times, in succession. and the analysis of bubble size taken off a number of prints to:-

- a) obtain a sufficiently large sample to give a smoothed histogram
- b) reduce any effects of bubble size change with time off the generator
- c) determine the minimum sample required for good results.

Each print was analysed using an X-Y digital analyser, with paper tape data output. Each side of the bubbles on the prints were measured using the cursor, which was coupled to a digital voltmeter to give co-ordinate readings. The scale of the prints could be used to calibrate the electrical system, data being obtained directly in microns. The data tape was then reduced via a program to bubble diameters and number distribution.

The repeatability of individual bubble sizing off the prints was ± 2 microns, the magnification on the prints being such that bubble wall shadows were approximately twice as

thick as the cursor mark. The magnification on to the prints were:-

- a) 116 x lifescale for the pump generated foams
- b) 30 x lifescale for the sinter generated foams

5.2.1) Results for Sinter Generator

Bubble size information at two foam flow rates were obtained, such that $Q_A = 2 \cdot Q_W$. The resultant histograms can be seen in Figure 5.1, for the case of foam produced by the sinter, in an 0.5% Teepol solution, flowing in the 17.3 mm pipe. The plots have many similar features normal of foams produced from such sinters.

- a) no bubbles smaller than 100 microns are present
- b) only a small fraction, less than 0.05, are greater than 650 microns.
- c) both histograms have the same general shape, the distribution being even about the mean with a tail extending to higher bubble sizes.
- d) The mean value, $\bar{D}$ is only 7.5% greater than the medium, showing the standard deviation as being a good indicator of size distribution.

The data for each of these curves has been taken from three prints of each foam. The prints of the foam for the high gas rate can be seen in the Appendix in Figure E.1, and the individual histograms in Figure D.1. From these it can be seen that the sinter produces an even bubble size distribution, a sample of about 100 bubbles being sufficient to indicate the size distribution. The bubbles are evenly distributed about the mean and the generator shows no tendency to produce changes in bubble size at any fixed

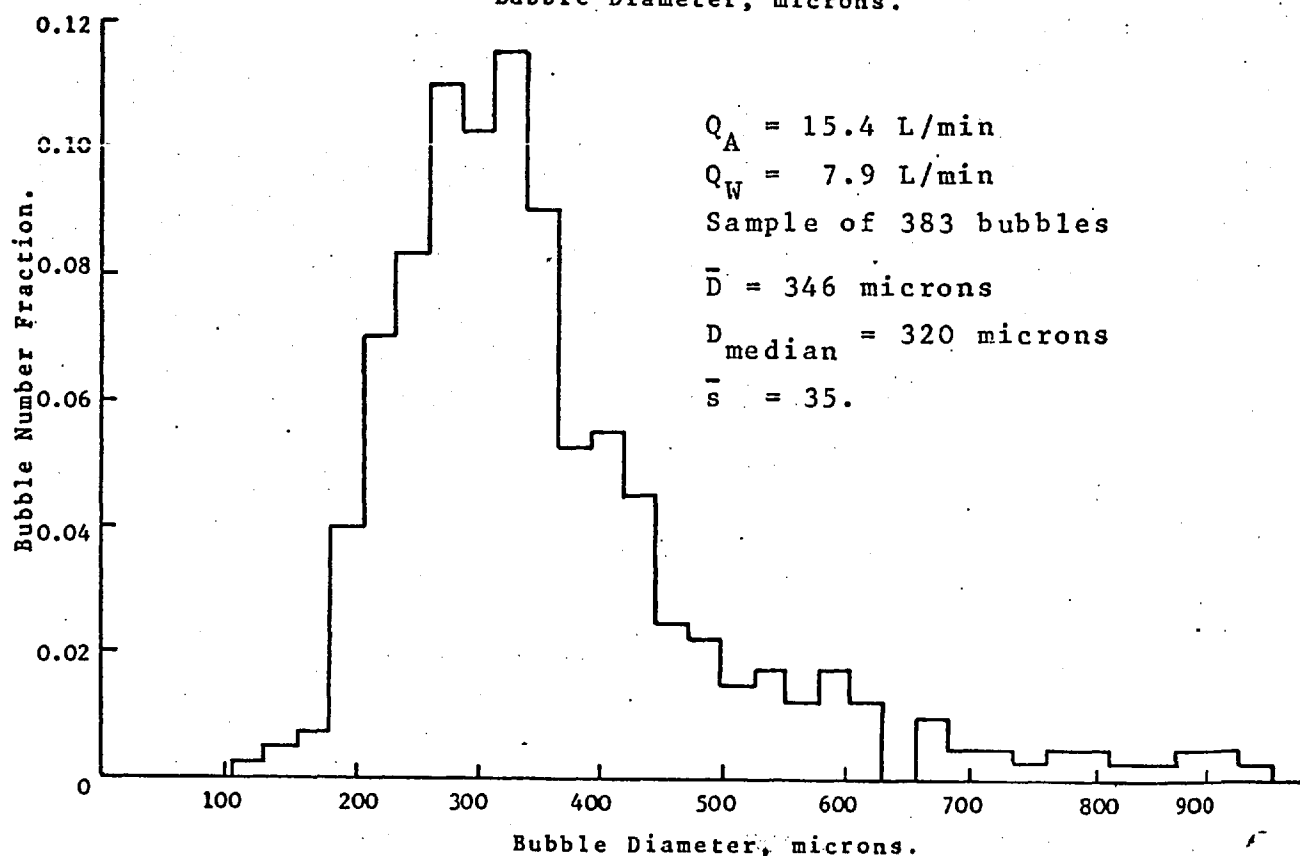
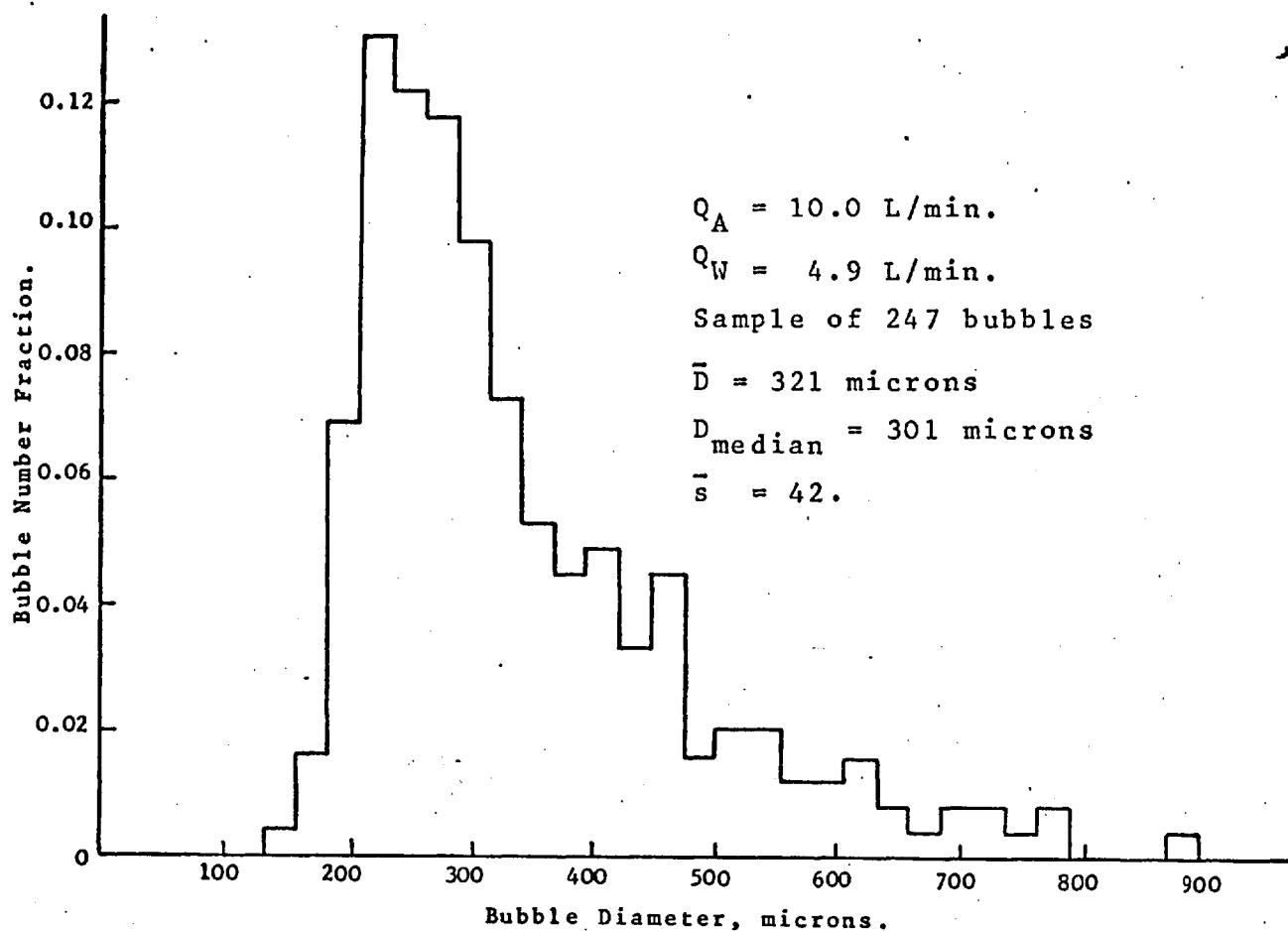


Figure 5.1 : Bubble size Histograms for Foam produced by the
 sinter generator in the 17.3 mm. pipe. $Q_A/Q_W = 2.0$

condition. This behaviour was as expected for the sinter generator.

A slight change in bubble size can be seen to occur as the gas rate increases. The number of pores on the sinter producing bubbles will depend on the pressure across the sinter, as the gas rate changes the pressure must increase and hence different pores come into operation. Results show that size and distribution do not change to any marked degree. A gross change in bubble size could be effected by

a) a change in liquid properties. A change of detergent concentration to 0.05% detergent solution results in a larger bubble size, as can be seen in Figure ES in the appendix.

b) a change in sinter material or pore size. The use of different porosity sinter discs could be used to alter bubble sizes.

The average bubble size for foam produced with the sinter generator can be considered to be within the range

$$300 < \bar{D} < 400 \text{ microns}$$

with a maximum size of 1,000 microns and a minimum of 100 microns, 80% of the bubbles will be within the range

$$200 < \bar{D} < 500 \text{ microns}$$

on an approximately bell shaped curve

5.2.2) Results for Pump Generator

The typical histograms of bubble size with frequency obtained for the case of the foam produced, in the 0.5% detergent solution, by the pump generator are shown in Figure 5.2. The figure shows the histograms of foams produced with a constant amount of water, the air rate being $\frac{1}{2} Q_w$ and Q_w respectively.

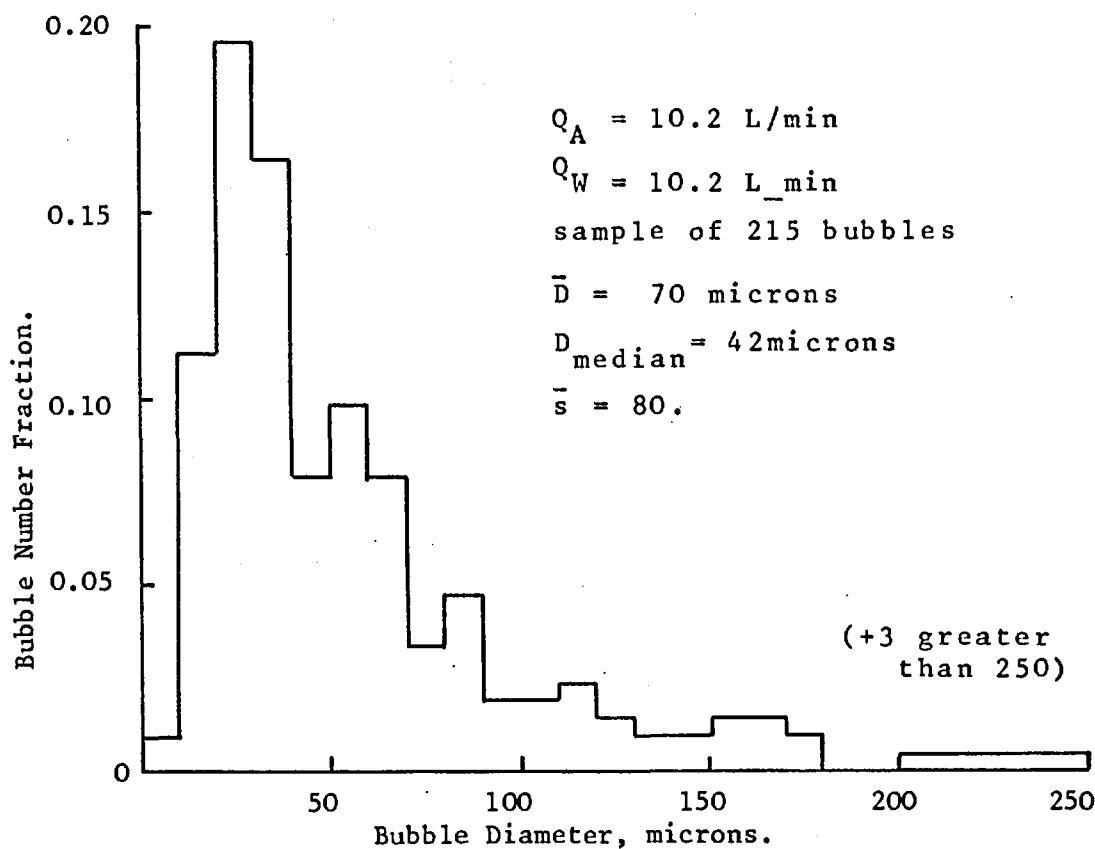
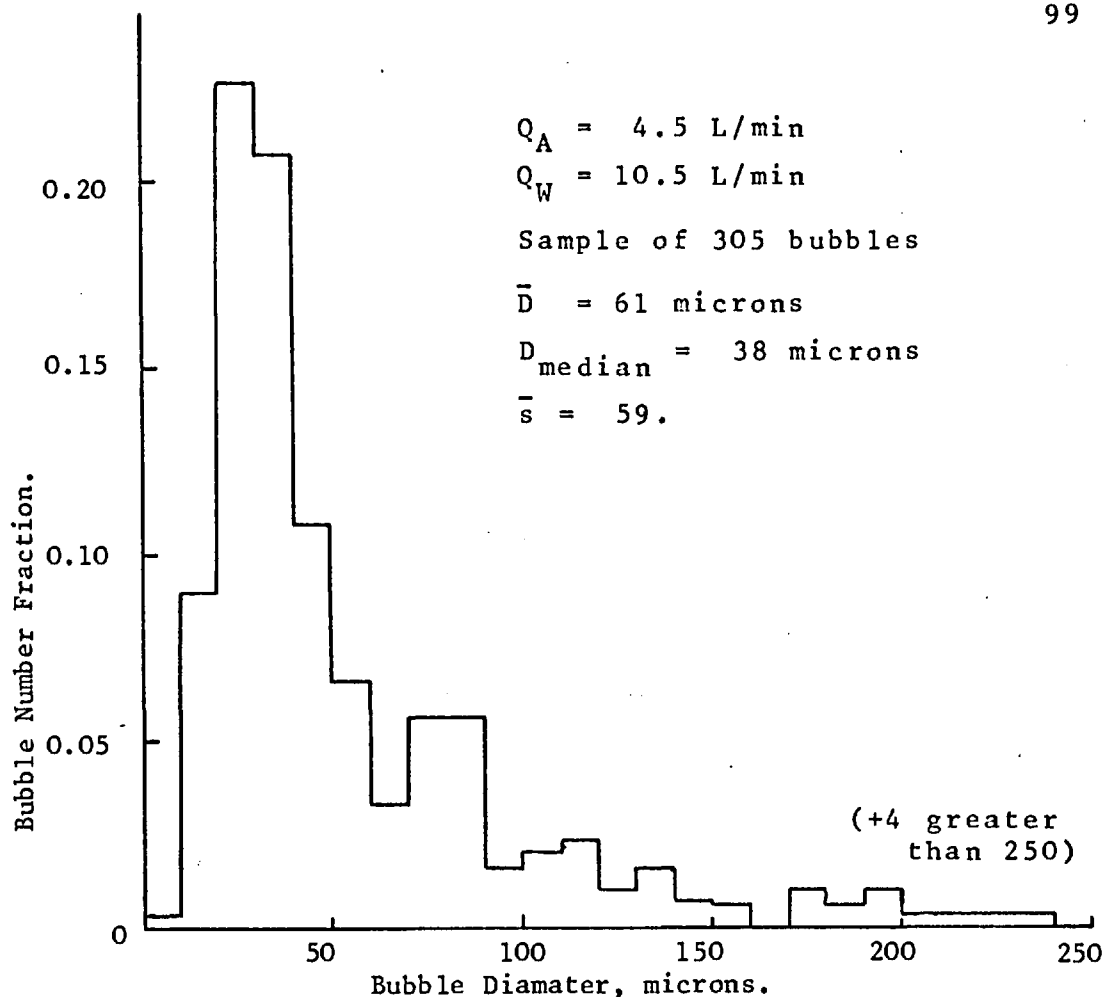


Figure 5.2 : Bubble size Histogram for Foam produced by the pump generator, in the 17.3 mm pipe.

The pump generator was fed with a mixture of air and liquid from the sinter generator, the bubble size in the feed being that produced by the sinter generator. Points of note of the histograms are:-

a) A few bubbles, $\simeq 1\%$, in each case were in the range 250 to 500 microns, these have not been shown on the plots.

b) the bubbles of diameter 10 microns and less were at the limits of measurement of this system.

c) The histogram show a marked skin with a long tail. This makes the use of standard deviation a poor indicator of spread of data.

d) In both cases $\simeq 65\%$ of the total bubbles are within the range 10 to 60 microns, while the mean values, $\bar{D}$, lie outside this range, 61 and 70 microns respectively.

The prints used to obtain the histogram when Q_W was 10.2 L/min are shown in the Appendix in Figure E.2. The individual histograms taken off each print are shown in Figure D.2. Inspection of these prints and histograms show that greater variation was obtained in the bubble distribution, for pump generation of foams, from each photograph, when compared to sinter generator results. A larger sample, about 200, of bubbles was required to make the contribution of the large bubbles obvious.

Inspection of the data at the two flow rates would suggest that there was no major change in bubble size with flow rate through the pump. The distributions have the same general shape, and the marked change in $\bar{D}$ is due to the effect of the few large bubbles present, a rather random selection. A reworking of the data, ignoring all bubbles greater than 250 microns in size gives

$$\bar{D}_1 = 56 \text{ microns and } \bar{D}_2 = 61 \text{ microns}$$

Significant changes in bubble size could occur if a different high intensity mixer was used. The effect of the degree of dispersion of feed to the pump was not found to be an important factor. Results for the pressure drop with different air feed arrangements were equivalent, the sinter air feed was used in all subsequent experiments.

The presence of the very small bubbles, $D < 25$ microns, could be alternatively observed in the drainage experiments. Rise velocities of these bubbles was less than the liquid drainage velocity, hence they would be washed out of the foam, and be unable to rise fast enough in the liquid to rejoin the foam, until the liquid drainage rate was sufficiently reduced.

5.2.3) Comparison of Foam Made with Sinter and Pump Generators

The size ranges of bubbles in the sinter and pump generated foams were such as to be nearly mutually exclusive. This can be clearly seen in Figure 5.3.

The difference in flow behavior of the foams can be attributed to the markedly differing bubble size range which are present in the foam. The very fine bubbles in the pump generated foam result in a pseudo-homogeneous fluid which has a half time of 0 [2½ minutes], and only flows in the dispersed regime. The sinter generated foam in contrast has a much larger bubble size, cannot be considered a homogeneous fluid, the half life being only 0 [½ minute], and experiences a wide range of flow regimes due to bubble coalescence and liquid drainage.

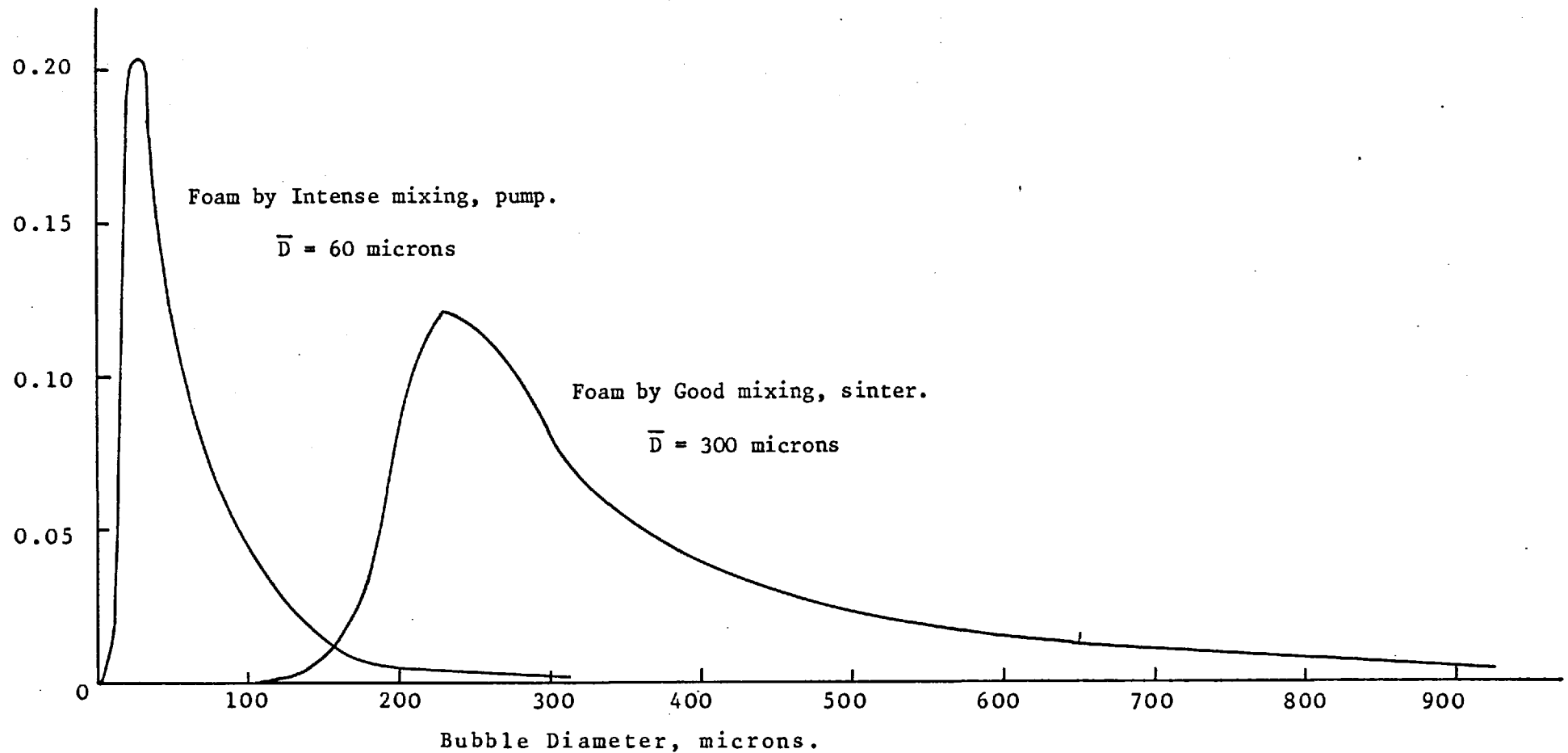


Figure 5.3: Comparison of bubble distributions for the foams generated with the pump, and sinter generators. Detergent solution = 0.5%. Data from photographs of foam flowing in 17.3 mmpipe.

5.3) Measurement of Excess Pressure in a Foam

5.3.1) Equipment and Data Collection

The excess pressure in a foam due to the surface tensions effects was measured using the small scale air-generation system, and the foam holding vessel as a constant volume system.

The foam sample was generated in the vessel, the taps at top and bottom closed and the pressure at the bottom of the sample measured at the start and when the foam collapsed by means of a water manometer. The readings taken were hence P_{B1} and P_{B2} the pressures being at the bottom of foam, and not equal to P_1 or P_2 .

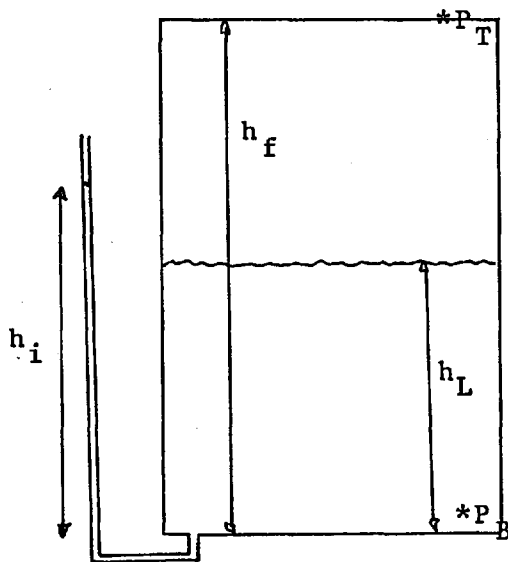


Figure 5.4: Foam pressure measurement vessel.

Let P_T be the pressure at the top of the vessel,

$$P_B = P_T + h_f \cdot \rho_f \cdot g \quad 5.1)$$

hence we can write,

$$(P_{B2} - P_{B1}) = (P_{T2} - P_{T1}) + (h_f \cdot \rho_f \cdot g) - (h_L \cdot \rho_L \cdot g) \quad 5.2)$$

but if a foam of uniform density is present in the sample,

$$(h_f \cdot \rho_f \cdot g) = (h_L \cdot \rho_L \cdot g) \quad 5.3)$$

and therefore,

$$\Delta P_B = \Delta P_T = (P_{T2} - P_{T1}) \quad 5.4)$$

Now P_{T2} , pressure at top of vessel when foam collapsed, is P_2 ,

$$\text{and } P_{T1} = \bar{P}_{\text{foam}} - (h_f \cdot \rho_f \cdot g)/2 \quad 5.5)$$

hence,

$$(P_2 - P_1) = P_{T2} - (P_{T1} + (h_f \cdot \rho_f \cdot g)/2) \quad 5.6)$$

$$\Delta P_{\text{foam}} = \Delta P_B - (h_f \cdot \rho_f \cdot g)/2 = \frac{2\sigma}{r} \quad 5.7)$$

hence from the difference in pressure, and the foam height of

the original sample, $\bar{r}$ can be calculated.

In the experiments carried out, $\Delta P_B = \Delta h_i \cdot \rho_l \cdot g$ and typical values of Δh_i obtained were 10 to 15 cms, while $\frac{h_L}{2}$ was 5 to 8 cm. The effect was measurable with the simple equipment used, and reasonable results obtained. The first reading of pressure must be made as soon as possible after the foam sample is stationary in the vessel as gas diffusion and drainage occurs, which change both $\bar{r}$ and the values of h_f and h_L . At a time $t > 0$ when the foam has drained.

$$P_B = P_T + (h_f \cdot \rho_f \cdot g)_t + (h_L \cdot \rho_L \cdot g)_t \quad 5.8)$$

where values
of h_f and h_L are
changing with t .

Hence it should also be possible to check the change in $\bar{r}$ with t if measurements of P_B and h_f and h_L are made, the value of ρ_f will however not be constant in the foam, an assumption of constancy only being possible at $t=0$ due to agitation of sample prior to start.

5.3.2) Results of Excess Pressure Measurements

An experiment was carried out to check the feasibility of using the excess pressure effect of the gas in a foam to obtain a measure of the average bubble size.

The experiment was carried out in the small scale air-foam generator, using a fine bore tube connected to the bottom to measure the pressure difference, see Figure 5.4. The vessel was filled to a depth of 18.2 cms with .5% Teepol solution, to give a ρ_f of .775 gm/cm³. A high density foam was employed for the tests so that good mixing and rapid generation would occur, the equipment producing a fine foam more rapidly at high density. The heights on the manometer were read at

the start of the time when the foam sample was trapped in the vessel, and 40 minutes later when all the foam had collapsed.

$$h_{t=0} = 32.0 \text{ cms}$$

$$h_L = 18.2 \text{ cms}$$

$$h_{t=40} = 47.0 \text{ cms}$$

The volume of gas at the end, $V_2 = 213.5 \text{ cm}^3$, and the change in total volume due to liquid being forced up the manometer tube = $.265 \text{ cm}^3$. The value of the surface tension, 30 dyne/cm., was taken from Figure A1..

Hence from equation 5.7 the average radius can be found:-

$$(47.0 - 32.0) \rho_L \cdot g - \frac{18.2}{2} \rho_L \cdot g = \frac{2 \times 30}{\bar{r}}$$

$$\bar{r} = \frac{2 \times 30}{(15. - 9.1) \cdot \rho_L \cdot g} = .0102 \text{ cms}$$

$$\text{hence } \bar{r} \approx 100 \text{ microns}$$

This value of $\bar{r}$ is large when compared to the values obtained for similar foams using a photographic technique, especially as it is obvious that a fair proportion of the bubbles have a radius of less than 15 microns.

5.3.3) Corrections

Corrections have to be made to the value of P_B . to account for certain inaccuracies in the equipment:-

a) the gas is not held at constant volume, a slight change in volume of the system occurs due to change in height of manometer pressure measuring system

$$V_2 = 213.5 \text{ cms}^3 ; V_1 = V_2 - (\text{Volume of liquid forced into manometer})$$

$$V_1 = 213.5 - .265 \text{ cm}^3$$

$$\therefore \% \text{ change} = \frac{.265}{213.5} = \underline{0.13 \%}$$

b) The vessel with the decaying foam might not be at a constant temperature, no attempt is made to control temperature, which could change by a degree or two during the 40 minutes that the foam takes to collapse. Ross (74) gives typical figure of about 4°C for the change in gas temperature due to energy release as the surface area of the foam decreases to zero, but this amount of heat would be absorbed by the liquid which has a very large heat capacity. The most likely temperature changes are due to natural exchange with the air in the laboratory, as heat is produced by the pump it might be expected to cool down as this heat is dissipated.

Hence equation 2.79 should be more exactly,

$$P_1 \cdot V_1 + \frac{2\sigma A_1}{3} = nR \cdot T_1 \quad 5.9a)$$

$$\text{and } P_2 \cdot V_2 = nR \cdot T_2 \quad 5.9b)$$

$$\therefore (P_2 - P_1) - \frac{2\sigma A_1}{3 V_1} = \frac{nR \cdot T_2}{V_2} - \frac{nR \cdot T_1}{V_1} \quad 5.10)$$

$$(P_2 - P_1) - \frac{2\sigma}{\bar{r}_1} = \frac{nR \cdot T_2}{V_2} \left(-1 + \frac{T_1 V_2}{T_2 V_1} \right) \quad 5.11)$$

sub for $\frac{nR \cdot T_2}{V_2}$ from 5.9b and rearrange

$$(P_2 - P_1) + P_2 \left(-1 + \frac{T_1 \cdot V_2}{T_2 \cdot V_1} \right) = \frac{2\sigma}{\bar{r}_1} \quad 5.12)$$

hence a correction of $P_2 \left(-1 + \frac{T_1}{T_2} \cdot \frac{V_2}{V_1} \right)$ must be made to account for the non uniform volume and T.

The value of $(P_2 - P_1)$ was measured in the experiment as $(h_2 - h_1)$ and was $\approx 6 \times 10^{-3}$ bar. For the correction factor, using the volume change, and an assumed change in temperature of 1°C, $T_1 > T_2$, the result is:-

$$P_2 \left(1 - \frac{294}{293} \cdot 1.0013 \right) = 0.9953 P_2$$

but $P_2 \approx P_{\text{atm}}$, $P_{\text{correction}} \approx 0.0047 \text{ bar}$.

which gives

$$\bar{r}_c \approx 56 \text{ microns}$$

This can be seen to be a very large change in estimated radius, and compares ^{better} with that obtained by the photographic method. The excess pressure in the foam was only $\approx 10^{-2} \text{ bar}$, for a foam at atmospheric pressure, this is $\approx 1\%$ of total pressure, hence both the volume and temperature control must be better than 0.1% if errors of approximately 50% are not to be made.

5.3.4 Conclusions

The measurement of the excess pressure in a foam due to the surface tension effects can be used to measure the bubble size of fine bubbled foams. The equipment used to test the theory was adequate to indicate the suitability of the results, and the results for bubble radius obtained were of the same order as those obtained by photographic.

Modifications to the equipment to obtain accurate results should aim to:-

- a) Use a more sophisticated pressure measuring system so that no volume change need occur to measure the pressure difference at start and end.
- b) Some precise temperature control or monitoring system is necessary, either to keep a constant temperature, or to be able to specify temperature changes. Correction for expansion of vessel and liquid will also be needed if these are to be used in cancelling out errors of type a).
- c) Facilities to make photographic checks on the bubble sizes, to check on the accuracy of the method.

5.4) Bubble Counting Methods

Alternative methods of obtaining bubble size and distribution information were also investigated. The use of overall sampling such as the excess pressure technique, does not offer any size distribution information about the foam, a serious defect as the presence of only a few large bubbles can greatly alter the value of the mean bubble diameter. The photographic technique is extremely slow and tedious, involving the production of prints and manual data collections. Methods which could be employed on a semi-automatic basis, without direct operator intervention to collect the data are suggested. The general feasibility of the methods have been studied for practical use, but have not been experimentally verified.

5.4.1) Bubble Sizing via Coulter Counter

The use of a Coulter counter to give a measure of bubble size and distribution is suggested. The Coulter counter operates on the change in electrical characteristics across a small orifice as a particle passes through the orifice. The signal change as the particle goes through the orifice gives a measure of the size. Calibration of the system, and careful choice of liquids is needed for reliable operation. With a sufficient number of particles passing through the orifice, the output can be analysed to give a histogram of bubble sizes. Sample measurement time is of the order of 10 seconds.

Standard Coulter counters are only devised to be able to count the bubble number in a particular size range, the calibration being altered via electrical means for other size ranges. This means that any sample has to be measured

as many times as the number of size ranges in the required histogram. This for a foam sample would be inconvenient. Alternative equipment which will produce a histogram from a single cycle of sampling are available at additional expense.

Problems of sampling a foam will arise due to the number of bubbles being so great that more than one will pass through the orifice at once. The analyser is designed to filter out non-perfect signals designed by such events, but threshold limits mean some dual particles will be treated as larger single particles. Foams investigated have bubbles which touch one another, and such samples could not be sampled. Dilution of the sample, a standard technique for use of the counter, so that individual bubbles would pass through the orifice, is required. For foams, of bubble size $20 \rightarrow 200$ microns, a dilution of 10^3 is necessary. Carefully controlled dilution would be necessary to avoid size changes, but sample measurement could be effected almost instantly using the spectrum analyser.

The use of this system would result in foam bubbles size measurements literally on the spot, which would be of great use in equipment optimisation or factor change studies.

5.4.2) Capillary Tube Counter

A technique for measuring bubble size has been developed which consists of trapping the bubbles in a fine capillary, such that the capillary diameter is less than bubble diameter

. The bubble is elongated in the capillary forming a cylindrical bubble, which can be measured. From the length of this bubble the size of the spherical bubble can be calculated, taking account of any correction necessary.

The minimum size of bubble which can be counted and measured in this way is equal to that of the capillary. Provision of small diameter capillaries of uniform bore is difficult and would be an important part of the successful operation of the equipment.

The length of bubble in the capillary can be measured by eye, using a Vernier microscope. Alternatively if a constant flow rate is maintained in the capillary the rate at which bubble walls pass a fixed point could be measured with an optical system coupled to an analyser to give automatic readings. Bubbles which touch one another could be measured as the bubble wall could be used as a triggering device. This would mean samples directly from the bulk foam could be sampled via a probe, extracting the foam through the capillary at a known rate.

Should dilution be required to separate the bubbles, so they were distinct in the capillary, this could be done on a continuous basis.

This method of bubble measurement could result in a great flexibility of foam sampling. Foam near to, and far from, walls could be sampled to investigate stratification effects. The provision of constant bore capillaries would be unnecessary if measurement of the moving bubble is used, the bore diameter at the measurement point only being required.

5.5) Conclusions.

Bubble sizes of foams are difficult to make accurately, and are often inconclusive once obtained. Typical values for the foams used in this work were obtained by photographic means for comparative use, no attempt being made to collect

extensive bubble size data.

Alternative methods were investigated and discussed, with the view to simplifying the sizing procedure. Use of the excess pressure in the foam shows potential for further investigation, although the technique does not give size distribution information. Automatic size counting techniques, with sophisticated equipment requirements, was also discussed and initial considerations are promising.

6. Flow Pattern Maps for Detergent Solutions.

The flow regimes which would exist in a horizontal pipe with the simultaneous flow of air and detergent solution were investigated. The effects of pipe size, detergent concentration and phase mixing methods on the flow regimes and the transition boundaries were investigated, and results were obtained for a representative sample of conditions.

6.1) Flow Regime Descriptions.

Any study of flow regime pattern maps needs to be based upon a universally agreed nomenclature, if meaningful comparisons between the maps of different workers and the flow systems they have investigated, are to be made. In the system under consideration the presence of the detergent gives rise to some marked changes in the typical flow regimes normally observed.

Part of the work involved the investigation of pure liquid/gas flows for comparative purposes, and the nomenclature used for these conditions, stratified, elongated bubble, slug and dispersed flow, follow the accepted definitions for these regimes, full descriptions of which can be found in the literature (28).

In the flow pattern maps which were developed for the detergent solution/gas flow system, the phase velocities were such as to generate four major flow regimes: dispersed bubble, stratified foam, slug and foam flow regimes. These regimes will be defined in detail below. Superficial air velocities of sufficient magnitude to be within the annular flow regime were not attained in this work.

6.1.1) Dispersed Bubble Flow

At high superficial water velocities the gas bubbles are

dispersed throughout the liquid, with a higher concentration of bubbles in the upper part of the pipe. As the gas rate increases the bubble size remains approximately constant and the bubble density increases, until the flow becomes opaque.

6.1.2) Stratified Foam Flow.

The gas and detergent solution flow in stratified flow in the pipe, with the gas phase forming the upper foam layer and the bulk of the liquid the lower layer. The bubbles formed in the generating device fill the upper part of the pipe so that the bubbles are closely packed, forming a relatively dry foam, which travels down the pipe with little bubble coalescence or breakdown. The relative movement between the bubbles is low, the upper foam layer tending to move as a plug, especially at low velocities. The lower part of the pipe is filled with the remaining liquid, part of the liquid being included in the foam. The lower layer is essentially clear of bubbles, but as the phase velocities increase so does the quantity of bubbles in this liquid layer. The liquid layer flows faster than the upper foam layer, with a velocity which can be many times greater than the foam flow velocity.

The interface between the foam and liquid is well defined and can be either flat or wavy. At low superficial water velocities the interface is flat, although not totally smooth, as the bubble size tends to give it a certain inherent "roughness", and only a few bubbles are being carried along in the lower water layer. These bubbles tend to travel along near the interface, but cannot leave the liquid and enter the foam, as the lower foam surface seems to have a certain rigidity which stops these few bubbles joining the main foam bulk. The bubbles on the interface are not moving with relation to

one another. As the water rate increases the interface becomes progressively more wavy, in a well defined manner with regularly spaced waves. The waves are initially ⁸nearly static with regard to the foam, moving forward in the foam only slowly. More bubbles are present in the liquid layer, a function of the generation process, as the interface is not shedding bubbles. The waves continue to grow in both magnitude and relative velocity to the foam, but change their wave length only slightly, until the flow becomes broken up and the interface "churns" and dispersed flow occurs.

These effects on the interface behaviour are most noticeable in the largest pipe (39 mm) and were not apparent in the smallest pipe (11 mm).

6.1.3) Foam Flow

The pipe is completely filled with a foam, in which the bubbles are closely packed and relative bubble motion is low. There is a thin water layer against the walls of the pipe. The bubble shape depends on the amounts of gas and liquid phases present; at the low gas/liquid ratios the bubbles are spherical, however once the close packing ratio has been reached the bubbles can start to become polyhedral. The bubbles next to the wall move together, and especially at high gas to liquid ratios it is apparent that the interior bubble layers are slipping past one another with progressively higher velocities the further into the foam the layer is from the wall

6.1.4) Slug Flow

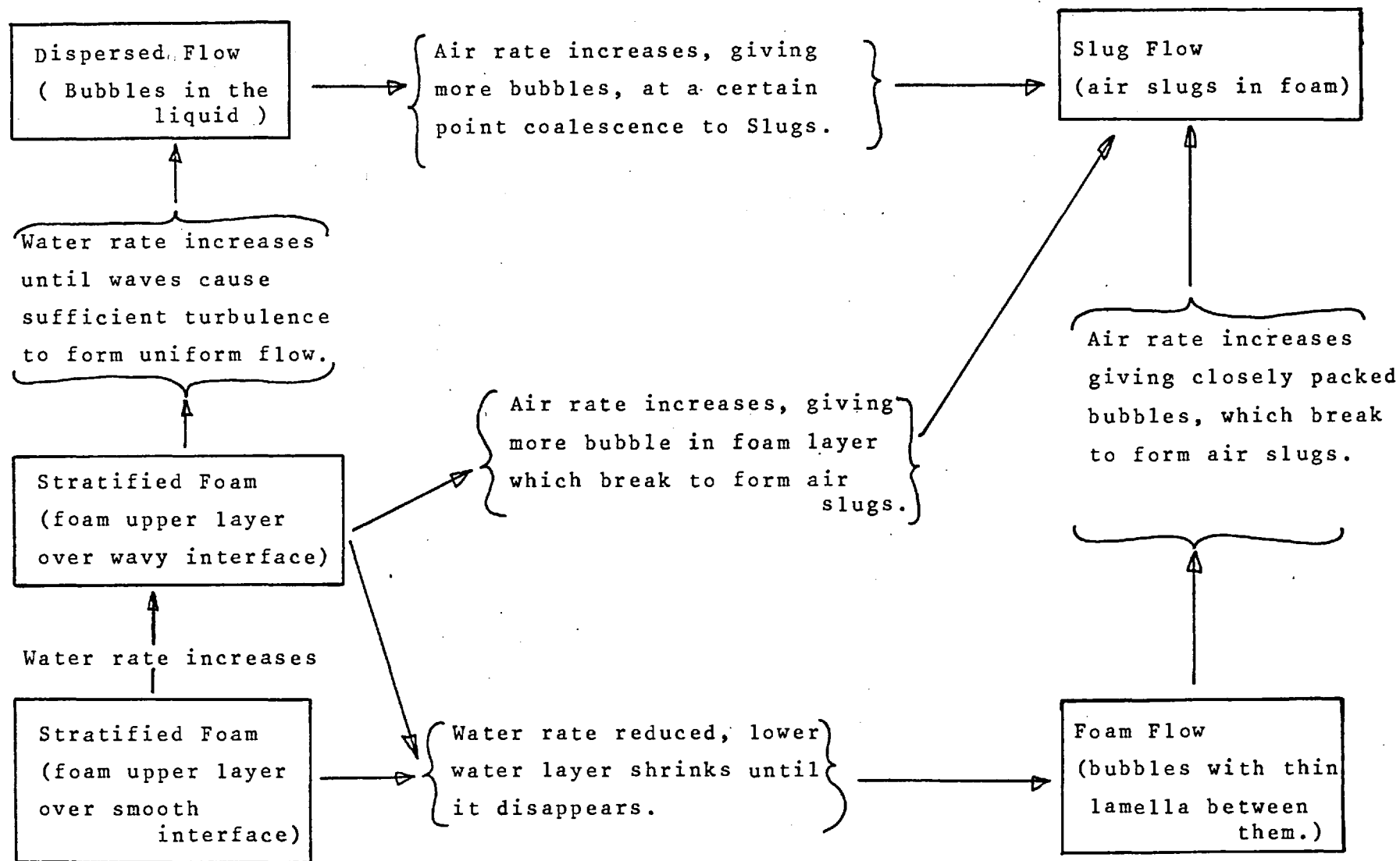
The slug flow regime consists of foam flowing in the pipe, with slugs of air flowing faster than the foam, passing along the pipe, in a manner analogous to that of gas-liquid slug flow. The air slugs are approximately the same diameter as the pipe and can be of varying length, from the diameter of the pipe upwards, as the gas flow rate increases. The slugs create much turbulence as they pass through the foam, a layer of foam being left on the pipe walls as each slug passes, and then swept up by the following plug of foam.

The slugs are not formed by the action of the generating device, but from some break-down mechanism of the foam, since uniform flow in a larger pipe will become slug flow in a subsequent smaller pipe coupled in series, after about 50 pipe diameters of flow in the pipe.

6.1.5) Regime Transition Mechanism

The flow patterns which result from the flow of air and detergent solution, could to a certain extent be considered to be those of three phases; i.e., gas, liquid and foam phases. The relative amounts of the gas and liquid phases will create in the pipe, either a mixture of gas-foam or foam-liquid or even gas-foam-liquid depending on flow rate, pipe size and detergent concentration. The changes from one regime to another will happen as the phase velocities are changed, eg., stratified foam flow will change to foam flow as water rate is reduced, the lower stratified water layer being slowly depleted until the upper layer descends and finally fills the pipe. The relationship between the various regimes, with the main factors involved in the transition is shown in Figure 6.1.

Figure 6.1: Flow regimes and Transition Mechanisms.



6.2) Flow pattern Photographs.

Photographs of the pipes were taken to aid in the regime definition and flow description. They were not used to establish transition boundary values, this being done by the visual estimation of the flow pattern in the pipe.

Photographs of the foam produced with the sinter and pump generators can be seen in the Appendix E, Figures E.1 and E.2. Photographs of the complete pipe to show flow regimes are given in Figure E.3 to E.5, and consist of,

a) Figure E.3; 26.0 mm pipe, at a constant air rate with water rate increasing. This shows the Transition from smooth to wavy stratified foam flow, the regular form of the interface can be clearly seen.

b) Figure E.4; 26.0 mm pipe, at a constant water rate with air rate increasing. This shows the transition from smooth stratified foam flow to foam flow as the Q_A/Q_W increases.

c) Figure E.5; 17.3 mm pipe, at a constant air rate with water rate increasing, at 0.05% and 0.5% Teepol concentrations. This again shows the smooth to wavy stratified transition, and also the dispersed regime at the maximum water rate. The effect on the bubble size due to the detergent concentration can be clearly seen.

No regime photographs of the slugging regime could be obtained, the photographic technique not showing this regime.

These few examples of flow regime indicate the difficulties involved in regime estimation. The static flow was often more difficult to place, as the flow patterns are essentially in a dynamic state, with behaviour more obvious to the naked eye. The dividing line between dispersed and stratified flows can be taken as an example of how difficult the definition of the transition point is to make.

6.3) Method of Regime Identification

The flow regime identification was carried out in all experimental work by purely visual means during the course of the runs. The pipe was well lit to aid in flow regime estimation, and the flow regime which existed in the pipe was noted along with the relevant phase flow rates and pressure drop data. The full range of each set of conditions was covered quickly without recording data to establish the likely flow rates and regimes and help define regimes. All the data for each particular set of conditions was collected before any attempt was made to plot the results; this was done to reduce the likelihood of regime mis-identification as a result of making the data fit the 'expected' pattern map. The data for all pattern maps, with very few minor exceptions, was collected over a comparatively short space of time, hence the regime descriptions and transition states can be considered to be internally self-consistent. This visual technique is by no means entirely satisfactory, and several workers (28,51) have stressed the need for additional aids, either via 'still' or cine photography, to allow subsequent checking and re-defining to be carried out, or via pressure drop or other parameter correlating techniques.

6.3.1) Stratified Foam to Dispersed Bubble Regime

The transition of stratified foam to dispersed bubble regime was the most difficult of the changes to define accurately. The velocity in the pipe is high, about 5 m/s, and the point at which turbulence of the lower water layer is sufficient to give an adequate 'uniformity' to the flow is a matter of arbitrary definition. Few points were

recorded in the dispersed bubble region, due to pressure limitation of the equipment. This is however not uncommon in many similar works on flow pattern maps.

6.3.2) Stratified Foam to Foam Flow Regime

The stratified foam to foam flow regime transition was more easily defined. The velocity of mixture in the pipe is lower than that for transition to dispersed flow with a maximum of about 2.5 m/s, but was normally much less. The presence of the lower liquid layer in the pipe was always easily noted, in part due to it's greater velocity and the few bubbles in the layer, and this layer's disappearance was found to be a fair indication of regime change.

6.3.3) Wavy or Smooth Stratified Foam Flow

Different states of the interface during stratified foam flow were noted as either wavy, smooth or undefined, but no further accurate measurements of the structure were taken. Change was most noticeable in the large pipe, and was not recorded in the smallest pipe, a fact due mainly to the range of phase velocities covered in the pipes.

6.3.4) Dispersed Flow to Foam Flow

This regime change was met only in the case of the 11 mm pipe with detergent solution. The boundary has not been defined on the plots, but a symbol change on the data plots has been arbitrarily used, to occur at a water velocity of 1 m/s.

6.3.5) Transition to Slug Flow Regime

The transition to and from the slug flow regime from the dispersed, stratified or foam flow regimes was found to be

very well defined for the detergent solutions. The air slugs formed when slugging first starts to occur nearly fill the pipe, and so cause a noticeable change in the flow pattern, unlike the pure liquid/gas system where the difference between a large elongated bubble and a slug can give rise to many differing interpretations. This transition manifests itself in the foam as a slight flicker as the slugs move rapidly down the pipe. As the gas rate increases the flicker tends to continue at approximately the same rate, but becomes more defined as the slugs become larger. The repeatability of the slug transition boundaries using this flicker criteria was found to be ± 0.2 m/s on the gas superficial velocity, a much clearer definition than is obtainable in similar runs with pure liquid-gas, in which the slugging transition point for a large elongated bubble to a slug can be open to a fair degree of uncertainty and personal interpretation.

6.3.6) Use of Photographs

A few photographs were taken of the pipe to check on regime description and estimation. A typical selection can be seen in Figure E.3 and E.4 for flows in the 26.0 mm pipe at constant air and water rates, and in Figure E.5 for flow in the 17.3 mm pipe at 0.05% and 0.5% Teepol concentration. These Figures were not used in the construction of the flow pattern maps, but can be seen to confirm the flow pattern descriptions and approximate boundaries. The difficulties involved in the visual identification of the various transition boundaries can be seen from these photographs, any criteria of transition is to a certain extent dependent on the precise definition of the flow patterns, in itself a rather arbitrary procedure.

The desirability of the use of photographs for regime identification is to a certain extent limited by the scale of data collection necessary for flow pattern map preparation, and in this case was not carried out to any large extent due to the fortuitiously clear definition of the slug regime boundary.

6.4) Range and Scope of Experiments

The parameters which were varied in the experimental work were:-

- 1) The pipe sizes in which the experiments were carried out
- 2) The gas and liquid velocities
- 3) The method of introduction of the gas and liquid to the pipe
- 4) The concentration of the detergent in the liquid, ie., changes in surface tension and foaming characteristics of the liquid phase

All experimental work was done using a test length of six metres, not including any pipework necessary for the gas/liquid mixing system. The mixing system used was coupled to the test pipework, with reduction or expansion pieces as required. No other changes were made to the mixing system to account for the varying gas and liquid requirements of the various pipe sizes. This tended to limit the maximum flow velocities in the larger pipe sizes.

6.4.1) Pipe Size and Type

The pipe sizes used in the experimental runs were all glass pipes and were:-

- 1) 11.4 mm diameter
- 2) 17.3 mm diameter
- 3) 26.0 mm diameter
- 4) 39.1 mm diameter

All except the 11.4 mm pipe were the standard QVF glass piping with the normal butt joints, the 11.4 was made up from glass tubing and flanges. See section 3.1 for pipe work and other details. The sections under test were six metres long, made up of several lengths of pipe.

6.4.2) Phase velocities used in the experiments

The maximum flow rate that the equipment was designed to give was 60 litres/min. The range of superficial velocities possible in the various pipes is shown in figure 6.2. The maximum value of water and air flow rates was not reached in the case of the 11 mm pipe due to pressure drop limitations. From the figure it can be seen that not all regions of interest could be covered with the equipment. The flow pattern map for air-water is included in the figure to give an indication of regimes covered.

The lower superficial velocities are set by the minimum controlable and accurately measurable phase rates; 1 Litre/min.

6.4.3) Mixing Method

The flow patterns that might be obtained could be expected to depend on the mode of gas/liquid mixing, as the interaction between the phases will depend on the degree of initial contact between the gas and liquid. The methods of injection gave three ranges of gas/liquid interaction, primarily a function of the bubble size formed in the generation equipment.

- 1) Very low mixing, a simple T-piece in which gas and liquid were introduced as an upper and lower layer with no bubbles being formed.
- 2) Good mixing, the gas was blown into the liquid via a sinter disc, bubble sizes being in the range 0.1 mm to 1 mm.

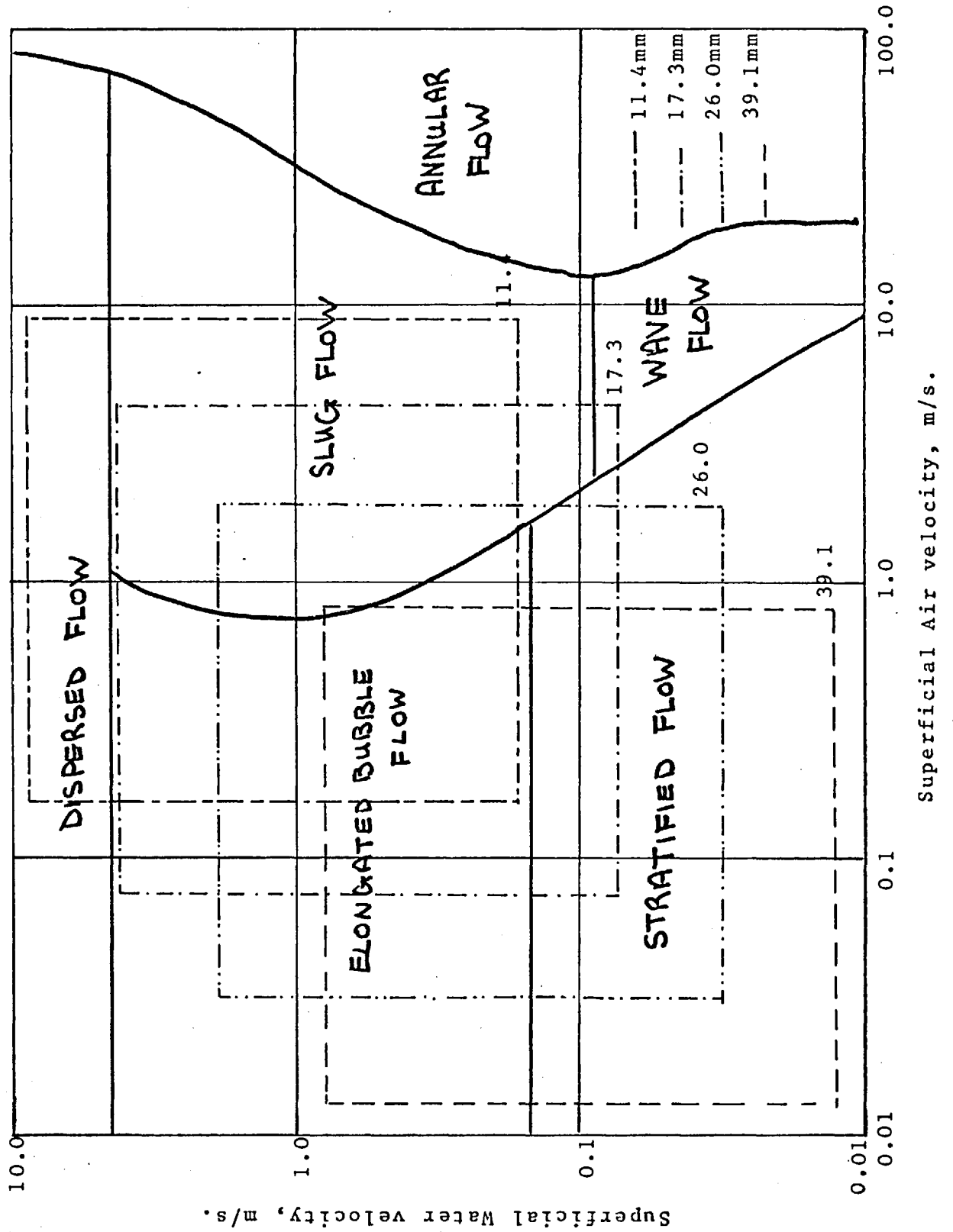


Figure 6.2: Phase velocities reached in the test pipes, with the Flow Pattern map of Mandhane et al(51), for comparison purposes in the Foam flow maps discussion.

- 3) Intense mixing, the gas/liquid stream was passed through a mixing valve and centrifugal pump, bubble sizes being in the range 20 - 200 microns.

Typical bubble size distributions for method 2 and 3 can be seen in figure 5.3 in the discussion on bubble size determination.

In all cases the flow pattern at any particular set of phase velocities and injection method was established after a short entrance length in the test pipe, typically 50 diameters. This has been found to be the case generally in two phase flow systems.

6.4.4) Detergent Levels

Four levels of the foaming agent, Teepol L, were investigated to find the effect on the flow patterns of varying surface tension and ease of bubble formation. All solutions were made up using mains water, the concentration being adjusted by addition of the required volume of detergent, the surface tension being measured by Ring Tensometer.

- 1) Zero - pure water (see section 3.1 for discussion on preparation for pure water experiments).
- 2) Trace of detergent - 0.002% by volume- surface tension = 59 dynes/cm. The solution would form a slight foam of only a few seconds persistence.
- 3) Below the C.M.C. - 0.05% by volume- surface tension = 36 dynes/cm. A stable long lasting foam formed.
- 4) Above C.M.C. - 0.5% by volume- surface tension = 31 dynes/cm. A stable long lasting foam formed.

See Appendix A for full details on behaviour of surface tension with concentration of Teepol L.

6.5) Treatment of Data

The experiments for the flow pattern maps were carried out in conjunction with pressure drop estimation in the test pipes, and the mode of control of the phase velocities was aimed mainly at the systematic collection of pressure drop data.

The tests were carried out by fixing a water flow rate, and then varying the gas rate in approximately constant steps. Control of the gas rate to specific values was found to be impracticable, as the in-situ gas rate can only be calculated after the average pressure in the pipe was found. As no control of pipe mid-point pressure was carried out, the in-situ pressure varied as the flow rates changed. The complete cycle of runs at the required water rates were carried out, and then additional points taken, if required, to fill in any main areas of uncertainty. Hence, of the range of values obtained, only about 1/3 of the points are near, or on, the pattern transition boundaries.

The flow patterns observed were categorised into two major groups.

- a) a definite flow pattern of one of the given types
- b) some intermediate or indefinite pattern, not clearly of either type of flow pattern.

The data was then plotted on to log/log graphs, with a series of symbols to denote the various flow patterns, and the intermediate cases. The gas velocities were calculated from a knowledge of the pressure in the pipe, and the standard flow rate, read off the rotameters. The full charts with the data plotted are given in the Appendix, Figures F.1 to F.10. The tabulated values used to plot these charts are given in

Tables I.1 to 14 ; for the various flow conditions investigated.

6.5.1) Example Flow Pattern Map

An example plot is shown in Figure 6.3, and on this the transition lines have been superimposed. The symbols used are more fully explained in the Appendix.

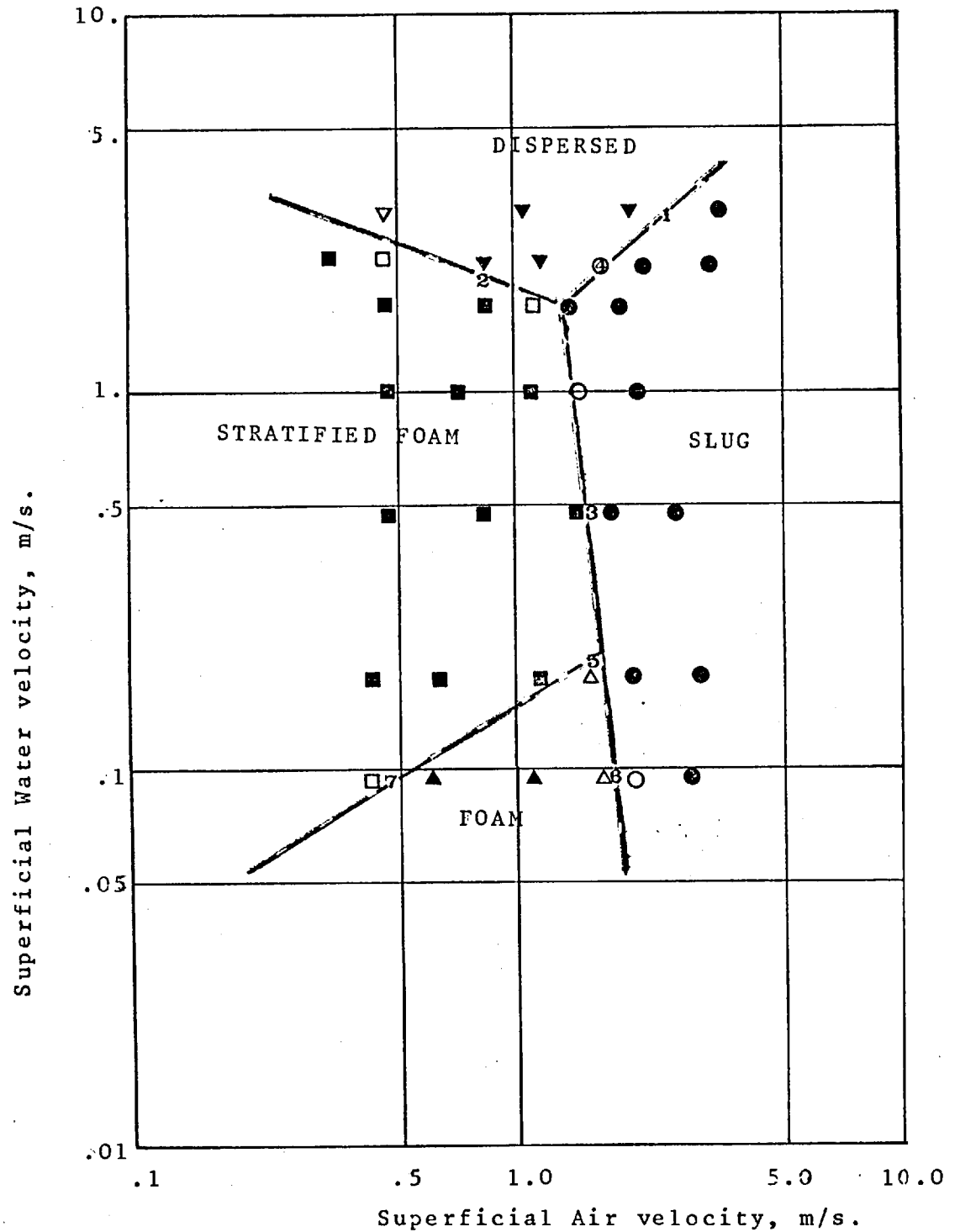
The transition lines have been plotted by taking the mid points between the definite flow pattern points (points 1,2 & 3) and the indeterminate points (points 4,5,6 & 7) and joining with the best straight lines. No attempt was made in boundary generation to include every regime point within the specified regime region, if the point seemed to fall outside the general area of the rest of the points. Where practical the lines are shown as curves if sufficient points are available to define the line, otherwise the boundary is given as a straight line.

The pressure drop data has on occasion been used to help in estimation of the transition points, as Pressure drop versus flow rate plots can sometimes show sharp changes. See Section 7 for details.

The boundaries are shown as lines on the plots, in fact they themselves are regions, maybe of quite wide and varying value, over which the regime changes take place. A fair degree of latitude in the interpretation of regime map data is necessary, if meaningful results are to be obtained from the use of such simple parameters as phase velocities for the description of the extremely complex flow situations actually existing in the pipe.

6.6) Results of Flow Pattern Experiments

The results from the range of experiments, with various detergent concentrations and pipe sizes can be most usefully looked at in relation to the detergent concentration, and



Flow Pattern

Definite

Indeterminate



Disperse



Slug



Stratified



Foam



Figure 6.3: Typical Regime Pattern map, with construction of transition boundaries from data.

phase mixing methods. Full tabulated data and the plots used to obtain the regime boundaries are given in the Appendix

For comparison and discussion purposes the flow pattern map of Mandhane, Gregory and Aziz (51) has been used, and is shown in Figure 6.2, with a change of axes values from ft/sec to m/sec. The regime boundaries off this map are reproduced on many of the following experimental charts, to aid in the understanding of how the action of the detergent has changed the flow regimes. The regimes names on the various plots are those actually observed in the pipe, see Section 6.1 for full description of regimes.

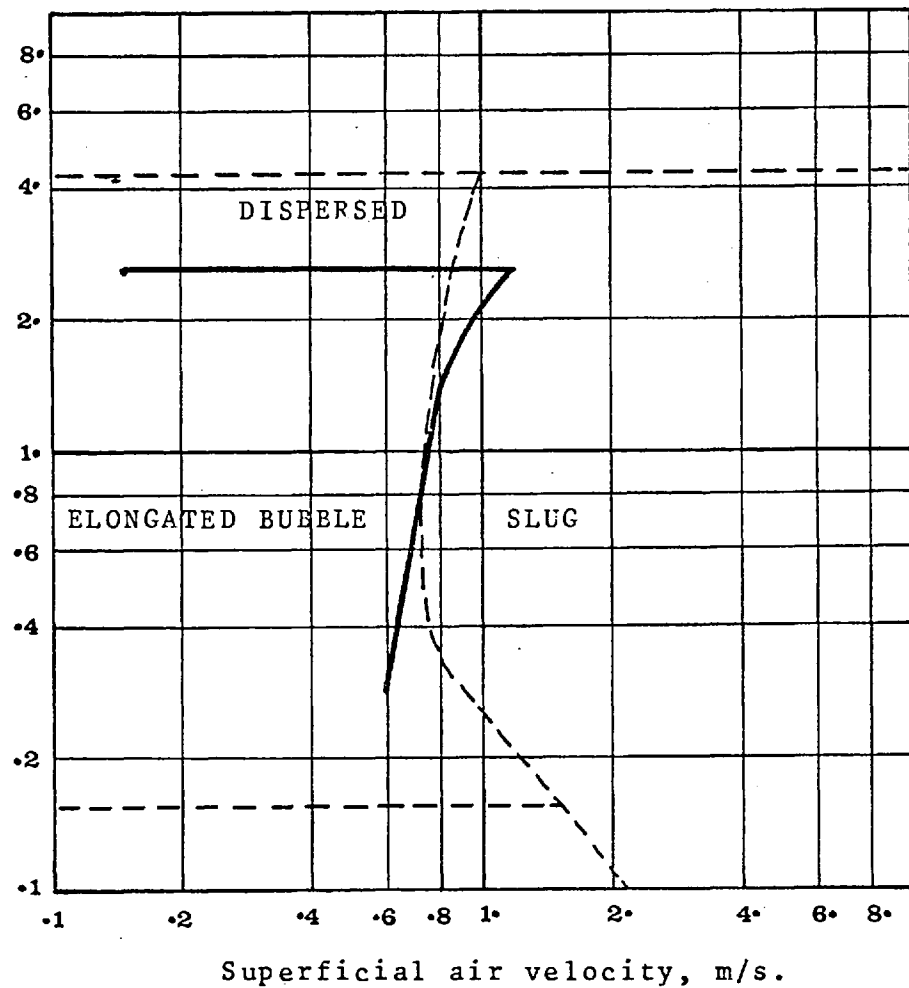
6.6.1 Flow Pattern for Water-Air

The flow pattern maps for pure water-air flow were constructed as a check on the equipment and for comparison purposes with the detergent flow pattern maps.

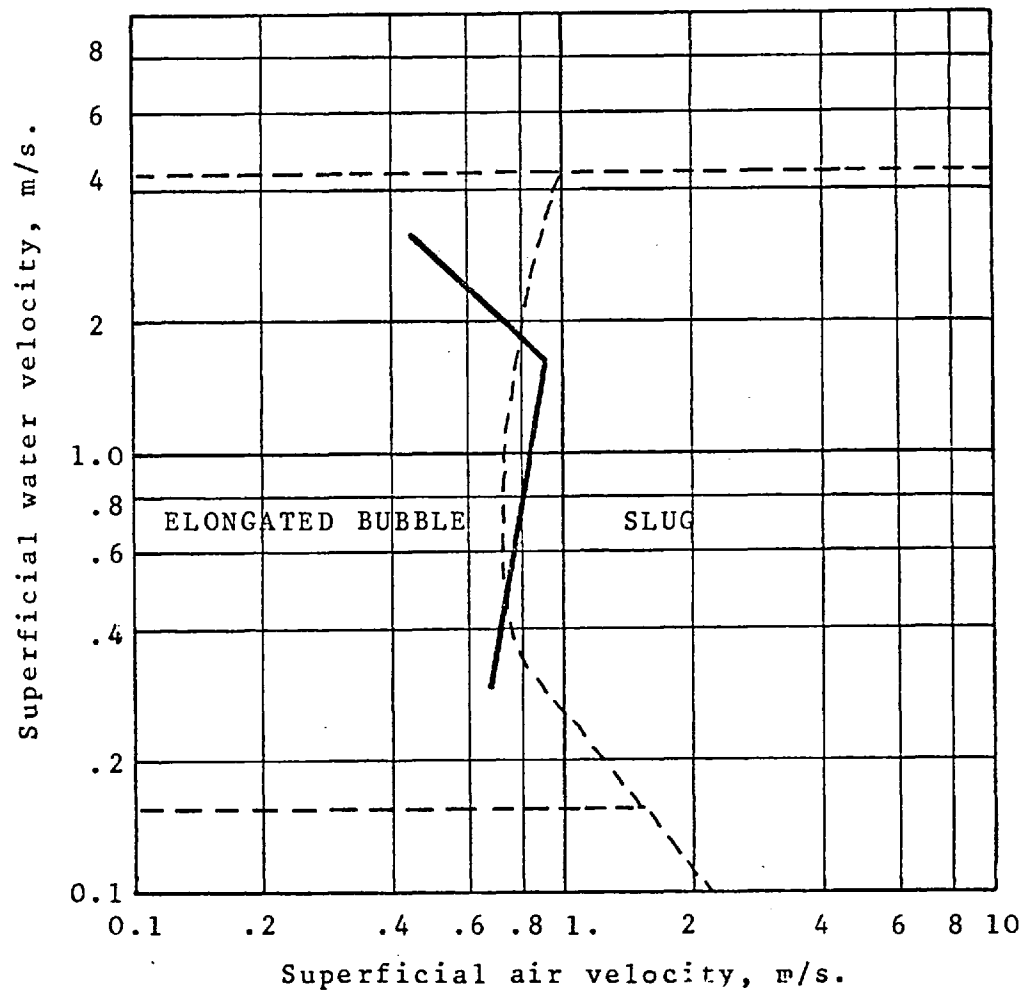
The results for the 17 mm and 11 mm pipes for both sinter disc gas injection, and intense mixing of the phases in the pump, gave approximately equivalent flow patterns and transition boundaries, and were found to be as reported in the literature, within the experimental error and latitude of the various flow maps. The plots of the data points for the case of phase mixing via the sinter are given in the Appendix F in Figures F.1 and F.2 for the 17 mm and 11 mm pipes.

The transition lines alone for these two pipes are given in Figure 6.4. The discrepancy with the reference pattern map is within the normal expected limits for this type of work. The difference in the boundary at a water superficial velocity of over 2 m/s is partly an indication of the difficulty of definition between the elongated bubble and slug flow regimes for pure-water/air flows.

Figure 6.4: Pattern maps for pure water/air flows.



Regime boundary map for 17.3 mm pipe



Regime boundary map for 11.6 mm pipe.

6.6.2) Trace of Detergent

A few runs were carried out with a trace of detergent in the liquid, to check on the effect of a slight lowering of surface tension without giving the ability to form stable, long lasting bubbles. The runs were of a cursory nature, with only two liquid rates being used. The main point of interest was the elongated bubble to slug transition flow rate valves.

The results for runs in the 17 mm pipe, with both sinter and pump mixers, were equivalent to those of the pure water. Few bubbles were produced, most breaking in the second or two it took to enter the test pipe. In the case of elongated bubble flow the long bubbles were often subdivided by several lamella as in Figure 6.5, which tended to drain and break as they went down the pipe.

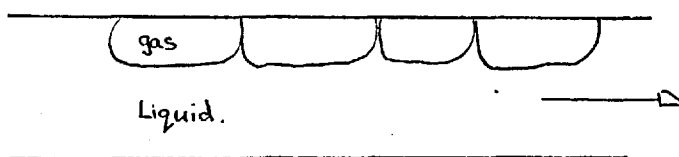


Figure 6.5: Lamellae spanning elongated bubble, with trace of detergent in the water.

In the transition and slug flow regimes there were more small bubbles present than in the case of pure water-air, but this did not alter the phase velocity necessary for transition from one regime to another.

6.6.3) Detergent Solutions

The major factor in the determination of the flow pattern maps which results from the various combinations of detergent concentration, pipe size and phase mixing methods investigated was found to be the generator used in phase mixing. The results

for the two detergent solutions, 0.05% and 0.5% were in each case similar, and hence are given together and compared in relation to the mixing method used.

6.6.3:1) Intense Mixing - Pump Generator

The intense mixing of the air and detergent solutions in the centrifugal pump and/or valve results in a very small bubble size, 20 to 200 microns, the foam produced being a white opaque mixture, with a long life time, having on average a t_{50} of over 2 minutes and bubble size of about 60 microns. See Sections 3.1 and 4 for details of generators and foam lifetime characteristics.

The pressure drop and flow regime results from the pump and the valve were found from initial experimental runs to be equivalent, and in all subsequent runs the pump was used, due to it's more flexible back pressure characteristics which allowed greater gas rate range and control.

All experimental runs with both the 0.05% and 0.5% detergent solutions in the pipe were in the dispersed flow pattern. The intense mixing in the pump meant that no stratified foam flow took place as no drainage had occurred in the time taken to pass through the pipe, 6 metres long. The initial turbulence from the pump had died away after about 1 metre of pipe, and the foam passed down the pipe in dispersed flow. No slugging occurred in the pipe even at the maximum gas rates, of about 5 m/s in the 17 mm, and about 10 m/s in the 11 mm pipe, even though these superficial velocities are on order of magnitude greater than the transition boundary to slug flow for pure water/air.

The superficial velocities covered in the experiments for the two pipes were as shown in Figure 6.2. The maximum phase

velocities reached were due to equipment pressure limitations, the minimum water rate, of 2 litres/min, due to the inability of the pump to operate satisfactory once the liquid flow rate drops below this value, the hunting of the pump speed causes inherent flow instabilities..

6.6.3:2) Good Mixing - Sinter Generator

The sinter generator results in a foam of larger bubble size, 0.1 to 1.0 mm, and when in the dispersed flow regime is a white opaque mixture. The foam coming from the generator was usually flowing in the stratified flow mode as it came round the top bend, details of precise generator configuration are given in Section 3.2. The foam flow regime was then established as the foam flowed down the pipe.

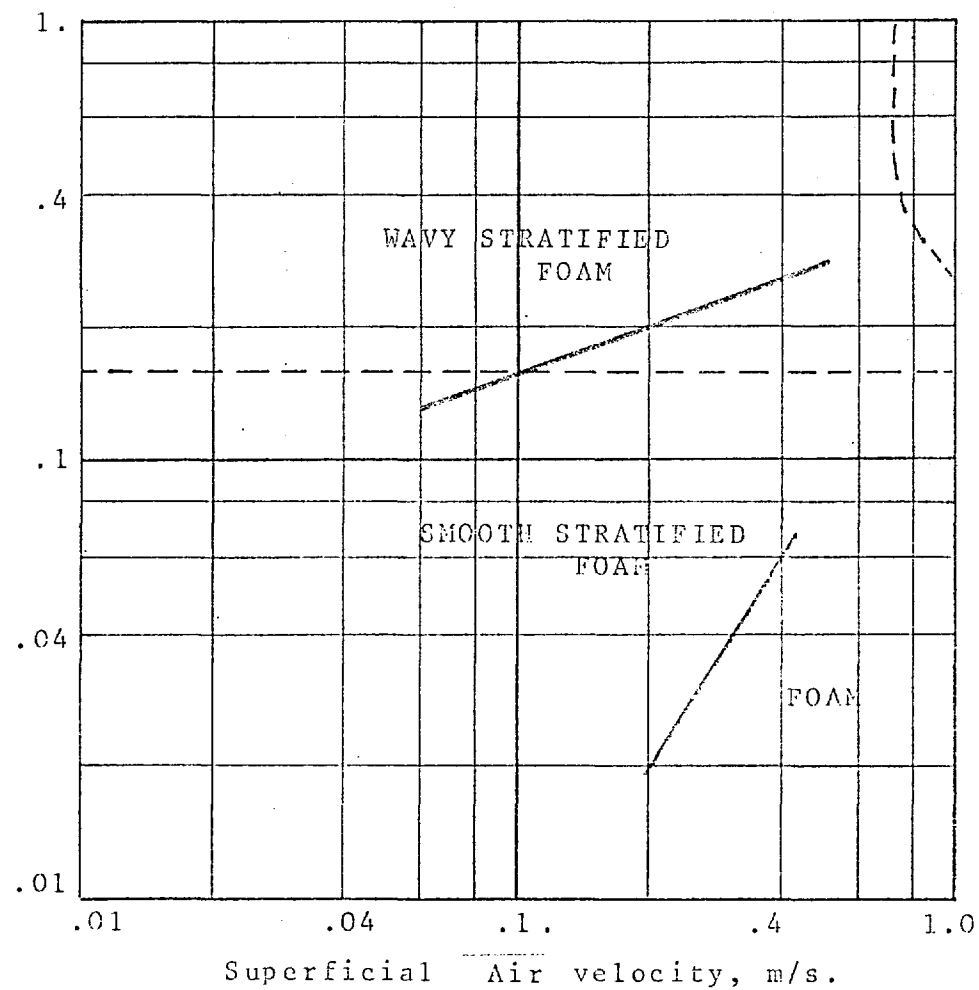
The effects of the two detergent concentrations investigated, 0.05% and 0.5%, are similar and the flow patterns are discussed with relation to the pipe size. The extent of the lines on the various plots is an indication of the areas covered in the experiments.

I. 39 mm and 26 mm

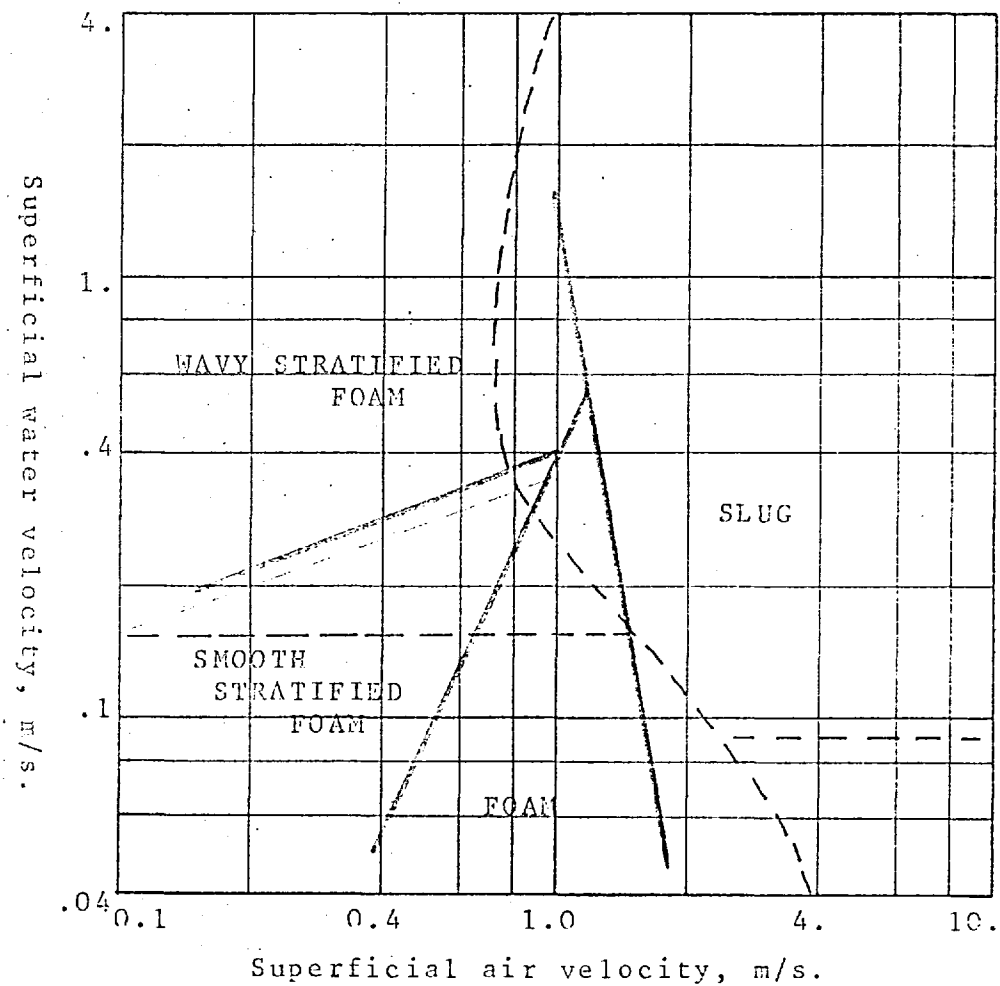
The range of flow velocities which could be reached with these pipes was comparatively low. In the case of the 39 mm pipe flow only in the stratified region was obtained, the plot in Figure 6.6 giving the transition boundary from wavy to smooth interface. The transition from stratified flow to foam flow was based on only a few data points, and was constructed with the aid of the 26 mm pipe flow pattern map.

The flow regimes which were obtained for the 26 mm pipe are given in Figure 6.6 and cover a far wider range of regimes, slug, foam and stratified flow regimes being observed.

Figure 6.6: Pattern maps for 0.5% solution in 39.1 and 26.0 mm



Regime Boundary map for 39.1 mm pipe, with 0.5% detergent, sinter generator



Regime Boundary map for 26.0 mm pipe, with 0.5% detergent, sinter generator.

Included on the map for the 26 mm pipe is the wavy to smooth stratified flow transition boundary for the 39 mm pipe, these lines can be seen to be identical within the experimental errors.

The full data plots are given in the Appendix in Figures F.3 and F.4.

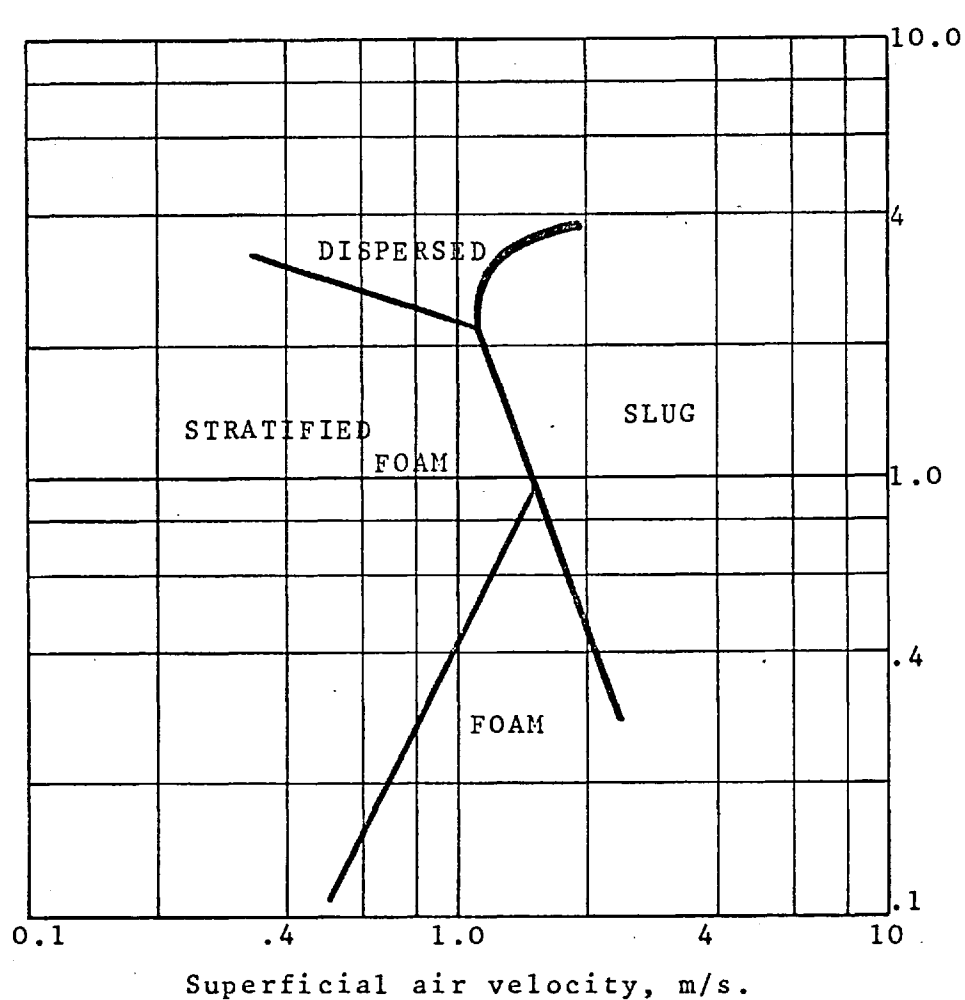
2. 17 mm Pipe

The flow pattern maps for flow in the 17 mm pipe, with 0.05% and 0.5% detergent solution, are given in Figure 6.7. The regimes covered by these plots cover the full range of slug, dispersed, stratified foam and foam flow and the data plots are given in the Appendix in Figures F.5 and F.6.

The plots for the different detergent concentrations can be seen to be of the same form, the only major difference being a shift to slightly high gas velocities of the slugging transition line at the higher detergent concentration. The lack of the transition line from wavy to smooth stratified flow in the case of the 0.05% solution was due to the fact that this minor regime change was not differentiated in the data collection for this particular set of conditions, rather than the non-appearance of this regime.

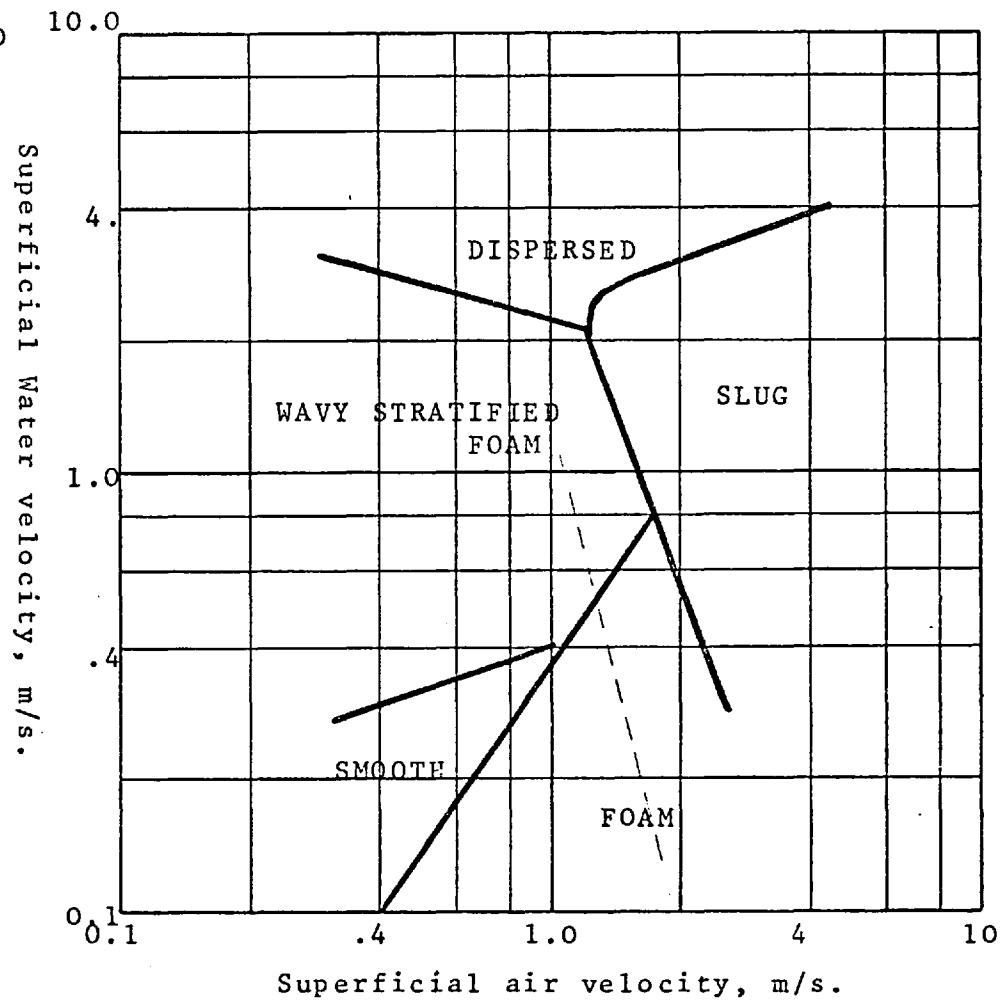
The slug to stratified flow transition for the 26 mm pipe is shown on the map for the 0.5% solution, of note is the shift to higher gas velocities for this transition in the 17 mm pipe.. The stratified foam to foam flow transition boundary is found to be coincident for the 26 mm and 17 mm pipes.

Figure 6.7: Pattern map for 17.3 mm pipe, 0.05% and 0.5% .



Regime Boundaries for 0.05% Solution.

Sinter generator



Regime Boundaries for 0.5% Solution.

Sinter generator.

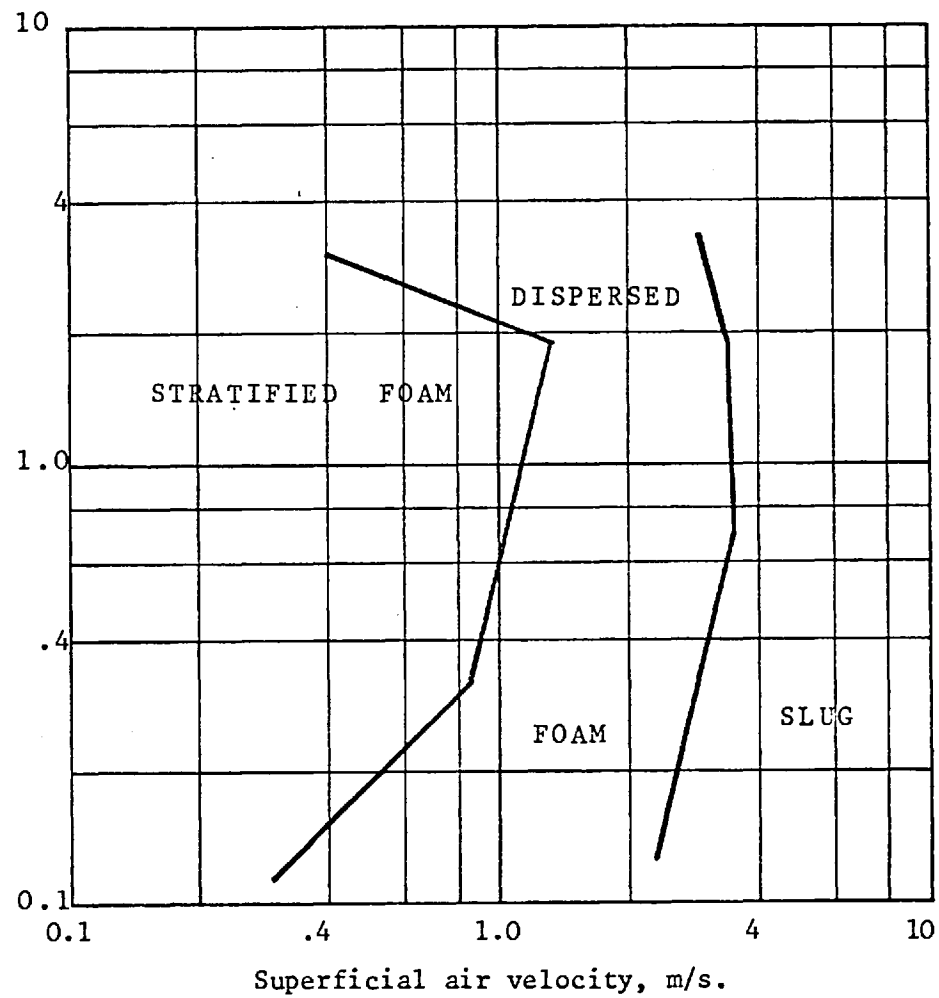
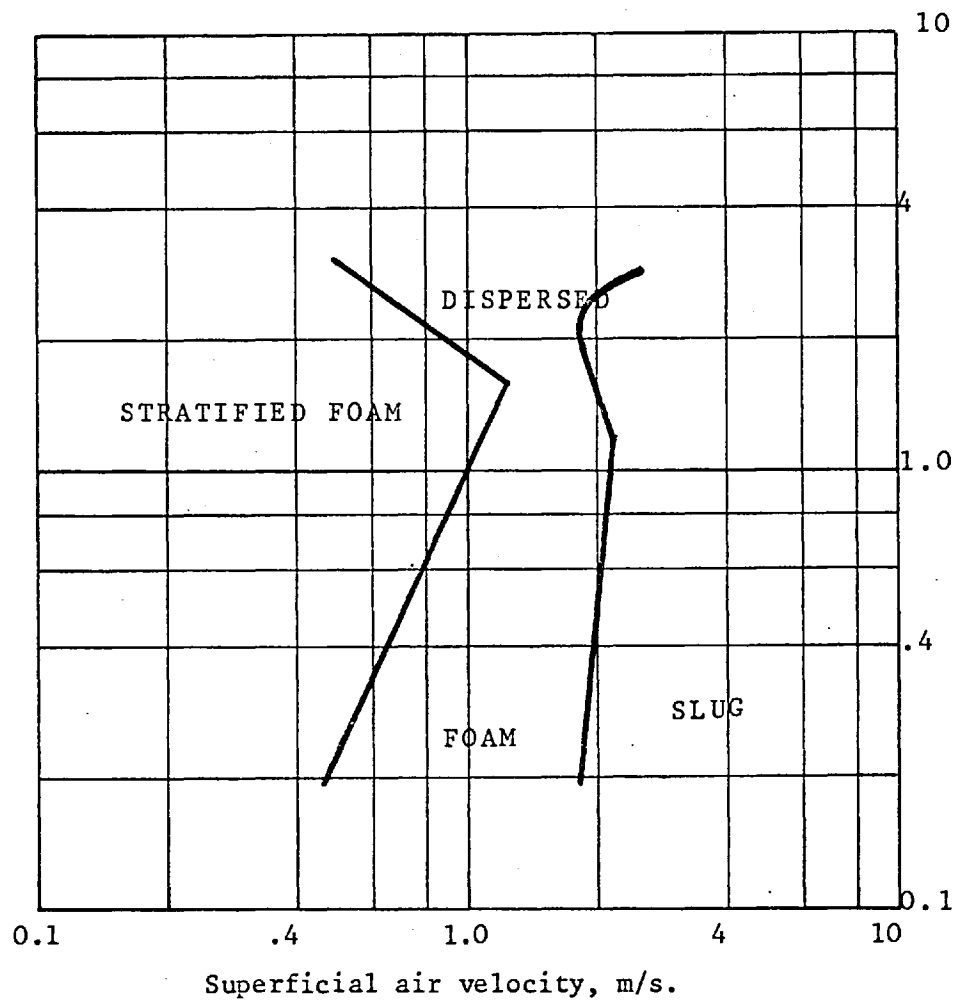
3. 11 mm Pipe

The flow pattern maps for flow in the 11 mm pipe, with 0.05% and 0.5% detergent solutions, are given in Figure 6.8. The regime covered by these plots are the slug, dispersed, stratified foam and foam flow regimes. The presence of the distinct change from wavy to smooth interface was not noted in the stratified flow regime, the interface never taking on a definite wavy or smooth structure at any flow velocities. The full data plots for these conditions are given in the Appendix in Figures F.7 and F.8.

The main points of interest from these plots when compared to those of the 17 mm pipe are:-

- a) The change in the slug flow transition line to higher gas rates, it has moved to a gas superficial velocity that is approximately twice that of the 17 mm pipe velocities. This change is most apparent at the high liquid rates, at the lowest liquid rates the transitional values are approximately equal.
- b) The shift to lower gas, or high liquid rates for the stratified foam to foam transition.
- c) Because of the gross shift of slug transition boundary, the dispersed flow regime now can change to the foam flow regime without passing through any other regime. Any division between these two regimes was unobservable, and would depend on an arbitrary definition of when bubbles in a liquid become the case of liquid lamella in a gas, the two extremes.
- d) The stratified foam to dispersed flow transition remained approximately constant, with relation to the 17 mm pipe transition boundary.

Figure 6.8: Pattern map for 11.4 mm pipe, with 0.05% and 0.5% solutions.



6.6.3:3) Poor Mixing - T Piece

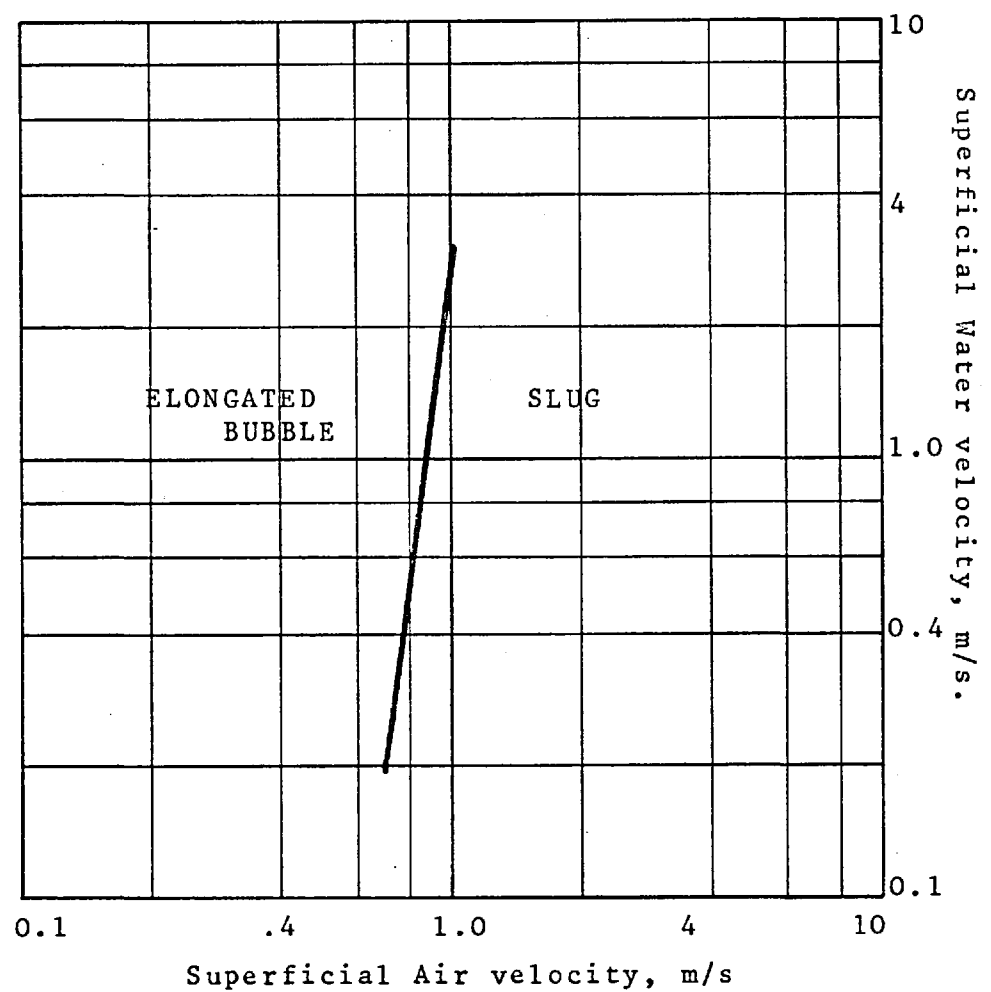
The flow regimes that are produced when the various detergent solutions, 0.05% and 0.5% are flowing in a pipe was investigated for the case when there is no initial mixing of the gas and liquid phases. The introduction of the gas and liquid phases was carried out using a simple T- piece, which allowed stratified introduction of the phases with little or no interphase mixing, see Section for details on T- piece.

The experiments were carried out in the 11 mm pipe as an initial test, and because of the indicated results were not repeated with other pipe sizes.

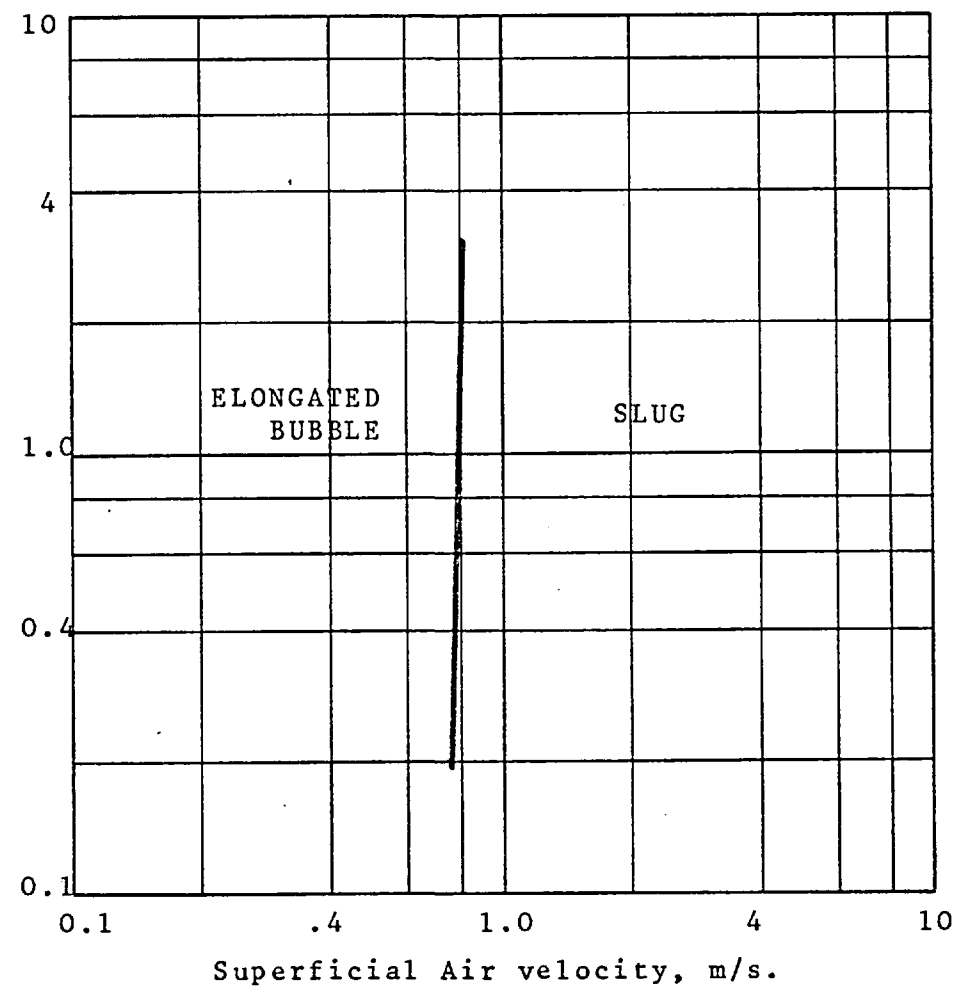
The flow regimes, and transition flow rates between them, were found to be the same within experimental errors for both solutions, and to correspond to those reported in the literature for pure water/air flows. This fact of correspondence could be expected as the effect of surface-tension, the only factor changed in these runs has been shown to be small in non-foaming systems. The phase introduction method was designed to eliminate bubble formation and hence the detergent solutions would act as liquids with low surface tensions but no foaming would take place, which would give the results obtained. However reported work has been for intrinsically non-foaming liquids, usually of high purity to limit any tendency to foam, and the effect of the detergent solutions in the slugging regime could have been expected to change the flow regimes.

The plots for the cases of 0.05% and 0.5% solution in the 11 mm pipe are given in Figure 6.9, the full data being given in the Appendix in Figures F.9 and F.10.

Figure 6.9: Pattern map for detergent solutions with low mixing in the 11.4 mm pipe



Regime Boundaries for the 11.4 mm pipe,
with T-piece mixer, 0.05% solution.



Regime Boundaries for the 11.4 mm pipe.
with T-piece mixer, 0.5% solution.

6.6.4) Regime Boundary Co-ordinates

The co-ordinates for the regime boundaries of the flow patterns shown on the various maps is given in the tables below. Where a line consists of several straight sections the co-ordinates of each turning point are given. If a line has a curved section in the line the end points of the curve are given as

$$(x_1, y_1), \text{ curve, } (x_2, y_2)$$

which means that point (x_1, y_1) and point (x_2, y_2) are joined with a curved section, inspection of the relevant figure giving the general shape of the curved boundary. A full description of the flow regime abbreviations is given in the appendix. The first figure given in the bracket is the superficial gas velocity, the second the superficial liquid velocity, in M/S ie., (V_G, V_L) .

1) 39.1 mm. pipe.

Sinter generator, 0.5% Teepol solution.

St-W	to	St-S	(0.06,0.13), (0.5,0.28)
St-S	to	F	(0.2,0.02), (0.4,0.06)

2) 26.0 mm. pipe.

Sinter generator, 0.5% Teepol solution.

S1	to	(St or F)	(1.06,1.0), (1.72,0.06)
St	to	F	(1.20,0.57), (0.42,0.06)
St-W	to	St-S	(0.16,0.2), (0.98,0.39)

3) 17.3 mm. pipe.

a) Sinter generator, 0.5% Teepol solution.

S1	to	(St or F)	(1.24,2.1), (2.55,0.3)
S1	to	D	(1.6,2.8), (4.2,4.0)
D	to	St-W	(0.3,3.1), (1.25,2.1)
F	to	St	(0.4,0.1), (1.75,0.8)
St-W	to	St-S	(0.4,0.3), (1.0,0.4)

b) Sinter generator, 0.05% Teepol solution.

S1	to	St	(1.1,2.22),(2.3,0.3)
S1	to	D	(1.1,2.22),curve,(1.9,3.9)
D	to	St	(0.33,3.2),(1.1,2.22)
St	to	F	(0.48,0.1),(1.58,0.9)

c) Sinter generator, Water.

EB	to	D	(0.15,2.6),(1.2,2.6)
EB	to	S1	(0.6,0.3),(0.8,1.5),curve,(1.2,2.6)

4) 11.4 mm. pipe.

a) Sinter generator, 0.5% Teepol solution.

S1	to	(F or D)	(2.3,0.13),(3.5,0.7),(3.4,1.9), (3.0,3.0)
St	to	(F or D)	(0.3,0.115),(0.85,0.32),(1.32,1.9), (0.4,3.0)

b) Sinter generator, 0.05% Teepol solution.

S1	to	F	(1.83,0.2),(2.2,1.2)
S1	to	D	(2.2,1.2),(1.85,2.0),curve,(2.7,2.9)
St	to	(F or D)	(0.46,0.2),(1.25,1.6),(0.5,0.3)

c) Sinter generator, Water.

EB	to	S1	(0.68,0.3),(0.92,1.6),(0.45,3.0)
----	----	----	----------------------------------

d) T-piece generator, 0.5% Teepol solution..

EB	to	S1	(0.76,0.2),(0.81,3.0)
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e) T-piece generator, 0.05% Teepol solution.

EB	to	S1	(0.7,0.2),(1.02,3.0)
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6.7) Discussion and Conclusions

6.7.1) Water Alone

The results for flow pattern maps for the water flowing alone with air were as reported in the literature, and confirm the operation of the equipment. The work showed the difficulty in establishing the elongated bubble to slug transition for pure water at high liquid rates, a factor which is brought out in the many flow maps given in the literature. The type of generator used to mix the phases was not found to effect the flow patterns, although detailed checks were only carried out with the sinter generator.

The many differences brought out by the use of the foaming solutions were highlighted to a much greater extent by their comparison with these pure water-air runs, and this type of comparison should be made in all such experimental work. The problem of cleaning the equipment to remove detergent traces should however be investigated, as repeated water to detergent solution changes can cause problems.

6.7.2) Trace of Detergent

The results of flow regimes for the case of a trace of detergent in the water, 0.002% Teepol by volume, were found to be similar to those of the pure water-air flow. The experiments were however only of a very cursory nature and by no means exhaustive. The experiments were carried out when it was found to be a problem cleaning the equipment for the pure water tests, see Section 3.1 for details of this preparation. The amount of detergent gave a lowering of surface tension to 59 dynes/cm without allowing the formation of any stable foam.

Even under intense agitation bubble sizes were as large as for pure water, and they broke within seconds of reaching the free air surface in the holding tank.

Hence the reported conclusion that surface tension has little major effect on flow regimes has been borne out, and also that excessive care to exclude any slight impurities to limit foaming has been shown to be unnecessary, if foam stability is extremely low. However the case of flow in long pipe lines in which any initially small effect could grow to dominate is still unclear, but the "degree" of foaminess of the solution will be the important factor, rather than its intrinsic surface tension.

6.7.3) Detergent Solutions

The effect of the various parameters of pipe size, generator type and detergent concentration, on the flow regime maps is a complex matter. The following discussion attempts to isolate each variable in turn, and indicate its effect on the regime maps.

6.7.3:1) Generator Effects

The type of generator used for phase mixing plays a very important part in determining the flow regimes which exist in the pipe with the detergent flows. Any discussion of foam flow regimes will need to include detailed information of the generator, and the type of bubble size produced by it. The three generator types used in this work gave a wide range of bubble sizes.

- 1) Intense mixing - very small bubble size, possibly the smallest bubble size that can exist for any time was produced by the pump.

2) Good mixing - small bubbles of a fairly uniform size were produced, but for the type of sinter used many other bubble sizes and size ranges are possible by quite minor design changes.

3) Low mixing - effectively no bubbles were produced by the generator, and any bubbles formed in the pipe will be a function of fluid properties and flow rates.

1) Intense Mixing

The flow was always dispersed, as could only be expected with the short in-pipe lines used. Samples of this type of foam in a static test condition show no separation of bubbles and liquid in the first 10 to 20 seconds, and hence the effect of shear in the pipe will help to keep a uniform foam with only a single 'phase' for a significantly longer time period in the pipe. This foam can be considered as the nearest approach possible to a homogeneous fluid for a gas/liquid mixture, and the presence of only one flow regime over the test range is not unexpected. The possibility of bubble breakdown and coalescence must however be borne in mind at much higher shear rates, but in these cases pressure drops will be high, and hence expansion effects as the foam passes down the pipe may be the predominate factor in it's behaviour, rather than surface tension effects.

2) Good Mixing

The flow regimes obtained with the sinter generated foams showed the widest range of regimes, with marked changes from the normal water-air flows. The main factor in these changes was the formation of large numbers of stable bubbles to form a stable foam. The stable foam then takes the place of the gas phase in normal liquid/gas flows, and the differences are due to the properties

of this new pseudo-phase.

The stratified foam flow region can be explained with relation to the foam and water properties. The situation being comparable to that of an oil and water mixture flowing in stratified flow, a system with very similar regime and pressure drop properties. The transition to and from the wavy and smooth stratified flow regimes is a function of the turbulence set up in the lower liquid layer.

The foam flow regime exists because of the inherent arrangement necessary to provide the water to surround the bubbles in the foam. The transition from stratified flow occurs when the water/air ratio is only just sufficient to fill the spaces between the bubbles, and hence no liquid layer is present.

The reason for the water rate for the dispersed flow region boundary being approximately half that of the pure water/air case is due to the small bubble sizes produced by the detergent. The smaller bubbles require less turbulence to create a flow sufficiently homogeneous to be described as disperse.

The slug transition boundary is shifted significantly to high gas rates. Slugging occurs at a point when large air slugs can be formed, the small bubble size requires greater shear forces to generate these conditions. The dependence of the transitional values on the pipe size is however an additional complication, and is in marked contrast to results from pure liquid-gas flow systems.

3) Low Mixing

The flow regimes observed with the low mixing of the phases were identical to those for the pure-water/air system. Hence, this further emphasises that the major factor in regime determination for detergent solution flow is the bubble size and number. In the case of very low detergent concentrations, bubbles will not form even with agitation, in the low mixing case no bubbles are generated within the test system by careful phase introduction.

This similarity however is unlikely to be the case when the pipe system involved is much longer than the 6 metres used in these experiments. Any bubbles that form in the pipe via flow disturbances or turbulence, will not coalesce rapidly, due to the detergent present. If a sufficient number of bubbles are formed the flow regimes could change, to match those obtained with the sinter generator.

The transition from the elongated bubble to the slug flow regime, with detergent present, occurred at the same superficial phase velocities as the pure liquid/air flows, but it was apparent that there was a larger number of small bubbles present in the bulk liquid between the elongated bubbles or slugs. These small bubbles were being formed at the rear of the large air bubbles, breaking off, and passing into the bulk of the liquid. The following elongated bubble which travels faster than the liquid, then tended to catch up these bubbles, carry some along with it, and allow the rest to pass around it with the bulk liquid. The small bubbles tended to gather in the upper part of the pipe and hence much interaction occurred, resulting in the formation of additional small bubbles and a reduction in the elongated

bubble size.

Hence at the higher gas and liquid flow rates, the gas, although still present mainly in the elongated bubbles, starts to travel as small bubbles at the front and rear of the larger elongated bubbles. This change in the mode of gas transport with position in the pipe was noted in the 6 metre pipe, but was not of sufficient magnitude to cause any regime changes, but does mean that the flow pattern can change along the pipe. The initial flow pattern is established quickly, within approximately 50 pipe diameters, as is normal for gas/liquid flows. Further down the pipe the elongated bubbles are starting to break up, and with sufficient pipe length at high enough velocities to produce total bubble breakup, the flow regime could change to stratified foam flow. The ultimate flow regime, in a very long pipe, for a foaming solution, with phase velocities which would be within the elongated bubble region for pure liquid-gas flows could be expected to depend on a balance between the forces causing bubble generation in the bubble wakes, and the bubble coalescence due to the surface active properties. The effect of generation method on the ultimate flow regime might be unimportant for very long pipelines. If bubble coalescence or drainage factors are controlling, in-situ pipe times of may be order $[10 \times t_{50}]$ might be needed to stabilise the flow, times of order $[200 \text{ secs}]$ or pipe lengths of order $[300 \text{ m}]$ under the conditions used in these experiments. These pipe lengths are well beyond any reported experimental pipe lengths, but are well within the scale of industrial situations, particularly for long pipe lines for oil transportation.

Recent work by Aziz et al (53) has indicated that the

elongated bubble region can be sub-divided into several sub-regions, and they discuss the mechanism of small bubble generation in the wake of elongated bubbles. Their system involved pure liquid, and bubble coalescence was the same order of time as the elongated bubble transit time, so small bubble build up was not noted. The detergent solution in this work results in a 'fixing' of any bubbles formed, and hence the effect on the flow pattern in long pipe is greatly enhanced. Their work also pointed out the need for photographic techniques on a large scale, to aid in consistent pattern identification when looking for such fine detail and also stressed the need to use relatively long test pipes, in their case 80 ft., if these types of flow pattern effects are to be identified and adequately investigated.

6.7.3:2) Detergent Concentration Effects

The effect of very low detergent concentrations at which the liquid cannot be considered to form a foam, have been discussed in the previous section.

The effect of varying detergent concentrations when the solutions formed are capable of forming a stable long lasting foam, can be seen in the comparison of plots for 0.05% solution in the 17 mm and 11 mm pipes, Figures 6.7 and 6.8.

The change in detergent concentration from 0.05% to 0.5% could be expected to give some marked change. The solution at 0.05% Teepol by volume is below the critical miscelle concentration (C.M.C) while at 0.5% it is well above it. This change would be expected to manifest itself in some measure of the foam stability.

The changes in flow regimes and their boundaries that occurred for this detergent concentration variation are small

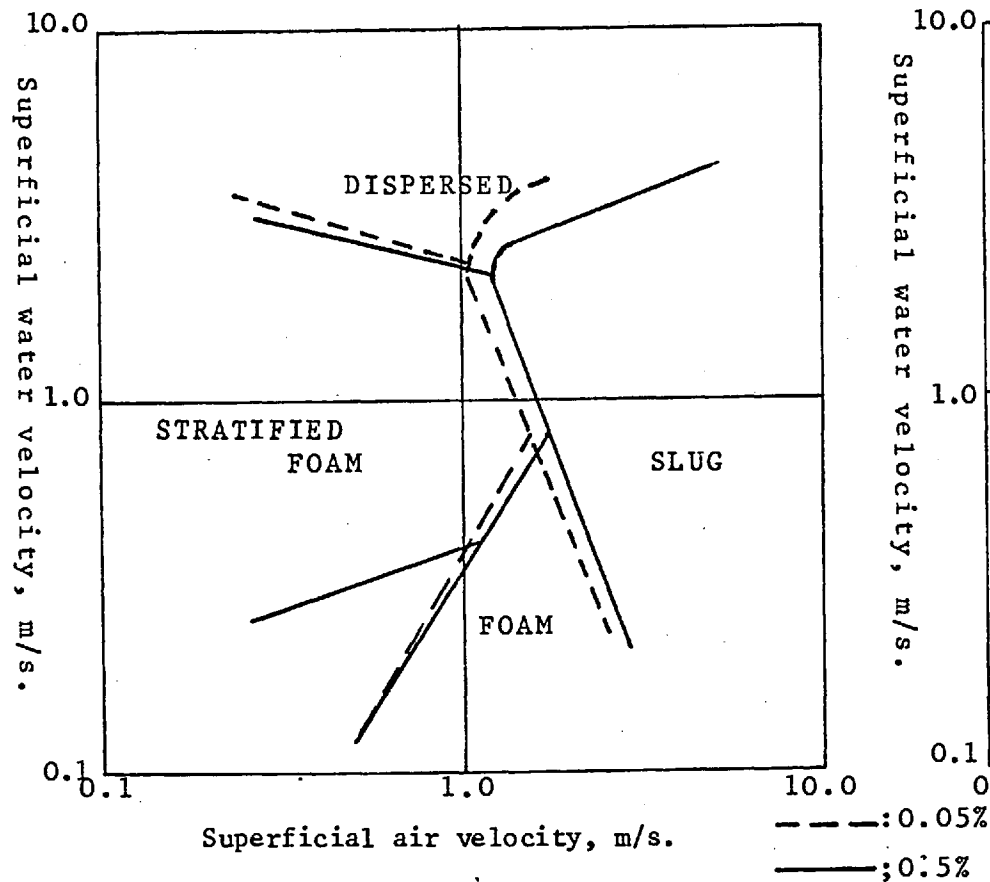
for the case of the 17 mm pipe. Only the slugging transition boundary has been effected, there being a shift of approximately $1.1 \times V_G$ at constant V_L , for the 0.5% solution when compared to the 0.05% solution. This comparatively small shift especially when compared to the boundary identification errors involved when using pure liquids, is however a real and significant change. The slugging transition for foam flows is very clear, and gives a better reproducibility than this difference. However the transitions from stratified foam flow to dispersed and foam flow are not too distinct, and hence these transition boundaries must be considered to be identical, the discrepancy on the plots being due to experimental and identification errors. These lines are shown in figure 6.10.

The differences for the 11 mm pipe with the two detergent solutions are again mainly in the slugging transition line, but there is a very marked shift to high gas velocities, $1.5 \times V_G$ at V_L constant. The differences in the other transition boundaries can again be attributable to identification problems. These lines are shown in Figure 6.10.

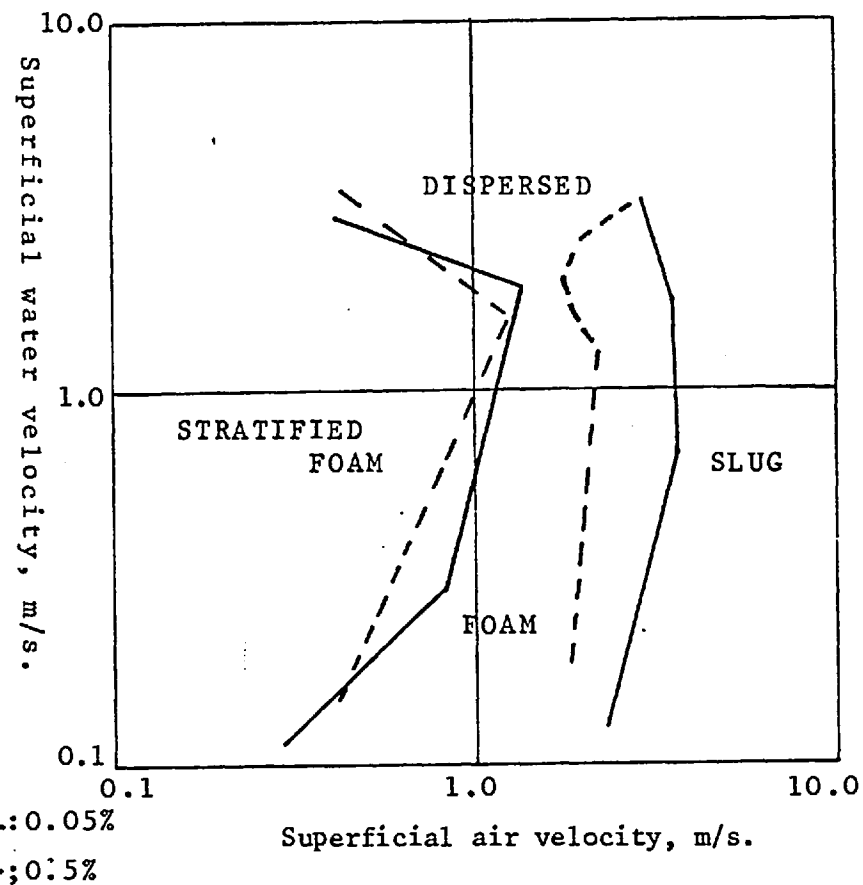
This shift to higher gas velocities, or greater gas/liquid ratios before slugging occurs can be attributed to the detergent concentration effect. The strength of the bubble depends on the surface tension properties of the liquid bubbles in 0.5% solution could be expected to resist breakdown to slugs better than in the 0.05% solution. The forces causing breakdown will be the shear forces acting on the foam, and these are capable of causing breakdown of the bubbles when phase velocities reach a certain magnitude.

This point is reached at a lower velocity for the less

Figure 6.10; Effect of detergent concentration on the flow pattern maps for the 17.3 and 11.4 mm pipes, with sinter generators



Comparison of 0.05% and 0.5% detergent pattern maps, in the 17.3 mm. pipe



Comparison of 0.05% and 0.5% detergent pattern maps, in the 11.4 mm. pipe.

concentrated solution.

The effect of different concentrations all beyond the C.M.C was not investigated, but would not be expected to change the slugging transition boundary that was obtained for the 0.5% solutions.

All detergent solutions, ^{when} tested in a mode which gave a 'good' foam, caused a marked shift to higher gas velocities when compared with the pure water case. The shift varied from about 1.5 x for the large pipe to over 4 x for the smallest pipe. This shift is highly significant and the detergent causes a marked change in the transition to the slugging regime.

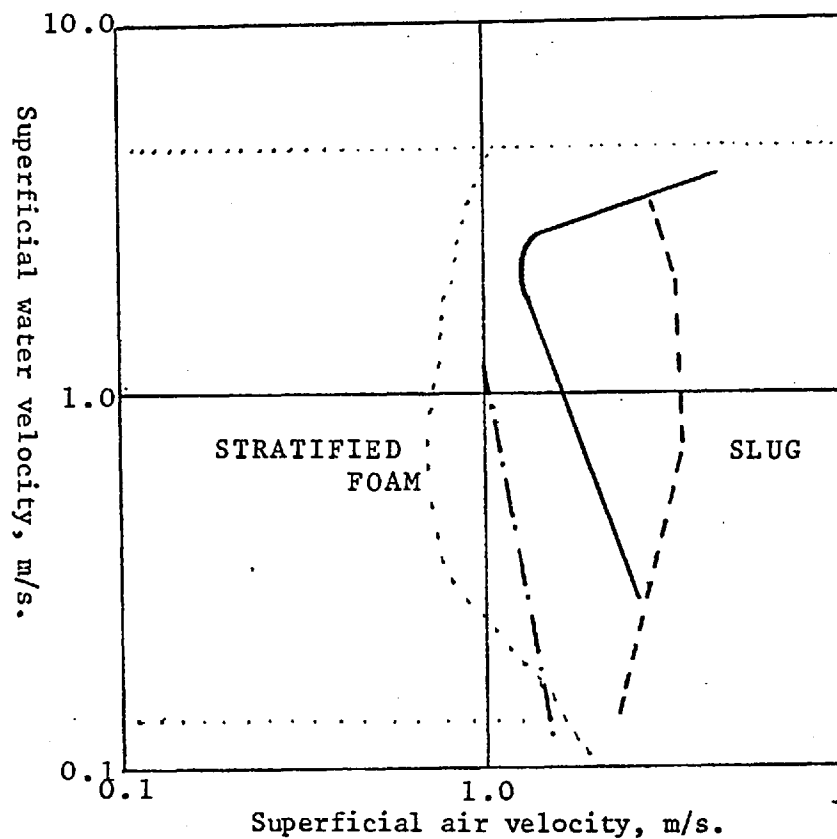
6.7.3:3) Pipe Size Effects

The effect of reducing pipe size on the flow regime maps was to cause a progressive shift to higher gas superficial velocities of the slug flow regime. The effect of the various pipe sizes can be seen in Figure 6.11.

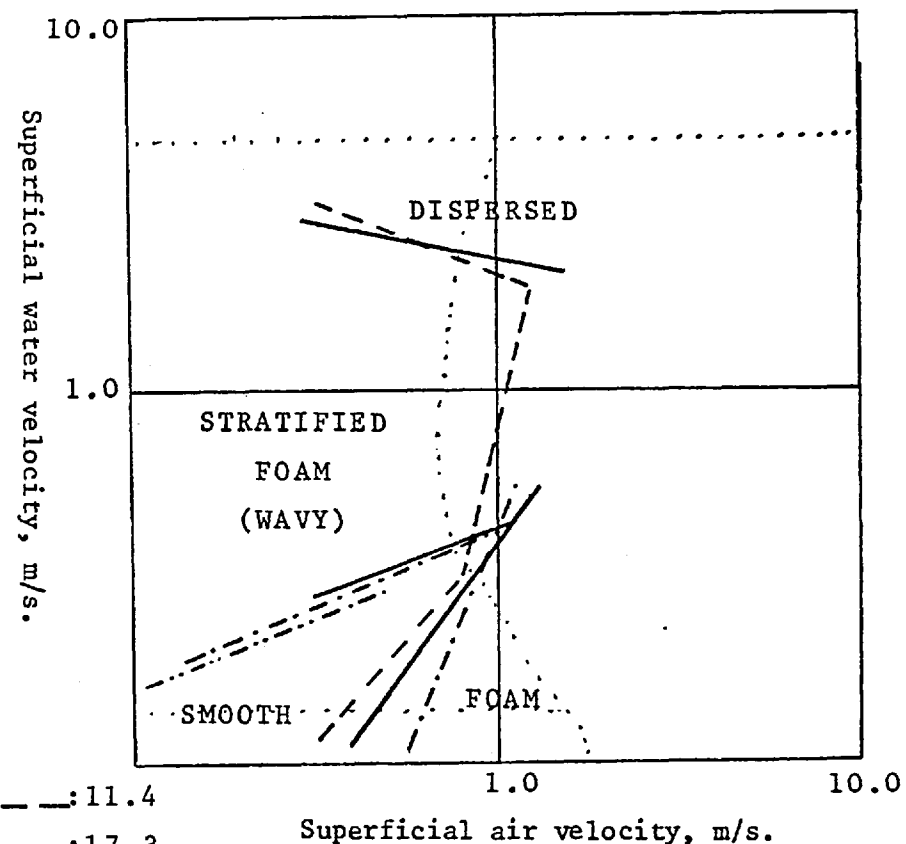
The other transition boundaries for the pipes, the dispersed to stratified foam, and stratified foam to foam flow, are approximately identical, allowing for identification errors. The wavy to smooth stratified flow transition, for the cases in which it was observed, that is all but the 11 mm pipe, was found to be coincident. These lines are also given in Figure 6.11.

The change in the slug transition boundary, as pipe size decreases, shows that the mechanism for air slug formation with a foam present is partly controlled by the pipe dimensions, unlike the pure water-air flow systems, where pipe size has been shown to be unimportant. The precise mechanism involved is not clear, the foam flow situation being in itself

Figure 6.11: Effect on the Transition boundaries of the different pipe sizes, 0.5% detergent, sinter generator.



Transition Boundaries for the slug regime, 0.5% detergent, sinter.



Transition Bounbaries for Stratified foam to dispered and foam regimes, 0.5% detergent, sinter generator.

---: 11.4
 —: 17.3
 -.-: 26.0
 -.-.-: 39.1
: pure water

difficult to model. The way the flow could generate sufficient shear or stress forces to change small bubbles to large air slugs is beyond the scope of this work.

6.7.4) General Foam Flow Map

A general foam flow map was prepared from all the pipe data and is shown in Figure 6.12.

The transition boundaries for the dispersed, stratified foam (wavy and smooth) and the foam flow regimes are the same for each pipe size, and are indicated with solid lines. The transition boundary for the slug regime depends on the pipe diameter and is given for the cases of 26, 17 and 11 mm pipes.

This flow map could be expected to apply to any 'good' foam flowing in a horizontal pipe of similar dimensions to the pipes used in this work.

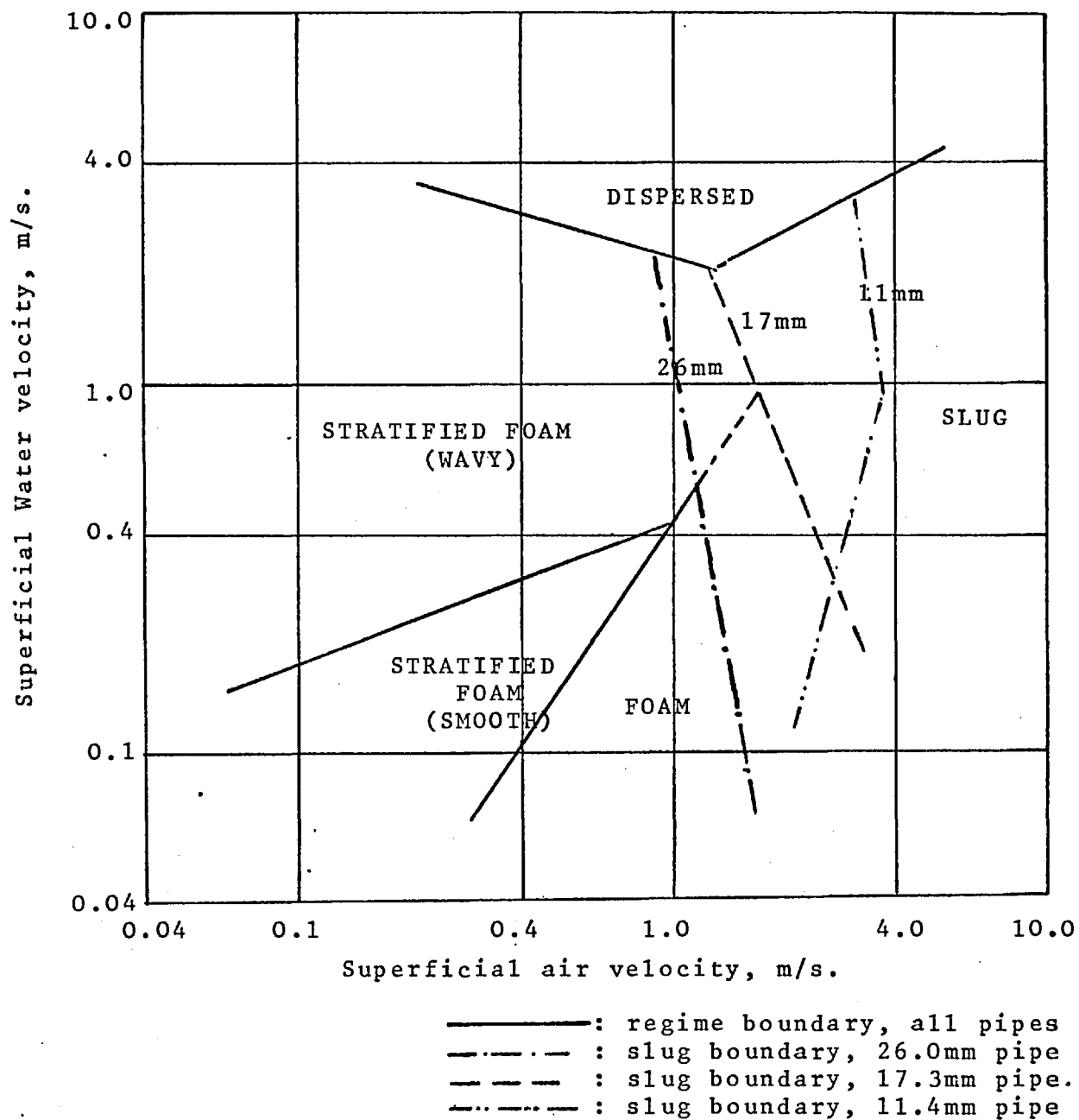


Figure 6.12: General Foam flow Pattern Map for Horizontal pipes.

Foams generated with sinter disc, and a good foaming agent.

7) Pressure Drop Results

7.1) Pressure Drop for Single Phase Flows

The pressure drop for detergent solution flowing alone in the pipe was found for the range of values covered in the foam flow experiments. The results were obtained over a large number of experiments, and were carried out to

a) calibrate the rotameters, flow rates being calculated both by rotameter chart reading, and by direct sample collection and measurement.

The rotameters were found to be accurate within $\pm 1.5\%$ over that range.

b) check the operation of the pressure measurement system, and compare with calculated pressure drops.

c) to obtain data for use in foam flow chart construction, and modelling purposes.

No pressure drop data for air flowing alone in the test pipes was obtained, the manometer system being unsuitable for the measurement of such low pressure drops. Values required were calculated by the use of fluid properties and the friction factor/Reynolds number correlation.

The results for the 17.3 and 11.4 mm pipes are given in Figure 7.1, and compared with the line used in the Lockhart-Martinelli correlation, which was based on properties of water. The pressure measurement system was found to work well, giving accurate and reproducible results. The manometer system, when using air-liquid flow in the test pipe, was checked at the completion of each series of runs by use of a water only flow check.

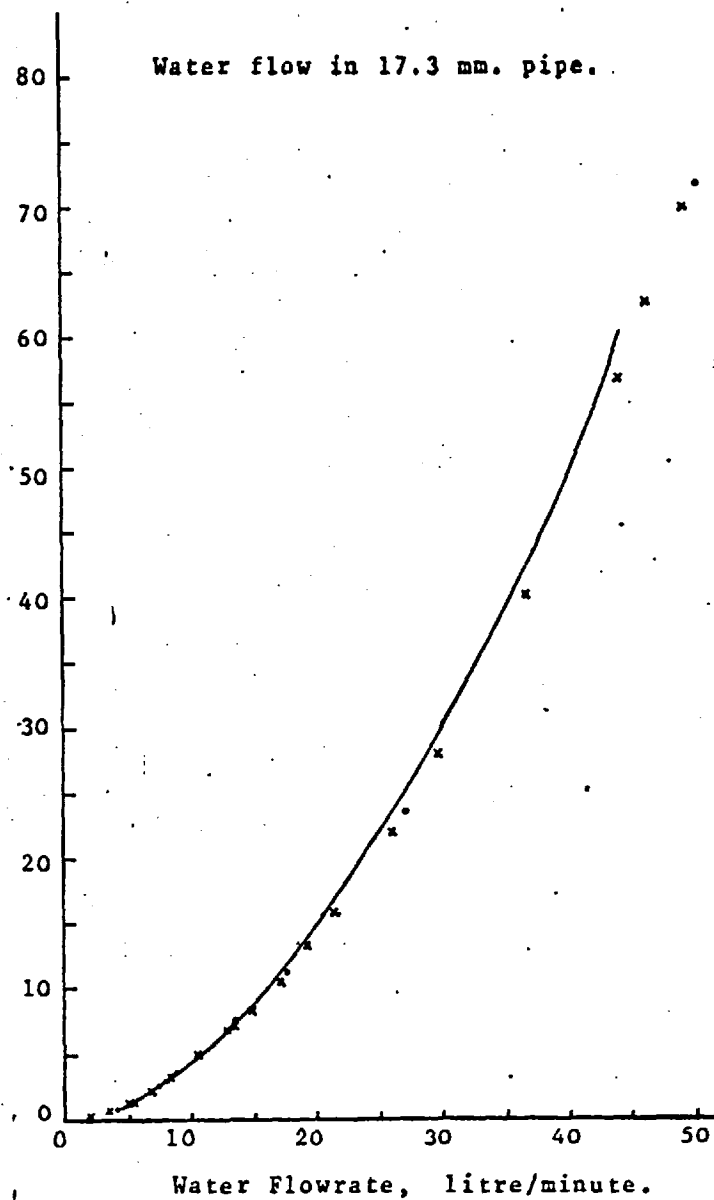
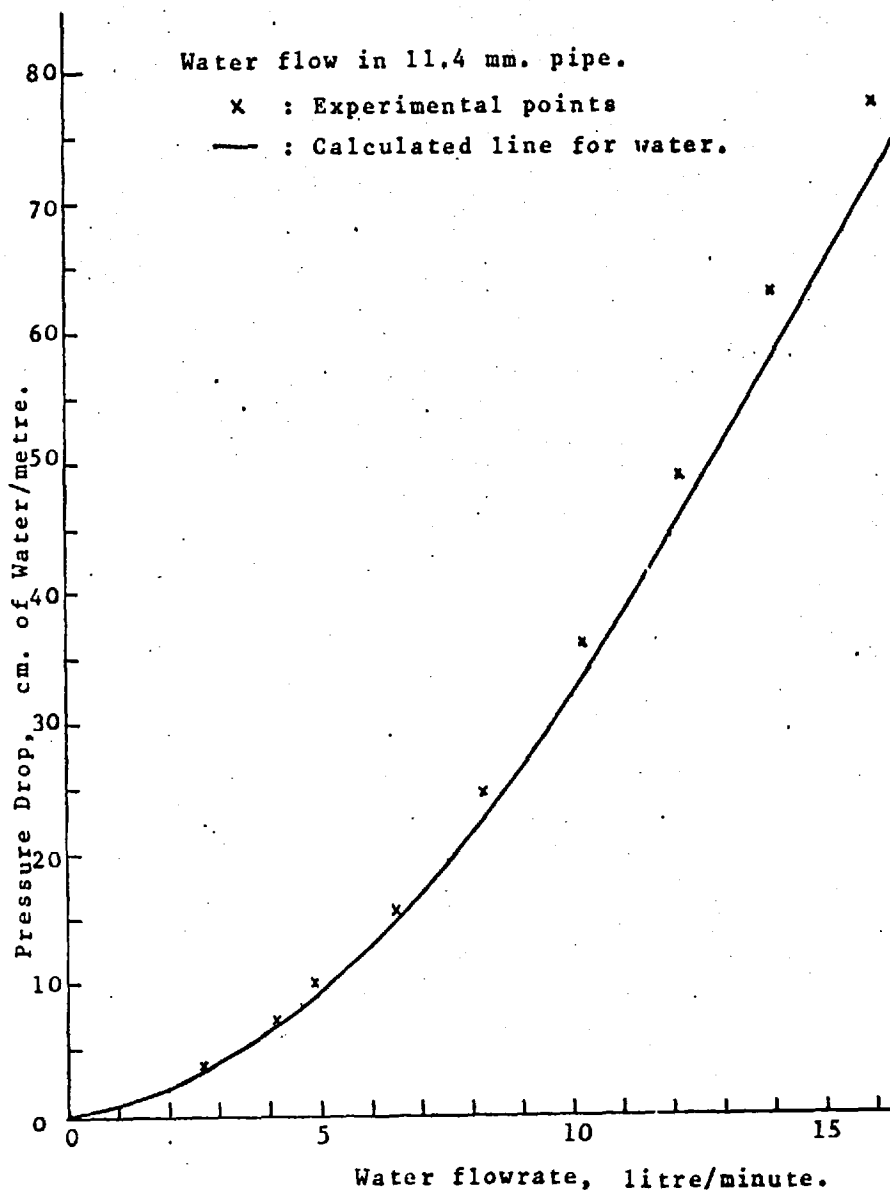


Figure 7.1: Pressure drop for water flowing alone in the 17.3 and 11.4 mm. pipes, with the results calculated for use in the Lockhart-Martineili plots.

7.2) Pressure Drop for Air-pure Water Flows

The results for the pure water-air flows in the 17.3 and 11.4 mm pipes are shown in Figures 7.2 and 7.3, for a range of air to liquid ratios. The increase in pressure drop with mixture flow rate, for any particular phase ratio, is clearly shown.

The data and plots used to construct these figures are given in the Appendix. The results for the 11.4 mm pipe were obtained at ambient pressures of 1.10 ± 0.10 bar and 1.30 ± 0.20 bar, and were found to give slightly different pressure drops. The values for the lower ambient pressure are given in Figure 7.3, the difference between results being small.

At any given water flow rate in the pipes the pressure drop increases linearly, from the water alone value, as the air rate increases. A change of shape occurs when the flow regime changes. This change, when apparent, always occurs at the regime transition flow rate, and results in a reduction in the slope. Pressure drops at the higher gas flow rates, when slugging flow occurred, were found to oscillate with slug passage in the test length. Variations were approximately ± 1 cm. water/metre, the length of manometer lines causing a reasonable degree of smoothing. When taking readings in the slug flow regime the taps on the relevant manometer lines were closed simultaneously, so that the readings would be an instantaneous measurement of the pressure drop.

The results of pressure drop V'S flow rate at constant phase ratio showed no evidence of any yield stress. The pressure drop at any volumetric flow rate depended on the mixture density, reducing density giving reduced pressure drops. At any particular water flow rate the presence of the air increased the pressure drop. The results compare favourably with similar results reported (27, 28), and correlate well with the Lockhart-Martinelli model.

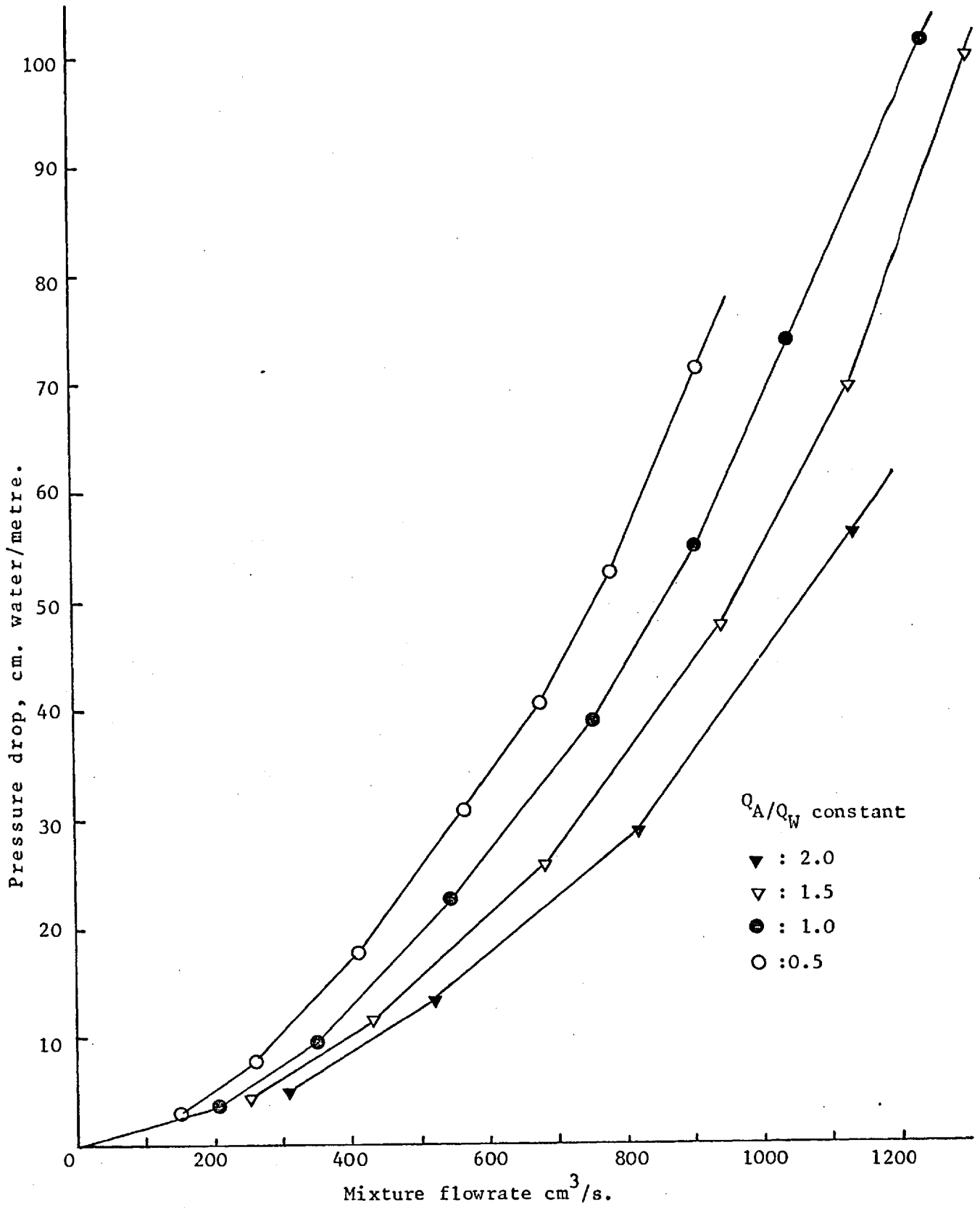


Figure 7.2: Pressure drop v's Mixture flowrate in the 17.3 mm. pipe, sinter generator, pure water/air flows.

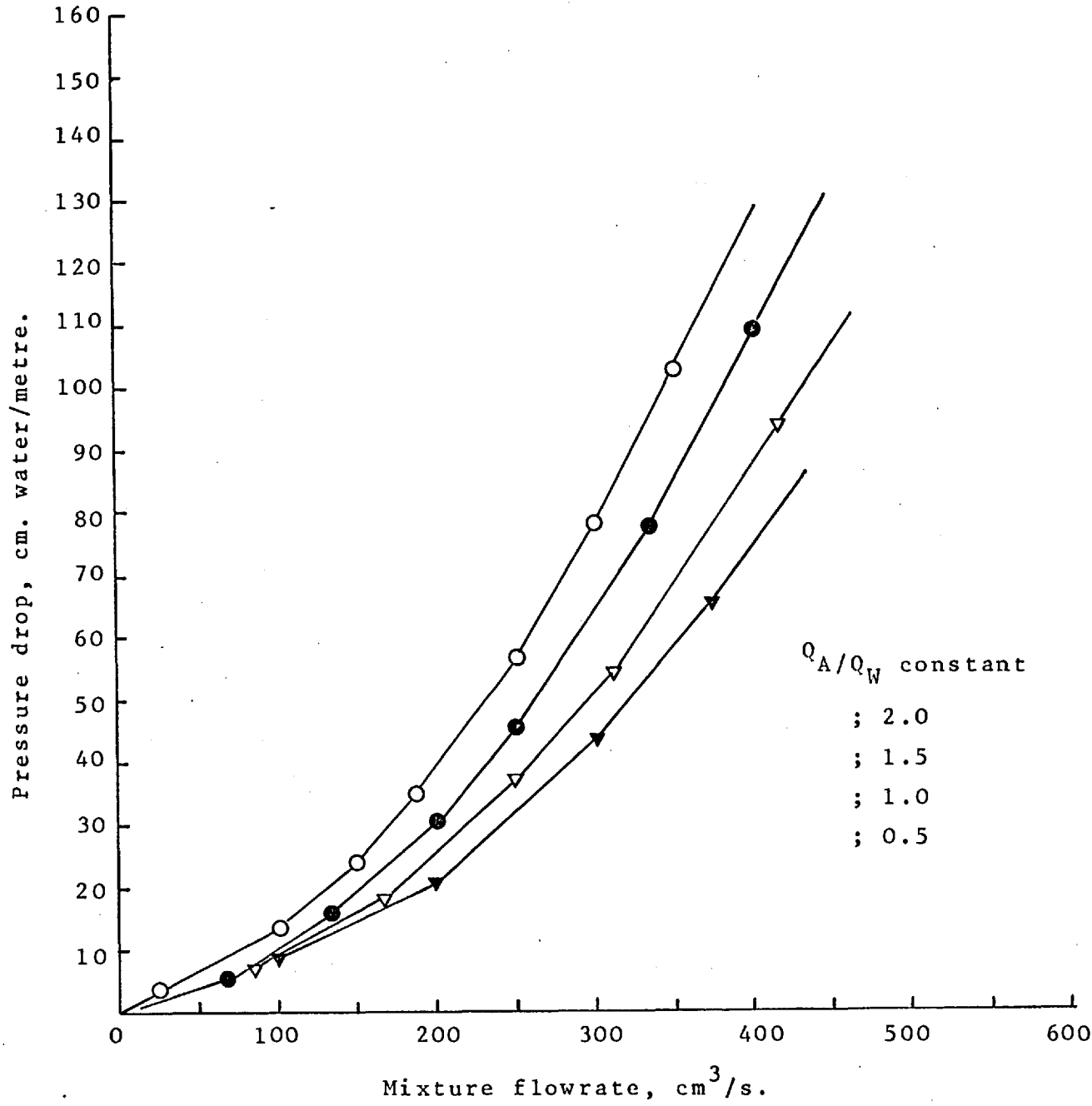


Figure 7.3: Pressure drop v's Mixture flowrate in the 11.4 mm. pipe, sinter generator, pure water/ air flows.

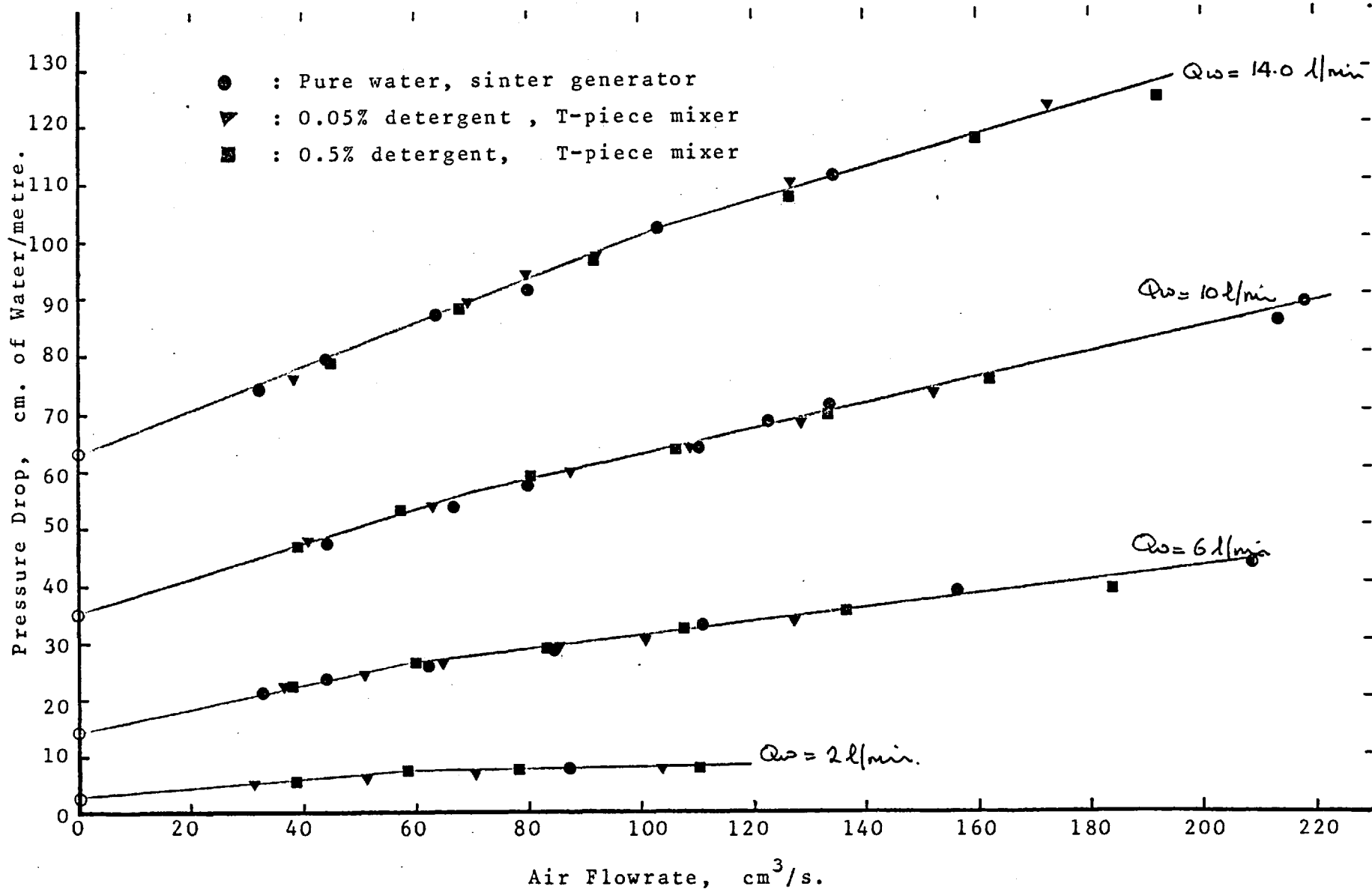
7.3) Pressure Drop for Air-detergent Solutions with Low Phase Interaction

The pressure drop, and flow regimes, obtained using the detergent solutions with the T-piece mixing, were identical to those obtained for the pure water-air flows.

Results for the flow of 0.05% and 0.5% detergent solutions in the 11.4 mm pipe are given in Figure 7.4, along with the results for the case of pure water at the same water flow rates. The data can be seen to fall on to identical curves, the error for these three different flow conditions being comparable to that obtained within any one set of data points.

The detergent solutions showed the pressure drop, at any liquid flow rate, going to that of water alone as $Q_A = 0$. The change in slope of the line on the transition to the slugging regime can again be noted, this can be seen more clearly in the plots in the Appendix, where the data for each detergent concentration is given separately.

Figure 7.4: Pressure drop results for Detergent solutions in 11.4 mm. pipe, with the T-piece mixer.



7.4) Pressure Drop for Air-detergent Solutions with Sinter Generator

The pressure drop results for foams generated by the Sinter indicate a complex behaviour. The flow regime depends on the phase velocities, which for the Sinter generator foams include the stratified foam flow regime. Flow mechanisms in this regime are complex, depending to a major extent on the rheological properties of the upper foam layer (75, 76).

The foam in the pipe cannot be considered a homogeneous mixture, and the application of normal fluid equations to the flow is not strictly possible. Results indicate however that for any constant gas/detergent volume ratio the mixture behaves as a Bingham type fluid.

Pressure drop V'S mixture flow rate for flow in the 11.4, 17.3, 26.0 and 39.1 pipes with a 0.5% detergent solution are given in Figures 7.5, 7.6 and 7.7. The pipe size range, and equipment flow rate restrictions, mean not all regions were investigated, but comparison of results indicate that the results for the 11.4 mm would apply for all the pipes at sufficiently large flow rates.

The data used to construct these charts was obtained from the original data by plotting ΔP V's Q_A at constant water rates. Straight lines result, with slope changes on flow regime transition. From these plots tables of ΔP at fixed Q_A/Q_W ratios were prepared. These plots and tables are given in the Appendix, as is the basic data.

The case of foam flow with a 0.05% detergent solution, in the 11.4 mm. pipe, is given in Figure 7.8. The curves show many similarities, the Bingham properties being present.

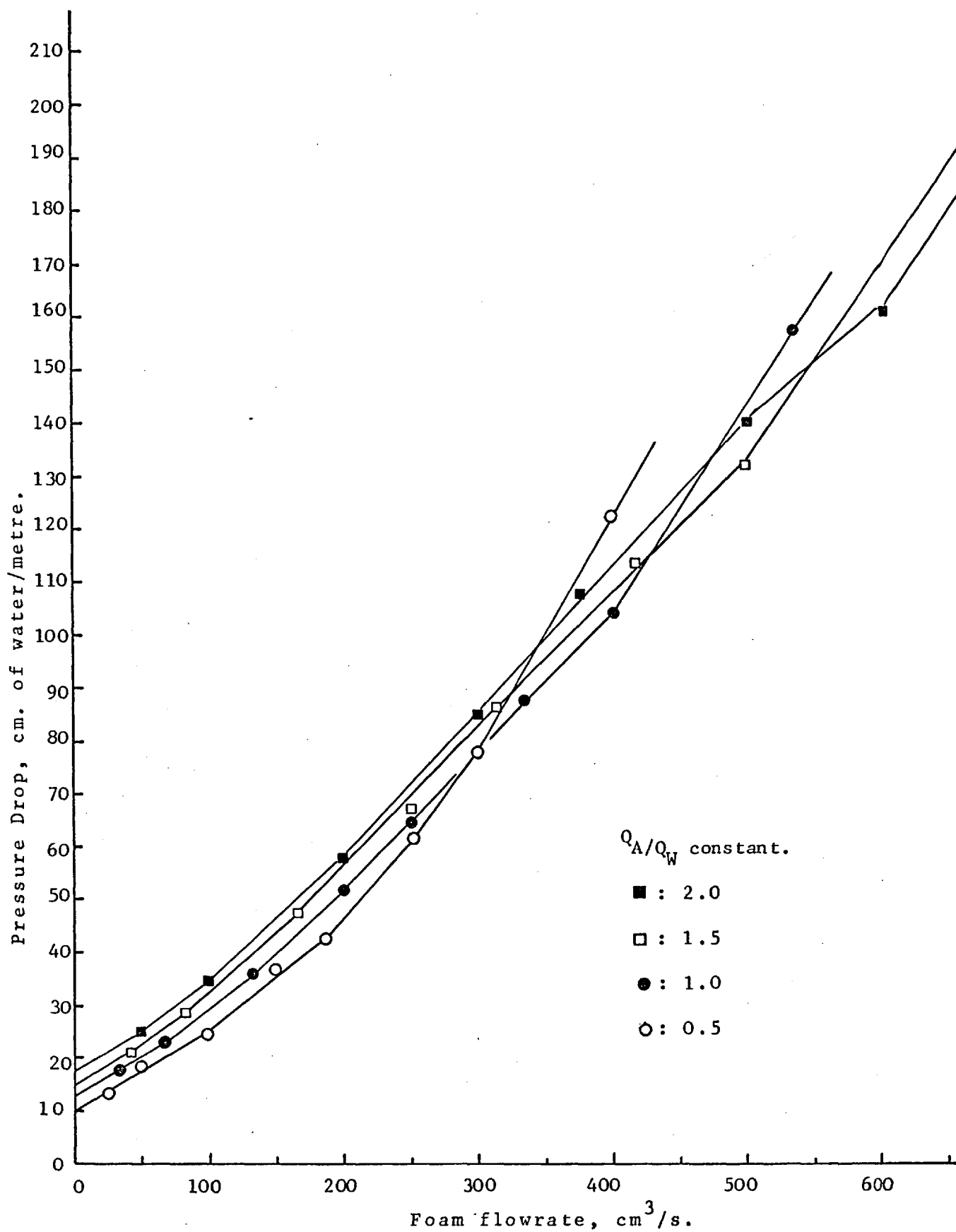


Figure 7.5: Pressure drop v's Foam flowrate in the 11.4 mm. pipe.

Sinter generator, 0.5% Teepol solution.

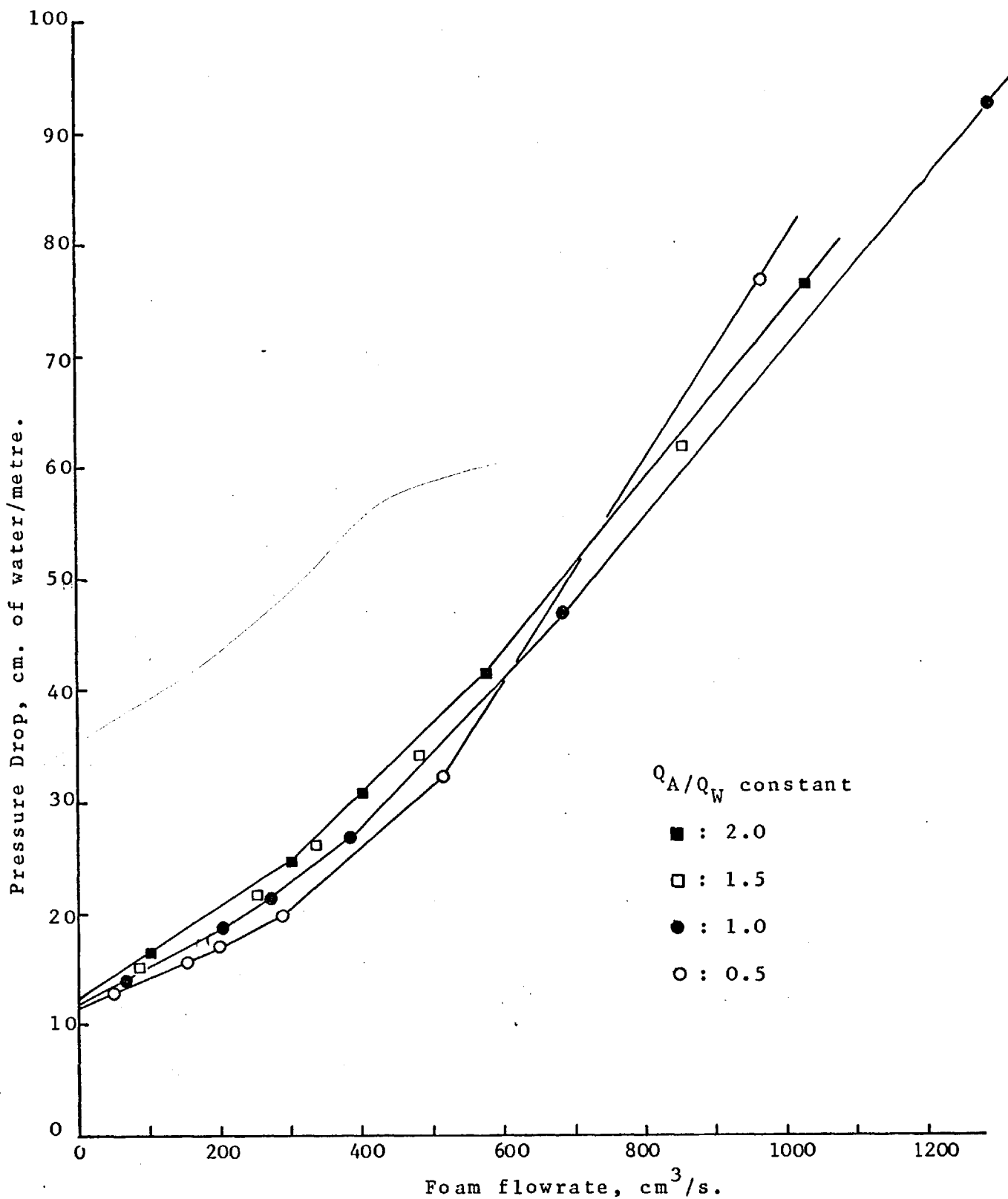


Figure 7.6: Pressure drop v's Foam flowrate in the 17.3 mm. pipe.
Sinter generator, 0.5% Teepol solution.

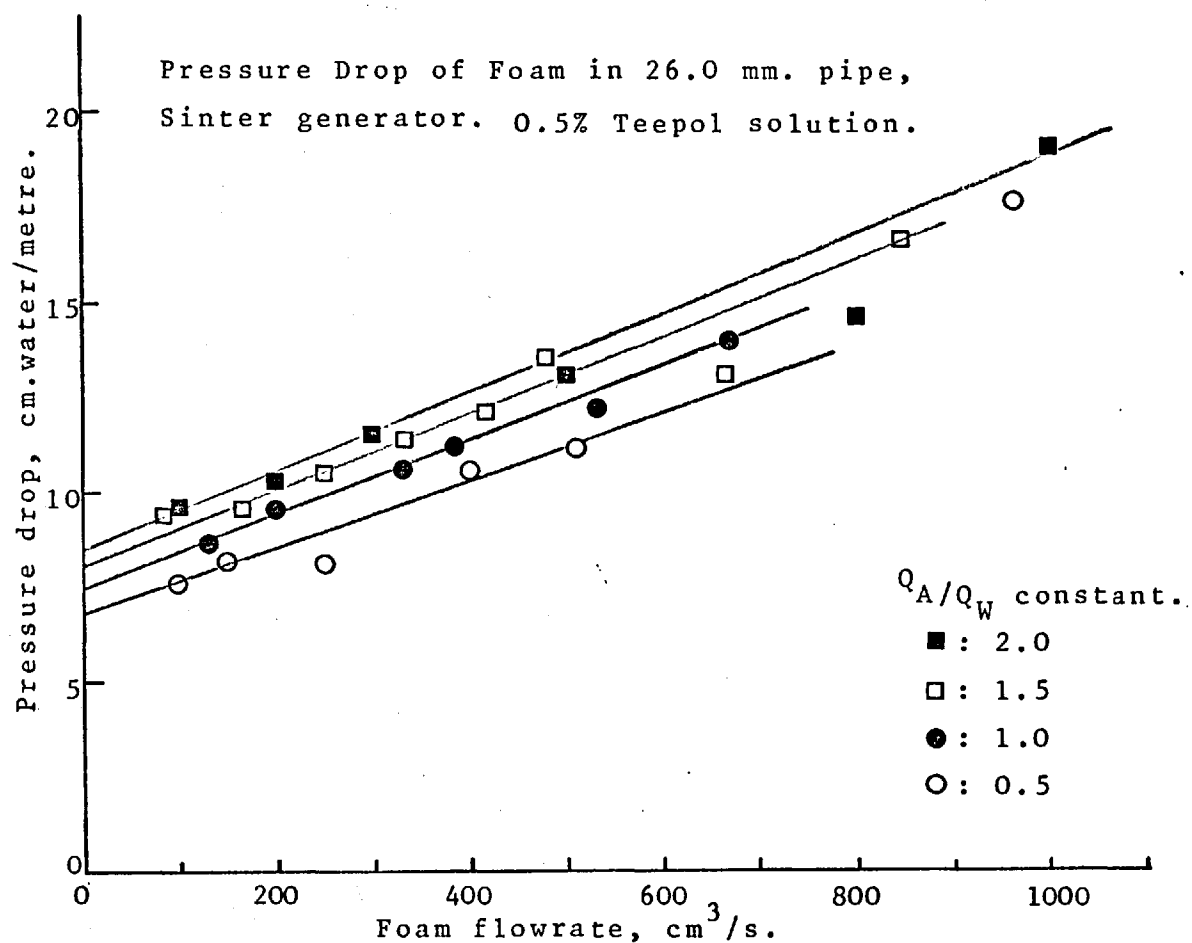
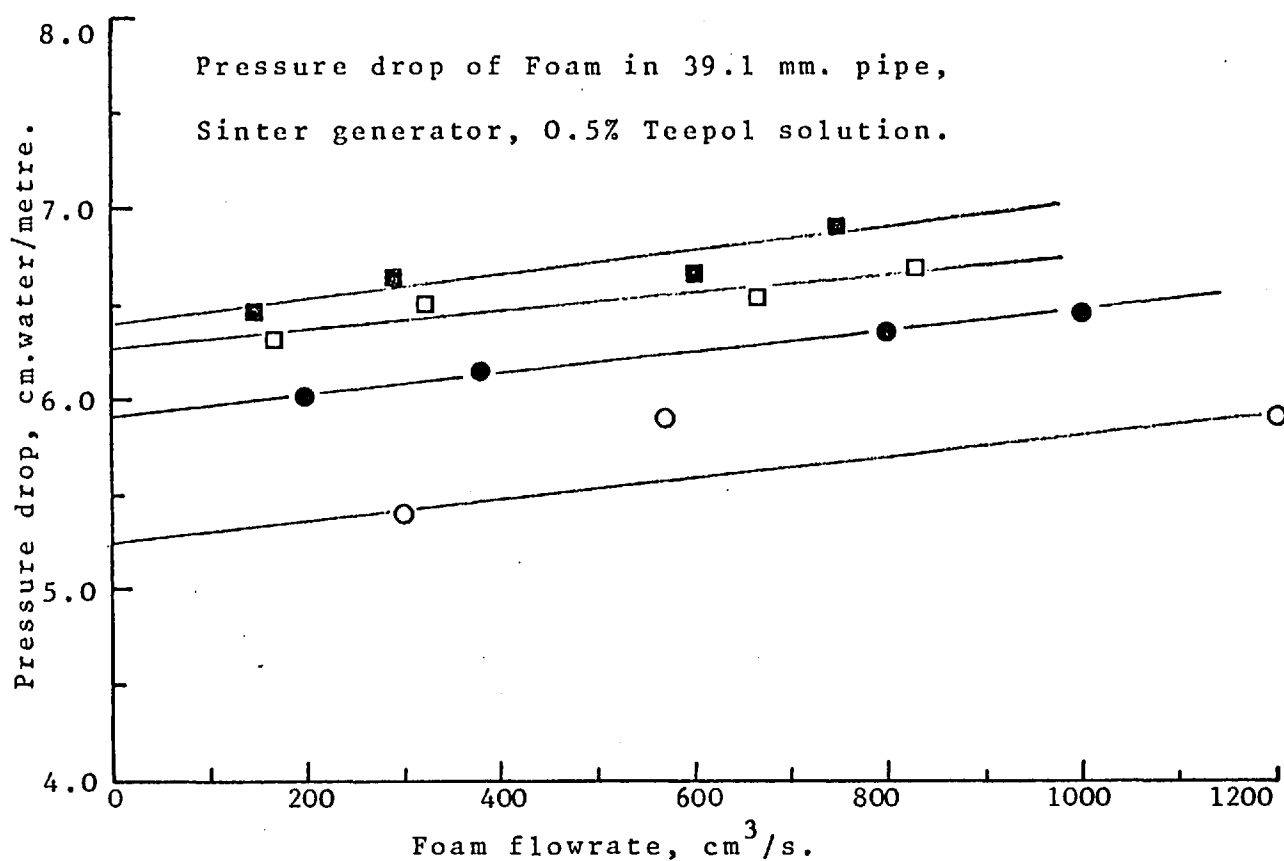


Figure 7.7: Pressure drop v's Foam flowrate in 39.1 and 26.0 mm. pipes. Sinter generator, 0.5% Teepol.

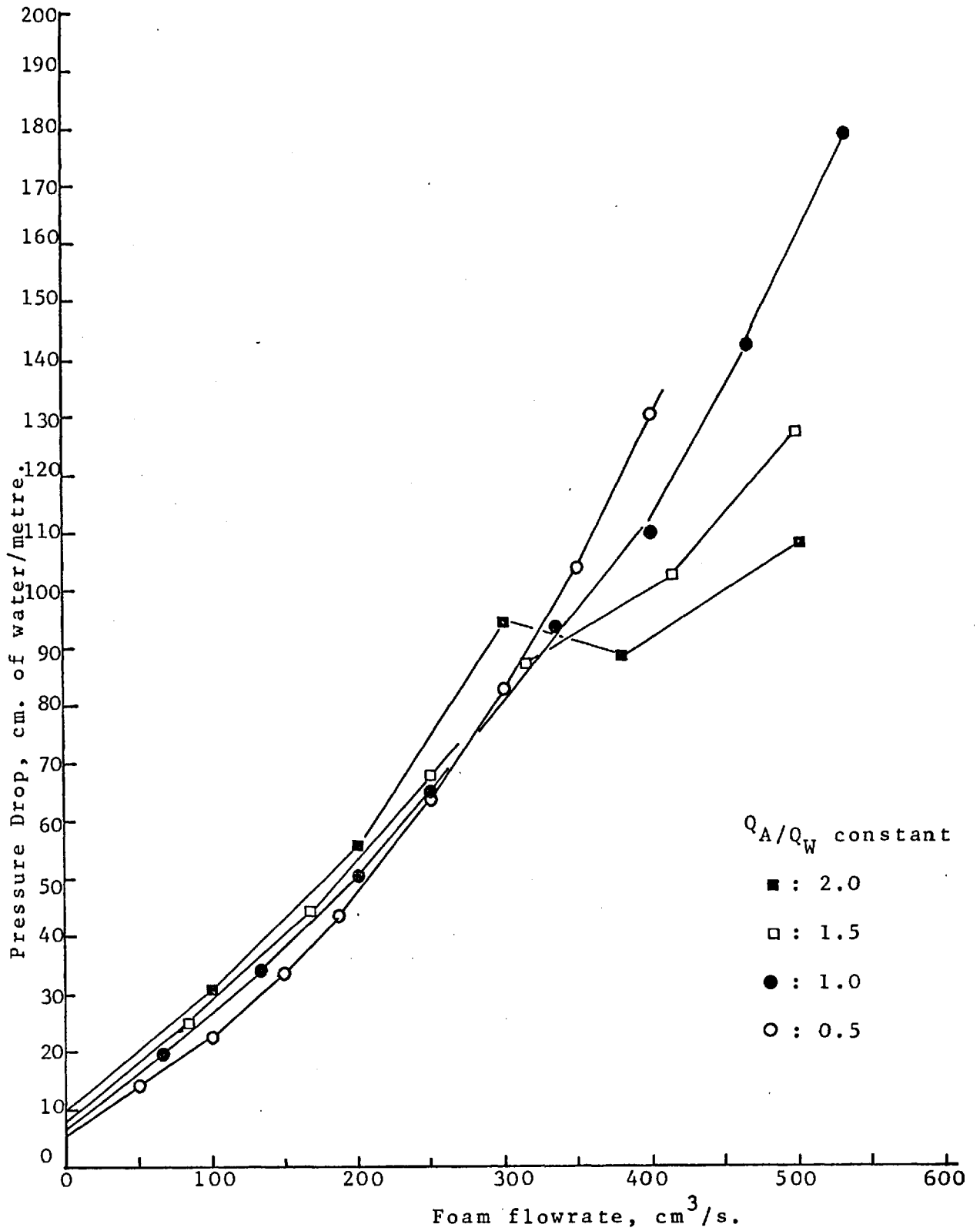


Figure 7.8: Pressure drop v's Foam flowrate in the 11.4 mm. pipe.

Sinter generator, 0.05% Teepol solution.

7.5) Pressure Drop for Air-detergent Solutions with Pump Generator

Results for the pressure drop V'S phase flow rates were collected for the case of foams produced by the pump generator, these foams were always in the dispersed phase. The results of the experiments at constant water flow rate and varying air rate are given in the Appendix. Charts constructed from this data, for flows of constant air-liquid ratio, are shown in Figures 7.9 and 7.10, for flow of the 0.5% solution in the 11.4 and 17.3 mm pipes. No experiments were carried out with the larger pipe sizes using the pump generator.

The plots show the same type of rheological behaviour that was obtained with the Sinter generator. The fluid has a yield stress, and the foams of various densities indicate the same behaviour. Data for the 11.4 mm pipe falls into a regular pattern, that for the 17.3 mm pipe is however not so uniform. Experiments with the pump generator with the larger size pipes was impossible, the flow regime which was obtained was a static upper dry foam layer with the foam produced by the pump flowing as a lower layer, in a manner similar to that met in the stratified foam flow regime, with the very fine bubbled foam taking the place of the water layer.

Additional experiments were carried out with 17.3 mm pipe by using a pair of pipes. The pump generator was set to produce a constant quality foam at a high flow rate, and this flow was then divided by a T-piece into two streams by suitable valving. The pressure drop in one of the lines was then measured. Foam composition was constant and no

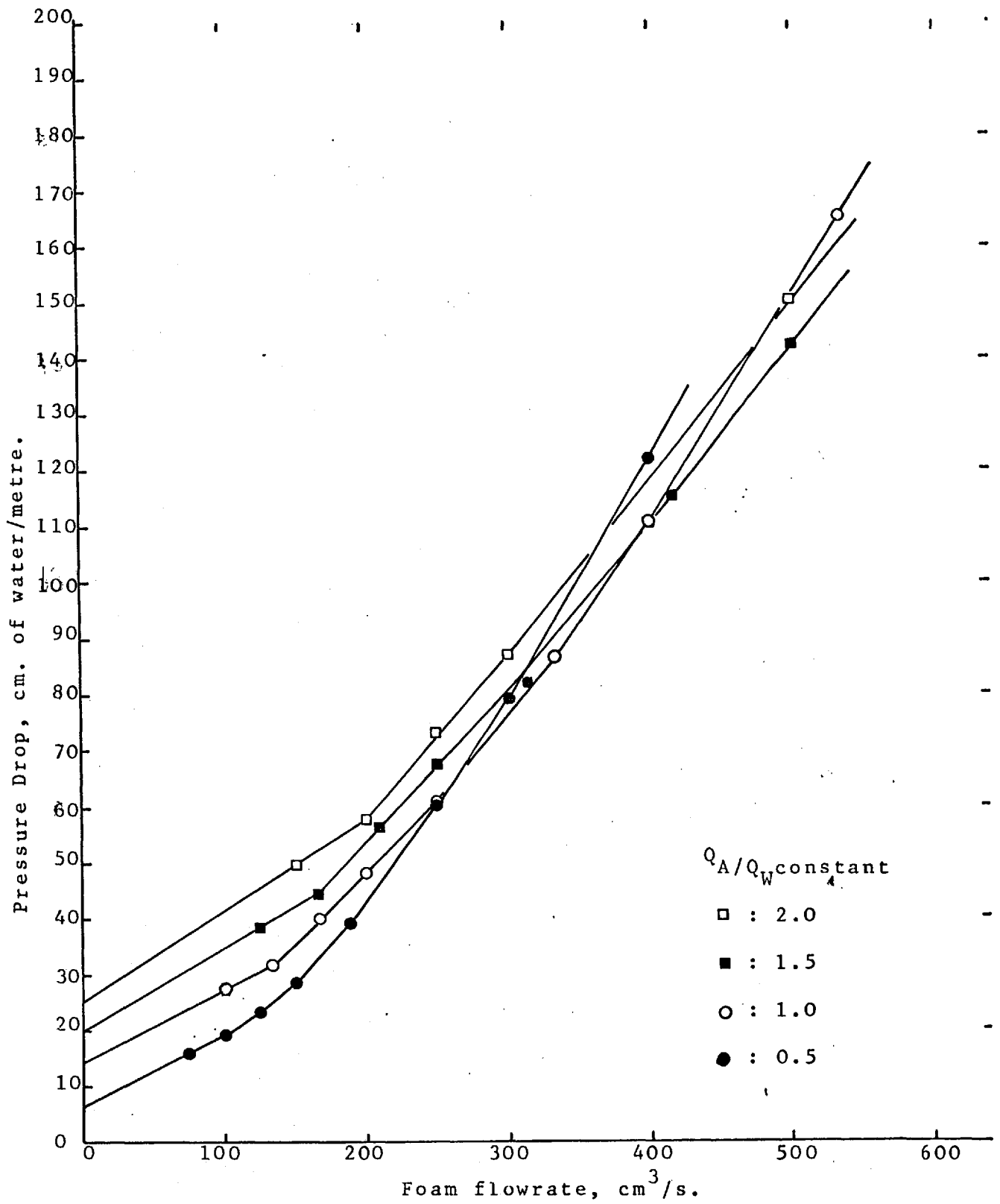


Figure 7.9: Pressure drop v's Foam flowrate in the 11.4 mm.pipe.
Pump generator, 0.5% Teepol solution.

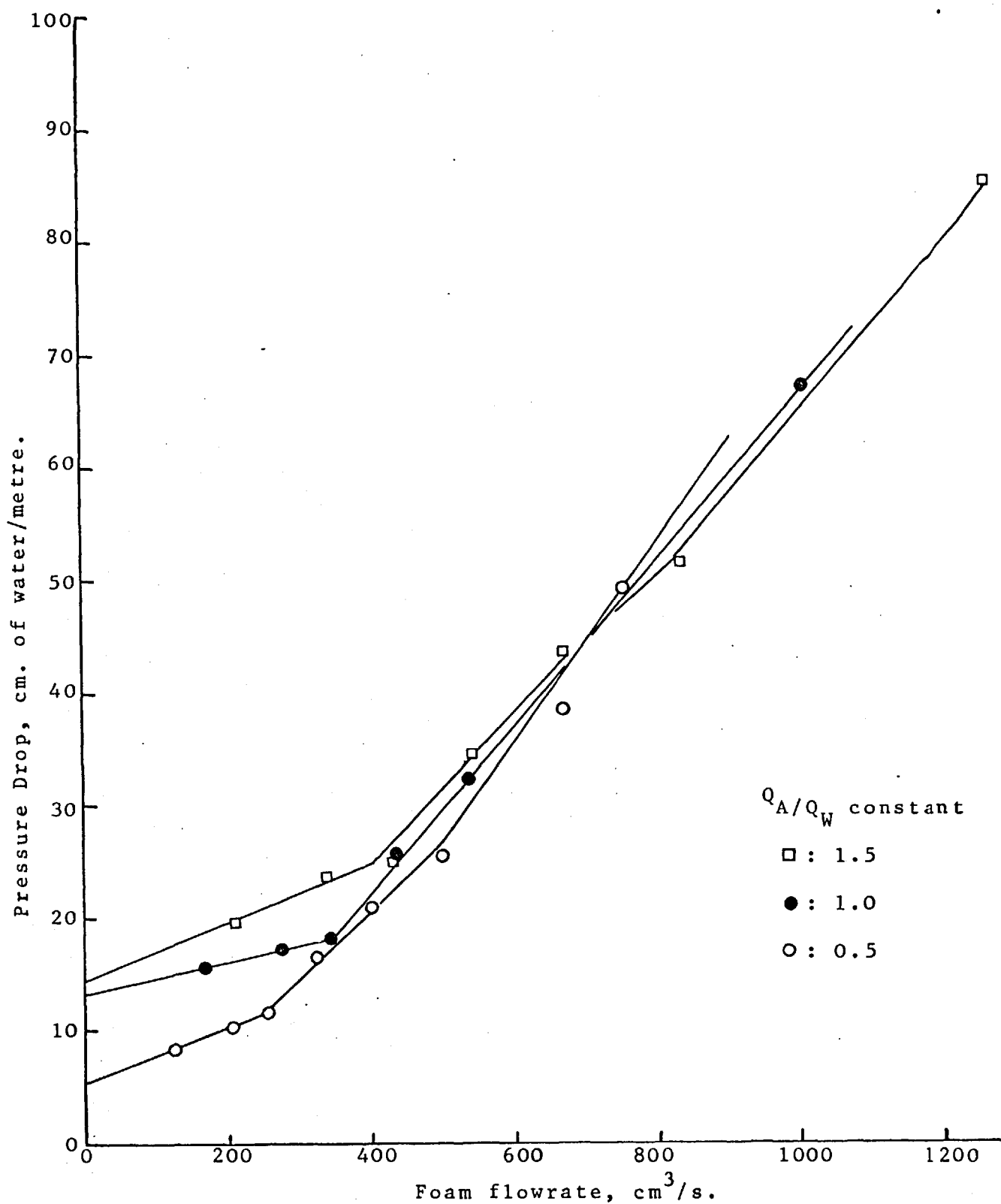


Figure 7.10: Pressure drop v's Foam flowrate in the 17.3 mm. pipe.
P mp generator, 0.5% solution.

intermediate charts had to be constructed. Every foam flow rate was sampled to measure gas/liquid ratio, although reported work on division of pseudo-homogeneous two-phase flows had indicated that no phase splitting at the flow division point would occur (83). Results are shown in Figure 7.11, where the Bingham fluid characteristics are more clearly brought out.

Results for flow with the 0.05% detergent solutions were similar, slightly lower yield stresses being noted. The plots are not given, inspection of the Lockhart-Martinelli correlation however show the trend.

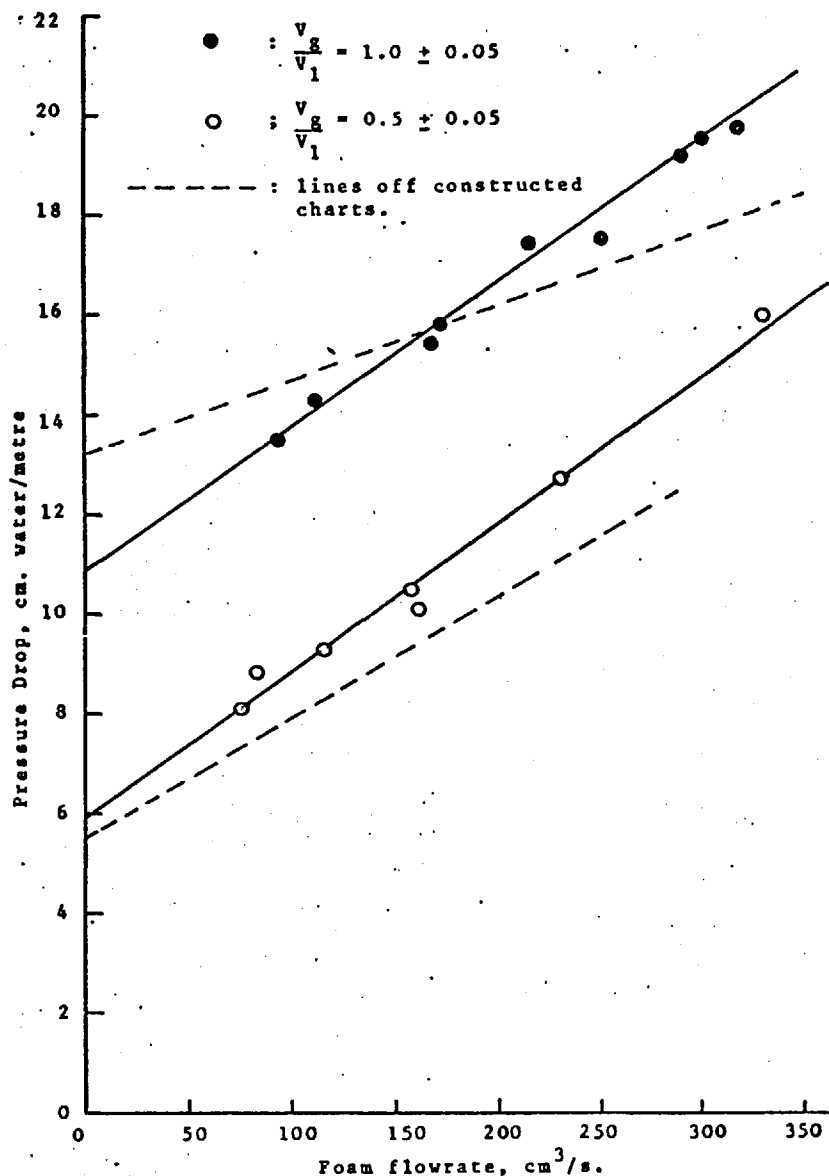


Figure 7.11: Pressure drop v's foam flowrate, using flow division of the main flow, for the 17.3 mm. pipe, pump generator.

7.6) Comparison of Pressure Drops

The pressure drop V 'S flow rate curves for all the detergent solutions with both pump and Sinter generators are similar, for flows with the same air/liquid volume ratio. All show a yield stress at zero flow, the value of the initial yield stress being dependent on the air/liquid ratio, the greater the amount of air the higher the yield stress. Curves for the similar conditions are in many cases coincident within experimental errors.

At high flow rates, where the flow regime was slugging in the case of pure water and the 0.5% solution with Sinter, and dispersed for the pump, the curves for all three flow conditions are approximately identical. This fact is further indicated in the Lockhart-Martinelli correlation for the flows. This behaviour can be seen on Figure 7.12 for flows at constant Q_A/Q_W in the 11.4 mm pipe. Similar curves for each Q_A/Q_W could be constructed, and would all show the same general behaviour.

The presence of detergent hence causes a marked change at low flow rates, imparting Bingham fluid properties to the foam. At high flow rates, when turbulence and slugging forces predominate, the surface forces play a smaller part in the flow regime mechanism, and results tend to those of air/pure water flows.

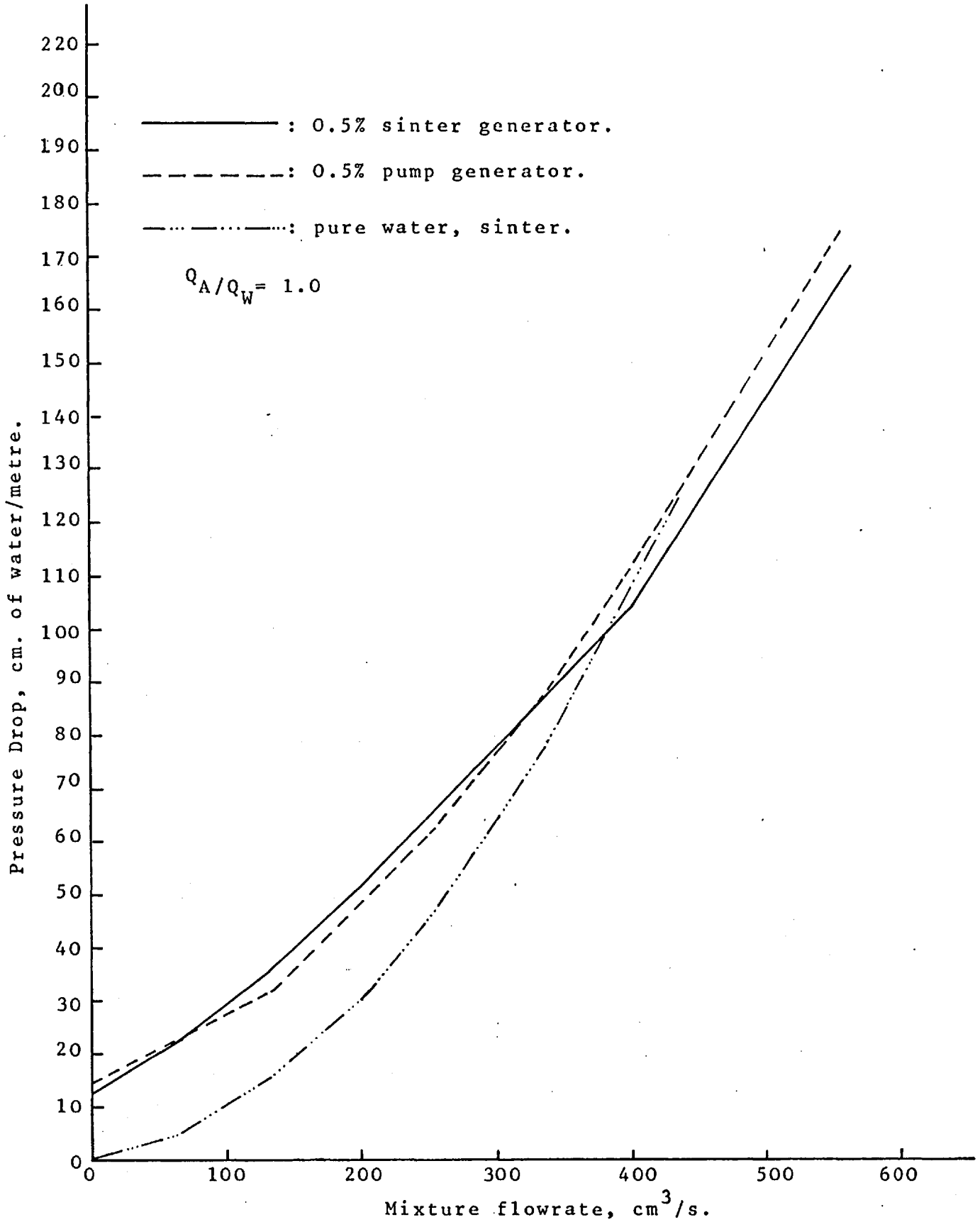


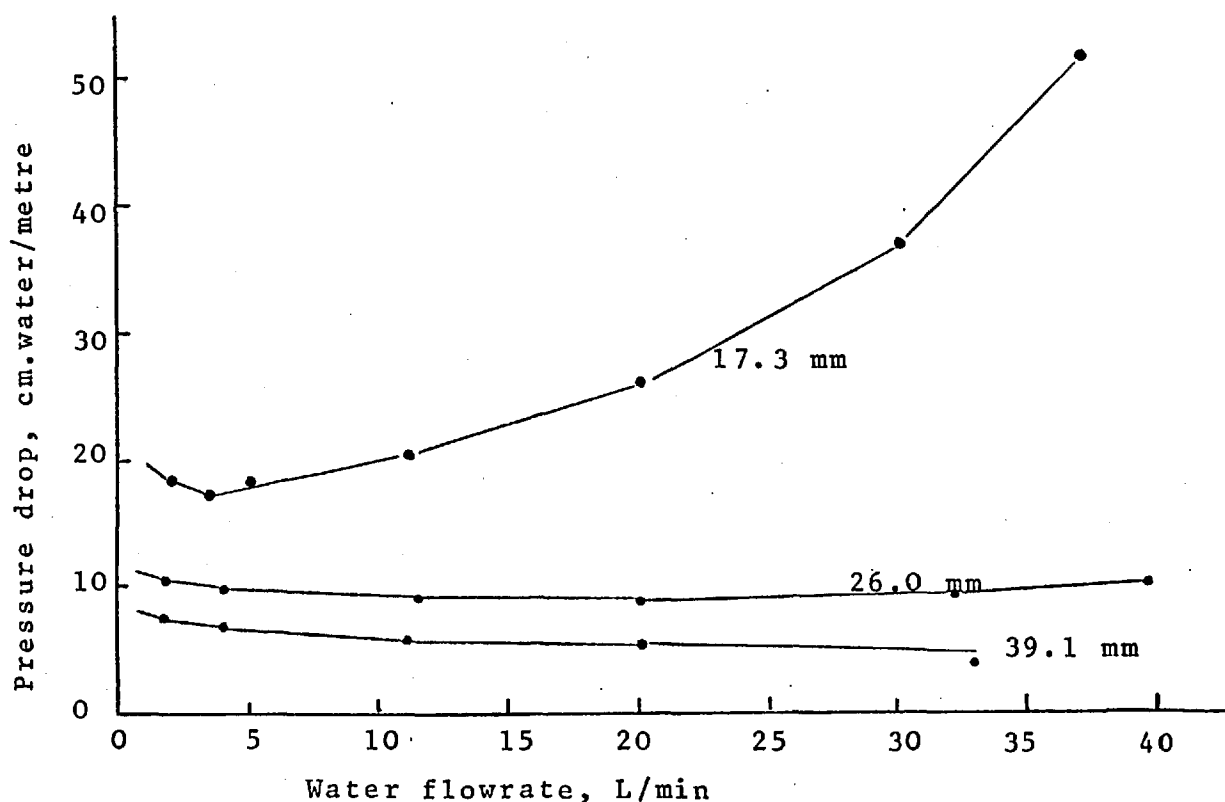
Figure 7.12: Comparison of Pressure drop v's flowrate for constant Air/water ratio, $Q_A/Q_W=1.0$, in the 11.4 mm. pipe, with pure water and 0.5% solution, sinter and pump generators.

7.7) Pressure Drop reduction at constant Air flowrate.

There was an apparent pressure drop reduction noted for certain flow conditions with the sinter generated foams.

At any fixed air flowrate, with increasing water rate, the pressure drop was found to pass through a minimum value. This effect can be seen in Figure 7.13, for a $Q_A = 6.0$ L/min. Results for the 17.3 and 26.0 mm pipes show the minimum, the water flowrate was not sufficient for the minimum to be reached in the 39.1 mm pipe. Similar curves can be drawn at different Q_A .

The model used to explain this flow behaviour was that of Stratified flow of a viscous liquid and water (75.76), and the flow mechanism is more fully discussed in section 9.



Air flowrate constant in all pipes, 6.0 L/min.

Flow regime in the stratified foam pattern.

Figure 7.13: Pressure drop reduction in Sinter generated foam, at constant air flow rate.

7.8) Bingham Model Parameters

The value of the parameters in the Bingham fluid equation, τ_o and μ_a , were calculated for the different flow conditions investigated.

$$\tau_o = \left(\frac{\Delta P}{L}\right)_o \cdot \frac{a}{2}$$

and

$$\mu_a = \left[\left(\frac{\Delta P}{L}\right)_Q - \left(\frac{\Delta P}{L}\right)_o \right] \frac{\pi a^4}{8 Q_f}$$

Values of $\left(\frac{\Delta P}{L}\right)_o$, etc, were read off the relevant charts. The results are given in Table 7.1.

Table 7.1; Bingham model paramaters for the Foam flows. Values of yield stress and apparent viscosity at constant Q_A/Q_W .

$\frac{Q_A}{Q_W}$	39.1 - sinter		26.0 - sinter		17.3 - sinter		11.4 - sinter	
	τ_o	μ_a	τ_o	μ_a	τ_o	μ_a	τ_o	μ_a
0.5	5.05	.032	4.35	.095	4.90	.045	2.86	.060
1.0	5.69	.033	4.74	.108	5.07	.058	3.62	.063
1.5	6.03	.026	5.18	.108	5.15	.080	4.21	.066
2.0	6.46	.035	5.44	.113	5.28	.091	4.91	.068

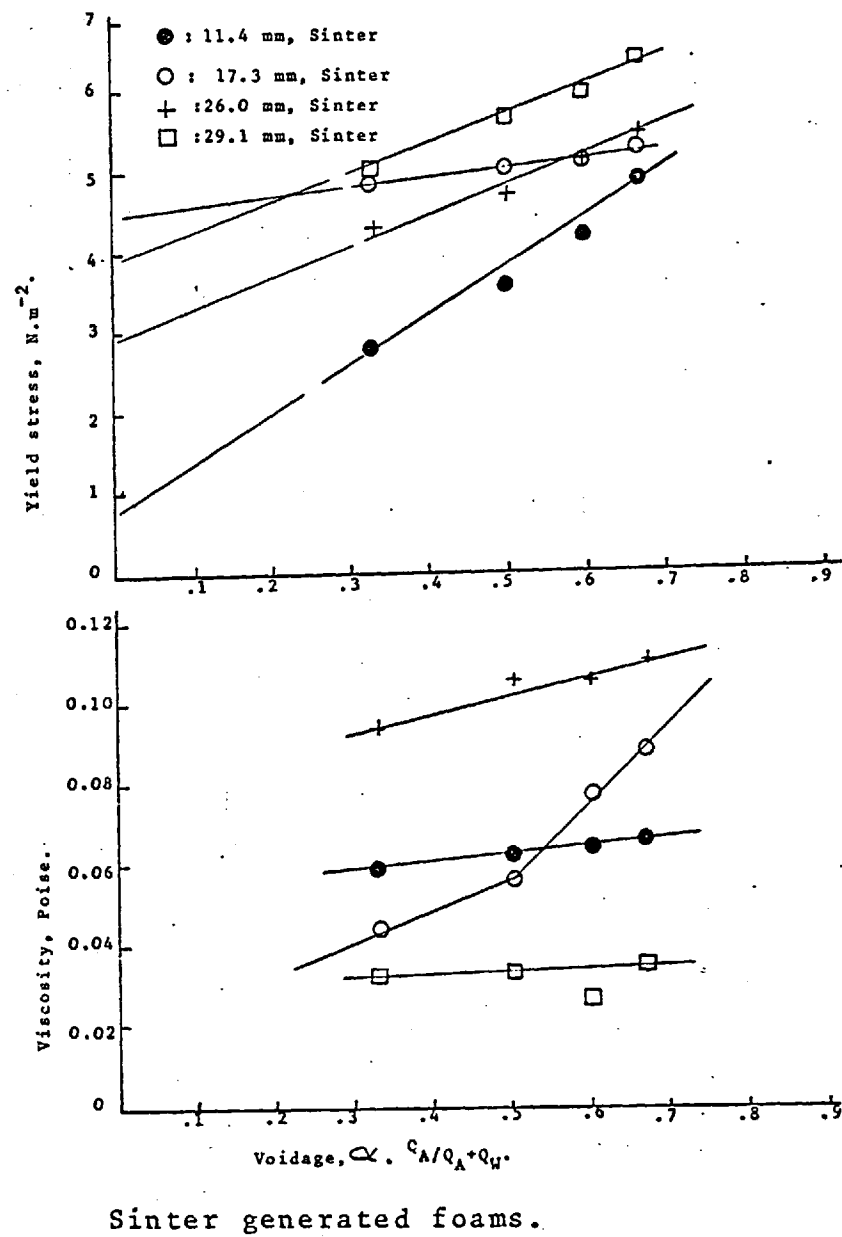
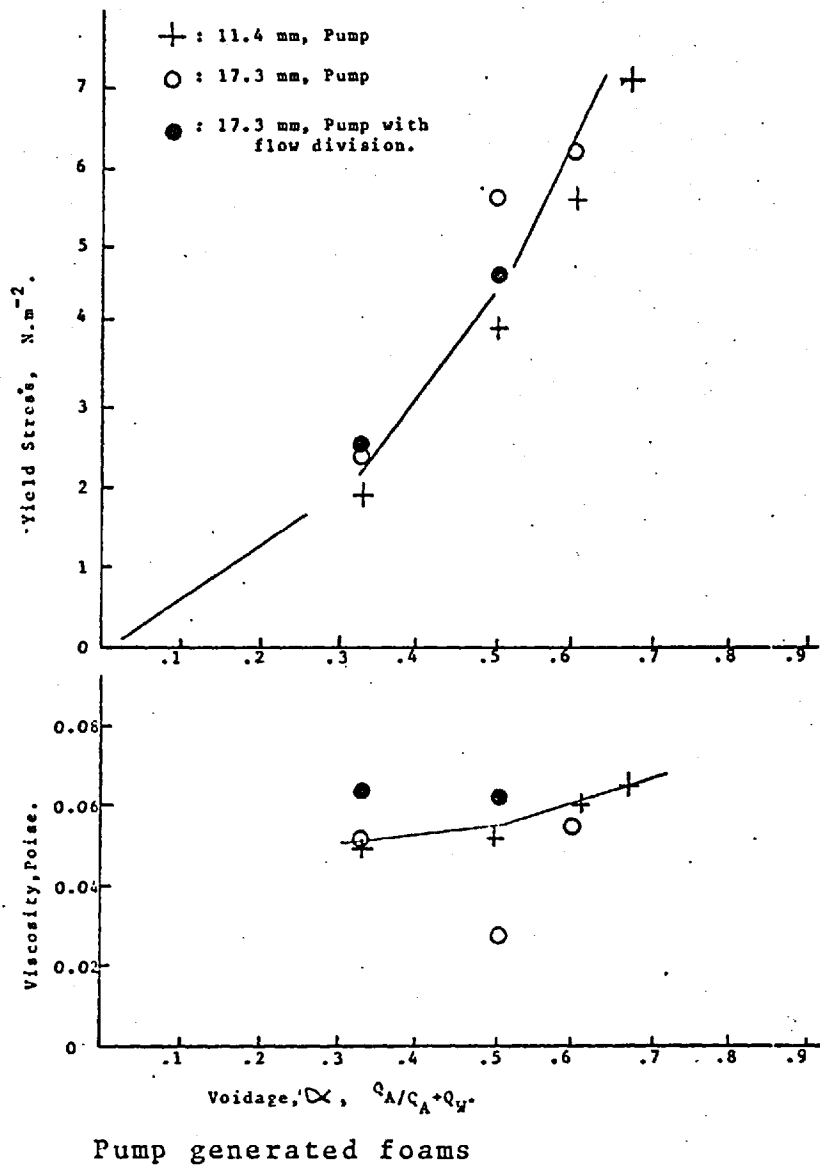
$\frac{Q_A}{Q_W}$	11.4 - pump		17.3 - pump		17.3 - pump flow division.	
	τ_o	μ_a	τ_o	μ_a	τ_o	μ_a
0.5	1.91	.050	2.34	.052	2.52	.064
1.0	3.98	.052	5.66	.029	4.65	.062
1.5	5.61	.061	6.17	.055		
2.0	7.15	.065				

Yield Stress, τ_o , Nm^{-2} . Apparent Viscosity, μ_a , poise.

These results are plotted in Figure 7.14, using the void fraction, $\alpha = Q_A/(Q_A + Q_W)$, to relate the factors.

The μ_a values for the foams in each pipe show an increase with increasing α , an average of approximately 20%, over the range investigated. The pump generated foam have a $\mu_a = 0.06$ poise, and both pipes give the same general result. The sinter generated foams show a much wider range of μ_a , from

Figure 7.14: Variation of Yield stress and Apparent viscosity with voidage, for the sinter and pump generated foams.



0.03 to 0.11 poise, indicating the complex behaviour of what is essentially two different liquids in stratified flow, the averaged properties of which depend on the flow geometry and pipe size.

The yield stress for the pump generated foam tends to zero as $\alpha \rightarrow 0$. This was as expected for a bubble dispersion, when voidage is low no interactions occur, and the yield stress properties of the fluid are lost. The sinter generated foams indicate that at low α the yield stress was not zero. The stratified structure always gave a foam layer, as α decreased this layer was smaller, but as long as sufficient bubbles were present to form a layer, the yield stress properties of the flow would persist.

These results, coupled with the regime maps, show that the pump generated foams can be considered a good approximation to a Bingham fluid, where τ_0 was a function of voidage, and μ_0 being approximately 6 centipoise in the range, $0.3 < \alpha < 0.7$. Results for the two pipes show an identical behaviour, little effect of pipe size on the flow properties being noted.

The sinter generated foams however had a far more complex flow structure. The lumping of parameters into an average value gives poor results, the flow behaviour of the liquid and foam layers needing to be analysed separately to give meaningful overall results.

7.9) Friction Factor Correlation

The results of the flow experiments in the pipes with 0.5% detergent solution were correlated by use of a friction factor/ Reynolds number plot.

$$\text{where } f = \frac{\tau_w}{\frac{1}{2} \rho_f u_f^2} \quad \text{and} \quad Re = \frac{u_f \rho_f D}{\mu_a}$$

Data from the charts of constant air/detergent solution flows were used to give pressure drop and flow rate information. The value used for μ_a was that calculated for the relevant flow for the Bingham model.

The results for the foams produced by the pump are shown in Figure 7.15. Lines for a range of Q_A/Q_W ratios are shown.

The results for the foams produced by the Sinter are shown in Figure 7.16. Lines for all the pipes are given for $Q_A/Q_W = 1$, other lines at different Q_A/Q_W values would give equivalent results.

Indicated on each line is the Hedstrom number for that particular flow,

$$He = \frac{\tau_o^2 D^2 \rho_f}{\mu_a^2}$$

The friction factor plot shows the typical result obtained with Bingham fluids (59). The lines are ordered in relation to their Hedstrom numbers. At the high Reynolds numbers the curves are tending to a single line, below the line for single phase flows. This is most obvious for pump generated foams, the values of He being in the range 1.6 to 6×10^3 . Other workers report this line to be parallel with the single phase curve, insufficient data at higher Re meant that nothing can be deduced in this case.

Of note is the result for the Sinter produced foams in regard to He number for the various pipe flows. An increase of He with diameter might be expected to be the rule for this

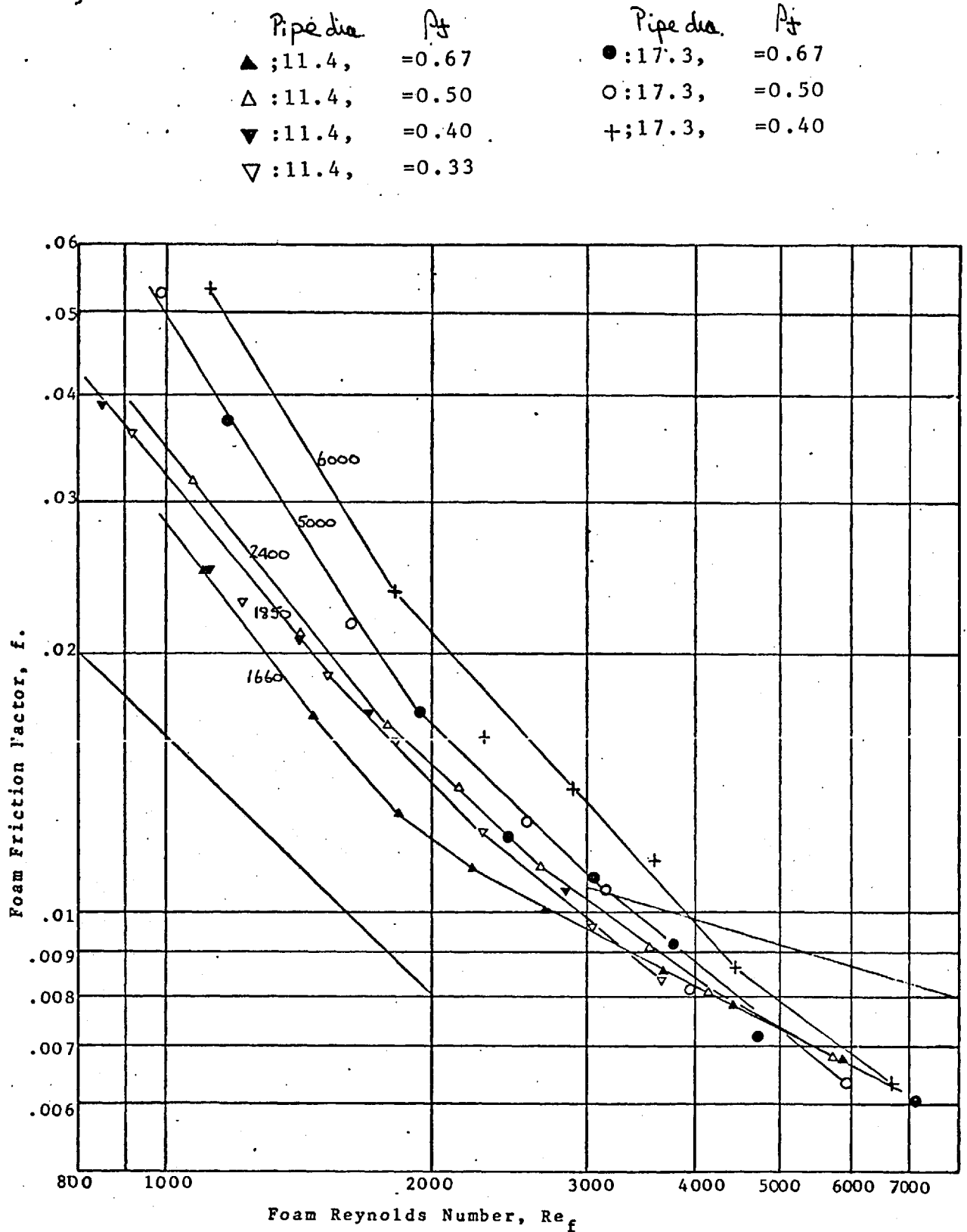


Figure 7.15; Friction factor/Reynolds number plot for the pump generated foams, for flow in the 11.4 and 17.3 mm pipes, at a range of Q_A/Q_W values, 0.5% Teepol solution.

●:11.4, ○:17.3, ▼:26.0, ▽:39.1mm pipes.

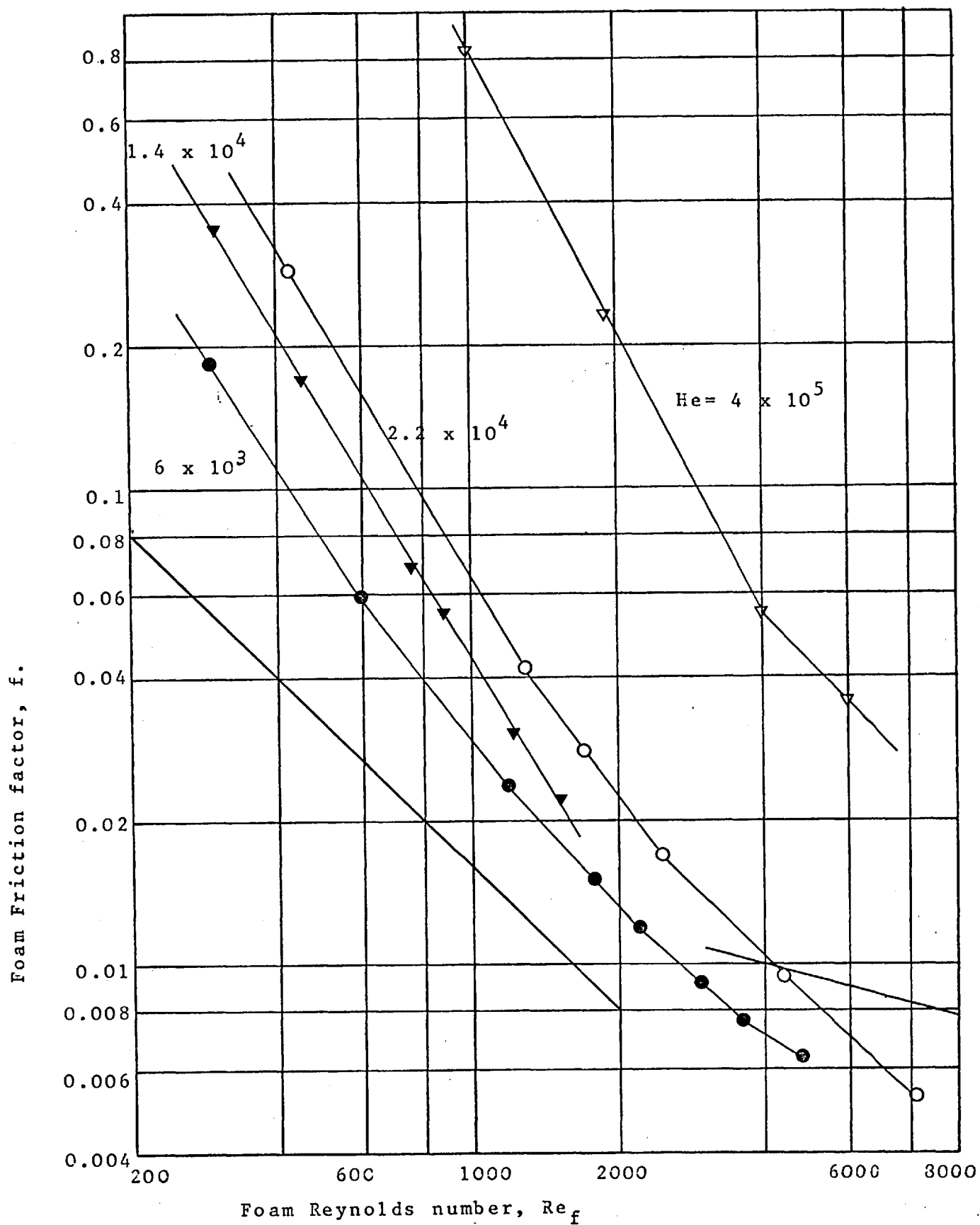


Figure 7.16: Friction factor/Reynolds number plot for the sinter generated foams, in the 11.4, 17.3, 26.0 and 39.1 mm pipes, at a $Q_A/Q_W = 1.0$, 0.5% Teepol solution.

foam, but the value of He for the pipes are

D .mm	He
39.1	4×10^5
17.3	2.2×10^4
26.0	1.4×10^4
11.4	6×10^3

The values of f v's Re however are such to follow the usual pattern required for the He value.

7.10) Slip at Wall Investigation

The pump generated foam was a homogeneous mixture, and the theory developed in Section 3.1 could be expected to apply. Wall slip has been reported for foams, but not as yet quantified.

The theory as postulated is applicable only for laminar flow. The friction factor, Reynolds number correlations indicates that only a few data points are available in the true laminar region, the bulk of the data being collected in the intermediate and turbulent region. This aspect will always be present in foam studies, for the restrictions previously mentioned.

Data at $Q_A/Q_W = 1$ and 1.5 for pump generated foams in the 11.4 and 17.3 mm pipes has been used to prepare a plot of τ_w v's $Q/\pi a^3$.

Theory indicates that for each fluid the data from the different pipe sizes should fall on the same curve. Results are given in Figure 7.17, where it can be seen that for each fluid the shear stress at low shear rates tends to the same line. At high shear rates the curves diverge, and the results for each pipe tend to fall on to the same curve, notwithstanding the differing Q_A/Q_W ratios investigated.

Hence experiments do not support the presence of wall slip in the laminar flow regime, but some effect of pipe was present at higher flows, which must be considered turbulent due to Re values. The use of the theory for slip is thus inapplicable, trials using the theory showing that values of Q_S could be obtained which were greater than Q_T .

Alternative theories to quantify and measure the slip are needed for foams, to take account of the turbulent foam flow, which will be the normal flow mode due to inherent restrictions which apply if a uniform foam system is to be maintained in the pipe.

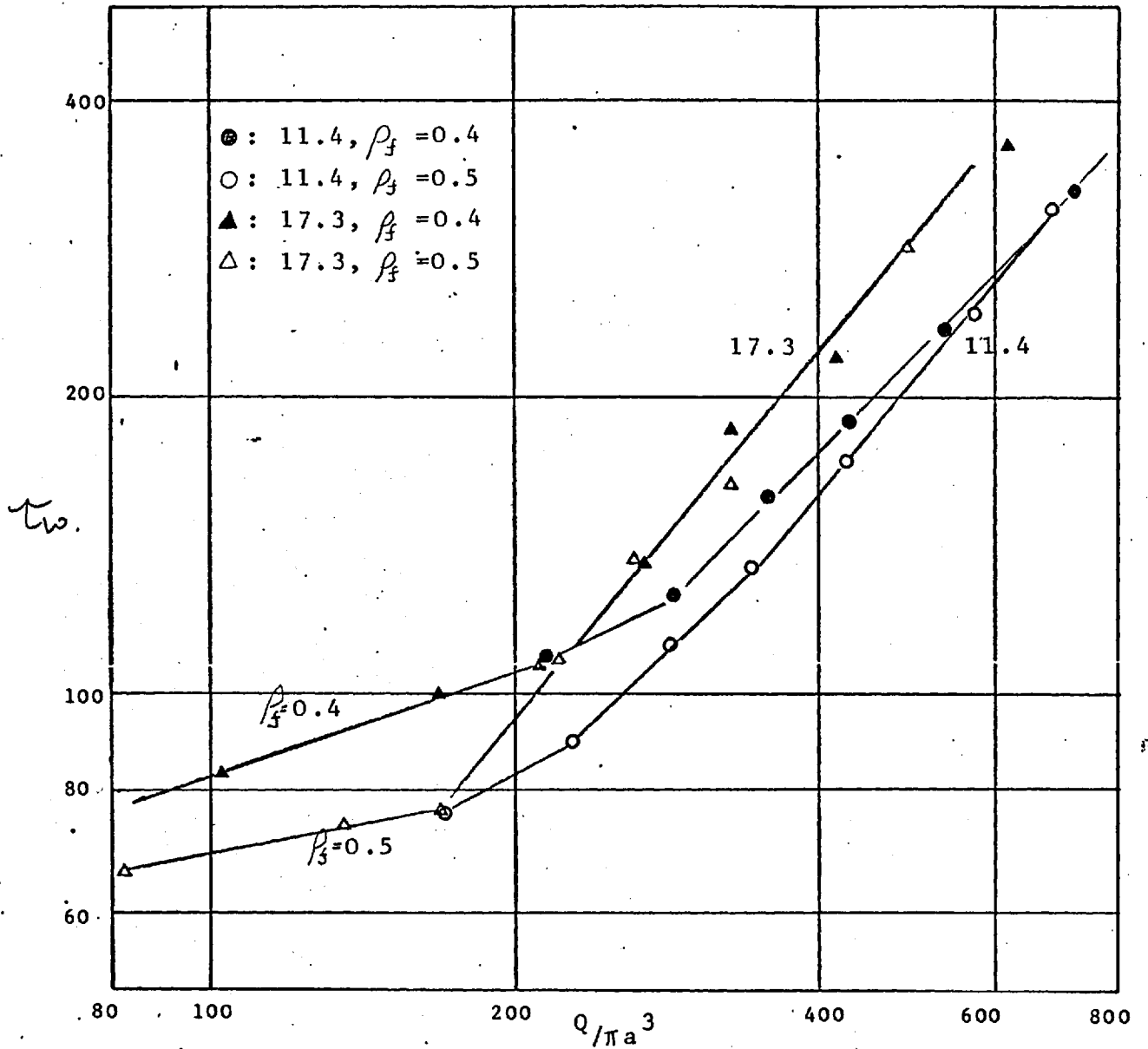


Figure 7.17: Plot of v' vs $Q/\pi a^3$, for the pump generated foams, 0.5% Teepol. Results for 11.4 and 17.3 mm pipes.

7.11) Foam Flow in Nozzle

Flow of two-phase mixtures in a nozzle is of an extremely complex nature. Most experimental investigations have used the dispersed flow regime, or some method of injecting the gas into the liquid stream just before the nozzle, so as to obtain an approximation to a homogeneous mixture. Equations describing such a flow system have been developed, but the experimental problems and errors have made the confirmation of certain details of the model difficult. Measurement of the pressure drop, air/water ratios and velocities all being subject to a high degree of error, due to the presence of the two phases.

The foams produced in the pump generator have an extremely uniform structure, with a reasonable degree of control on phase ratios, and so would be ideal for use in two-phase nozzle experiments. The presence of the detergent might cause some changes, but surface forces are likely to play only a small part in the rapidly changing pressure field. Initial experiments with a hand drawn venturi nozzle showed that a shock wave, indicating 'sonic' flow, could be obtained in the expansion side of the nozzle. An accurate glass venturi nozzle, of a known size, was then made. Many problems were experienced in its manufacture, the final method used being given in section 3.7. Only a few cursory runs were made with the nozzle, due to time limitations.

Experiments were carried out with the nozzle in the test pipe, looking at the foam flow in the contracting region of the nozzle. Foam generated by the pump was used, and the pressure just before the nozzle, P_1 , and at the throat, P_2 , measured. At high air rates a discontinuity in the foam in

the expansion section after the throat was observed. This 'shock front' was very distinct, oscillating back and forth over several millimetres, and spanned the pipe. The front would move towards and away from the throat as the air rate was lowered and raised.

The pressure drop for the foam flow in the contracting section was measured for a range of liquid water flowrates, with increasing air rates, and compared to the pressure drop for water flowing alone. The results are given in the Appendix, and the results of $\Delta P_f / \Delta P_L$ v's Q_A / Q_W shown in Figure 7.18. The presence of the gas can be seen to increase the pressure drop, compared to water alone. The equations describing the flow are complex, and are not developed in this work due to lack of experimental work to test them.

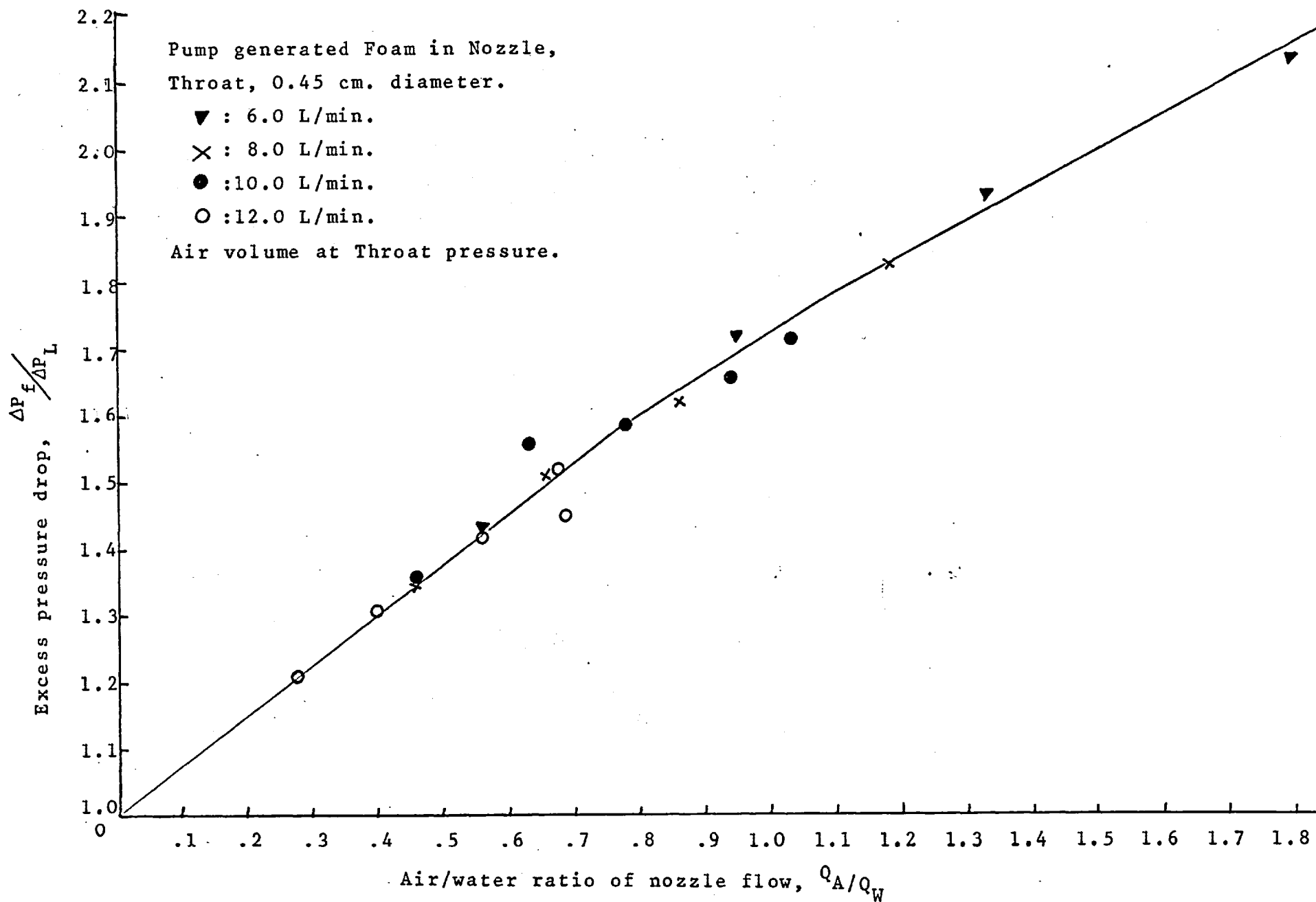
Inspection of the results indicate that there is a relationship between $\Delta P_f / \Delta P_L$ and Q_A / Q_W of the form,

$$\frac{\Delta P_f}{\Delta P_L} = A + B \cdot \frac{Q_A}{Q_W}$$

for $Q_A / Q_W < 0.8$; $A = 1.0$ and $B = 0.75$.

Visual indications of flow in the nozzle suggested sonic flow was reached in the nozzle. The velocity at the throat was calculated from the value of Q_{Ao} and P_2 and compared with the sonic velocity given by equation 2.89. Values of the Mach number of up to 0.98 were calculated for the throat, no flows which gave $M > 1$ were recorded. Results can be seen in the data in the Appendix. Sonic velocity at the throat was therefore reached, the error in the measurement of pressure and phase ratio and flow rate being such that error in $M_{\text{calculated}}$ would be about 5%. The equipment was limited in the maximum pressure which could be obtained, a

Figure 7.18: Pressure drop in the contracting section of the venturi nozzle, with pump generated foams.



more detailed experimental investigation would require that the equipment be modified to provide greater pressures.

These few experiments showed that foams are an ideal two-phase fluid for high speed nozzle investigation, the presence of the shock wave being easily maintained due to the high degree of homogeneity of the mixture.

8. Lockhart and Martinelli Correlation Results

8. 1) Water Alone

The results of the Lockhart Martinelli correlation for the cases with pure water-air flows are as shown in Figure 8.1 for the cases of

- 1) 11.4 mm pipe with sinter and pump generators
- 2) 17.3 mm pipe with sinter generator

The agreement with the original correlation can be seen to be excellent, a comparison with the scatter obtained in the original work can be seen in the figure below, taken from the paper(45). This close agreement is as might be expected as the phases, pipe size and material and phase velocities are all within the range used for the derivation of the correlation.

The tabulated results for these conditions are given in the Appendix, as well as more detailed plots of the charts, upon which is indicated the data grouped with the liquid flow rate indicated.

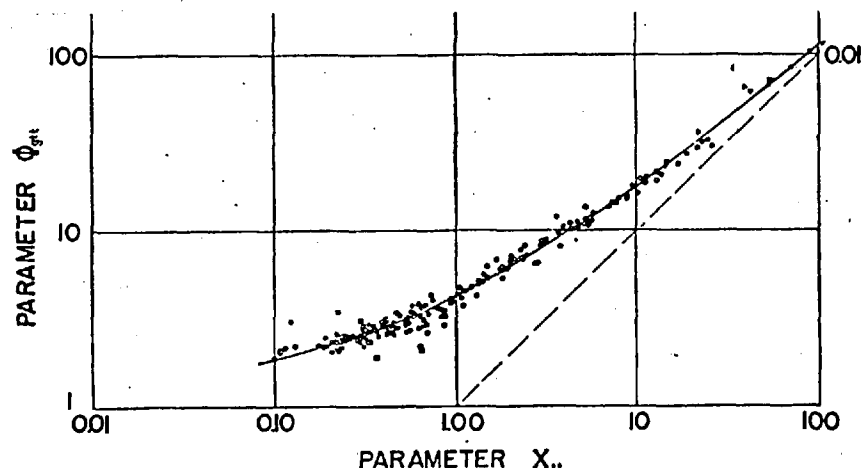


Fig. 1. Relation between R_L , ϕ_{gt} , and parameter X for mechanism 1, turbulent-turbulent flow. Experimental groups 3, 5, 8, 10, 13-17 are included. A few points from experimental groups 4, 7, and 9 for flow mechanism 3, turbulent-viscous flow are also plotted

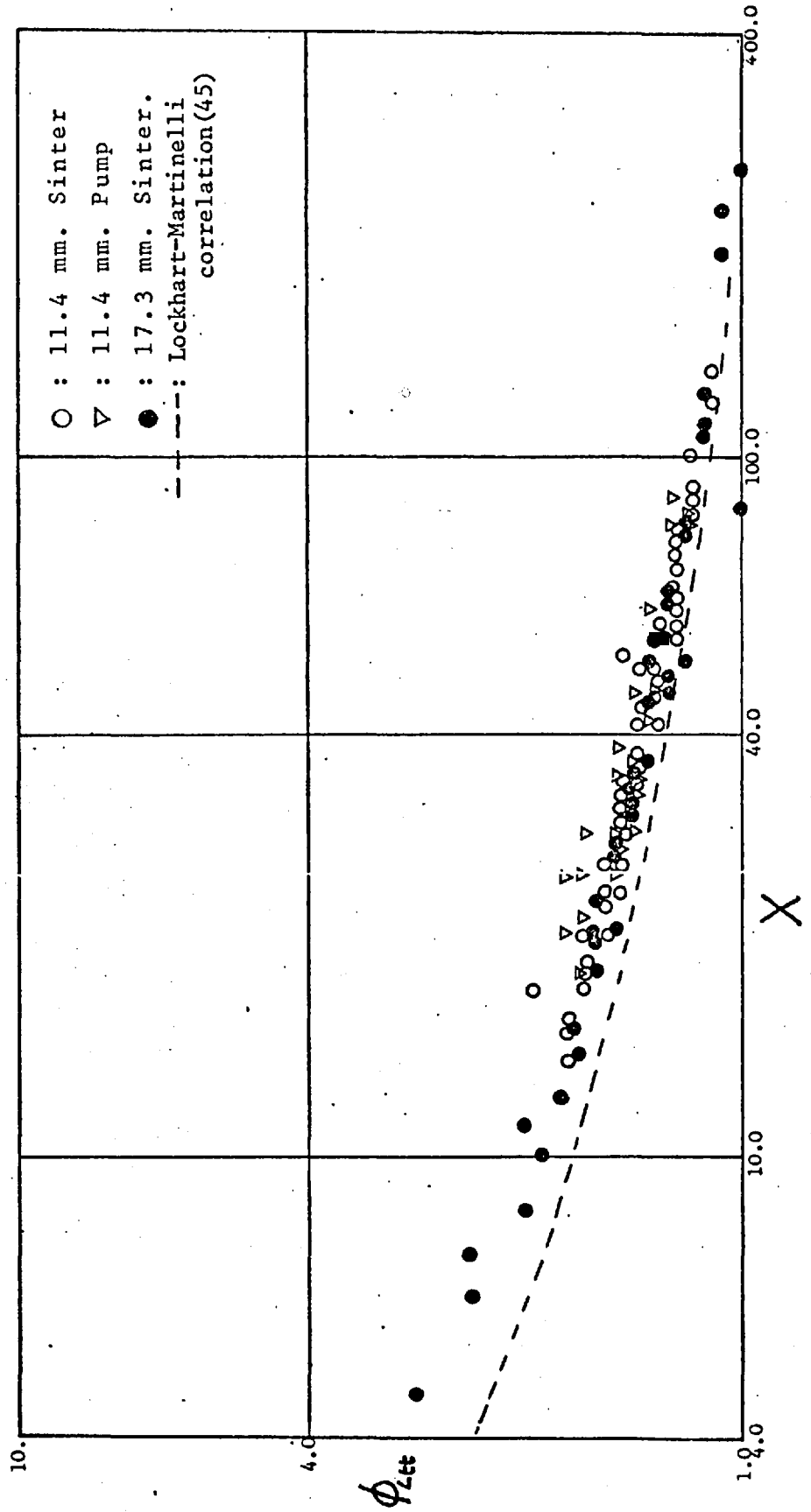


Figure 8.1: Lockhart-Martinelli correlation for the flows of pure water, in 17.3 and 11.4 mm. pipes

No difference can be noted on these charts for the various mass flow rates, other workers (46,47) on similar fluid systems have found that each particular mass flow rate can be best correlated with a single line, the average of which however is the Lockhart-Martinelli curve.

The lack of effect of a change in phase mixing method is shown by the comparison with the sinter and pump results for the 11.4 mm pipe. This is in accord with the findings of the flow regime pattern work which showed the phase mixing methods to have no effect on transition boundaries. The agreement shown in the pressure drop results also indicate this point.

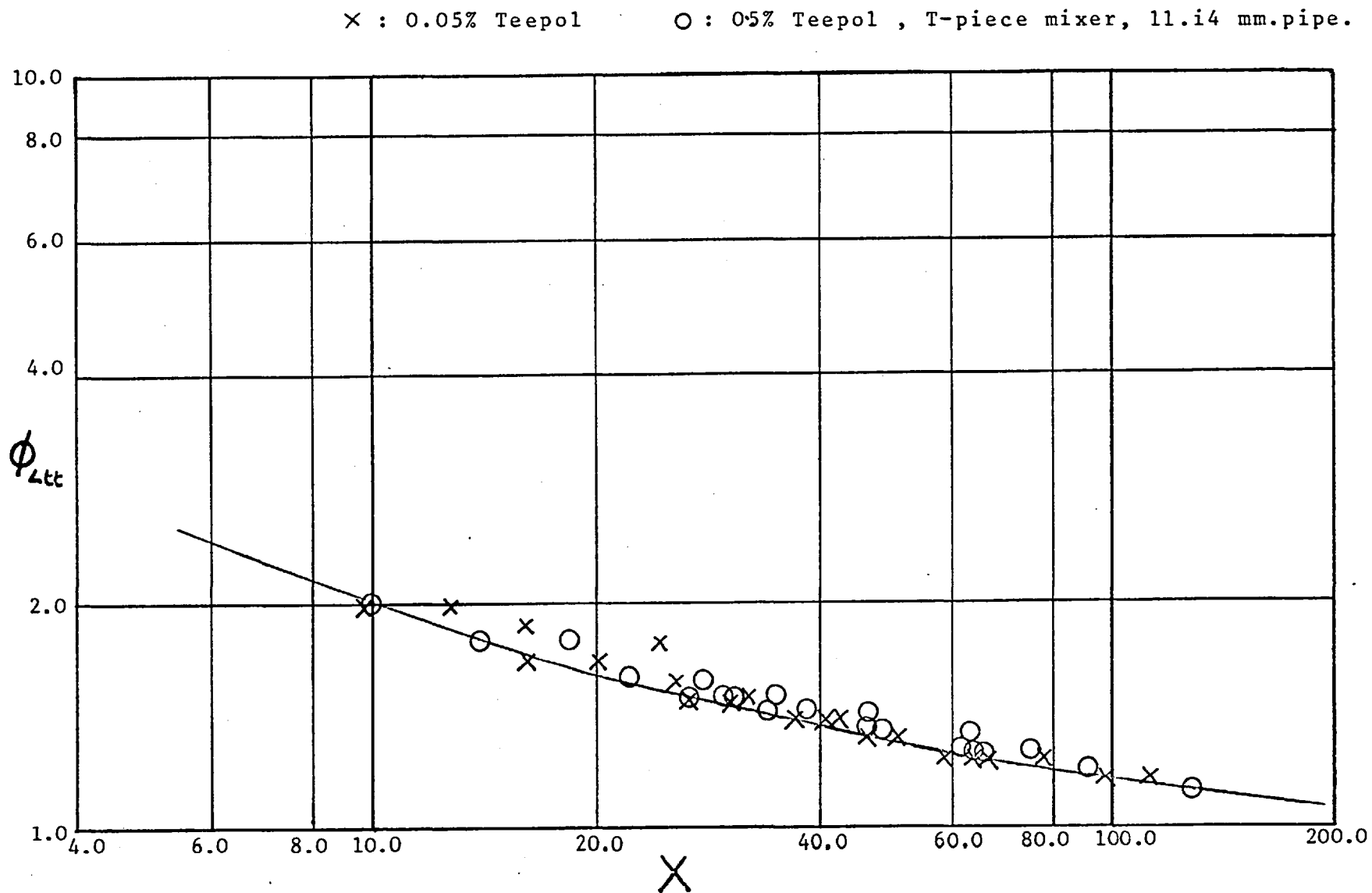
8.2) Detergent Solutions

The effect of detergent solutions of both the flow regime patterns and pressure drop results was to a large degree dependent on the phase mixing method used. The results of the pressure drop experiments are in marked contrast to those of pure water and this coupled with the change in flow regimes and major shift of transition boundaries could be expected to give a poor fit to the correlation.

8.3) Detergent Solution with Low Mixing

The results of the pressure drop and flow regime experiments with the low mixing in the 11 mm pipe via T piece, of the air and Teepol solution were found to be equivalent to those of the pure water-air system. When plotted for the Lockhart-Martinelli correlation the results were again similar to those of the pure water case, and were a reasonable fit to the curve as shown in Figure 8.2 for the 0.05% and 0.5% solution.

Figure 8.2 : Correlation for detergent solutions with the T-piece mixer, in the 11.4 mm pipe.



The tabulated results and full detailed charts are given in the Appendix in Tables I.3 and I.4, and Figure H.3. and H.4.

8.4) Detergent Solutions with Medium Mixing

The results for the pressure drop and flow regime experiments for medium mixing, via sinter disc, of the detergent solution and air were markedly different to those of pure water-air. See sections 6 and 7 for details.

8.4.1) 11 mm Pipe

When data for the 11.4 mm pipe at two detergent concentrations is plotted for the Lockhart-Martinelli correlation the results are as shown in Figure 8.3. There is a marked shift of the curve for each liquid flow rate in the pipe. A selection of the liquid flow rates investigated is shown on the figure, at the two different detergent concentrations, .05% and .5%, the agreement of each set of data is clear, as is the general shape of the curves. At higher liquid rates the curve approaches that of the case for pure-water-air and the Lockhart and Martinelli correlation. A full plot of these curves, with mass fluxes indicated is given in Figure 8.4.

8.4.2) 17.3 mm Pipe

The results for the flow in the 17.3 mm pipe was found to be equivalent to the case of the 11.4 mm pipe. Results at a series of liquid flow rates for the 0.05% and 0.5% detergent concentration are shown in Figure 8.5. The agreement for each concentration can be clearly seen. At high

Figure 8.3: Correlation for 0.05% and 0.5% Teepol solution in the 11.4 mm. pipe, sinter generator.

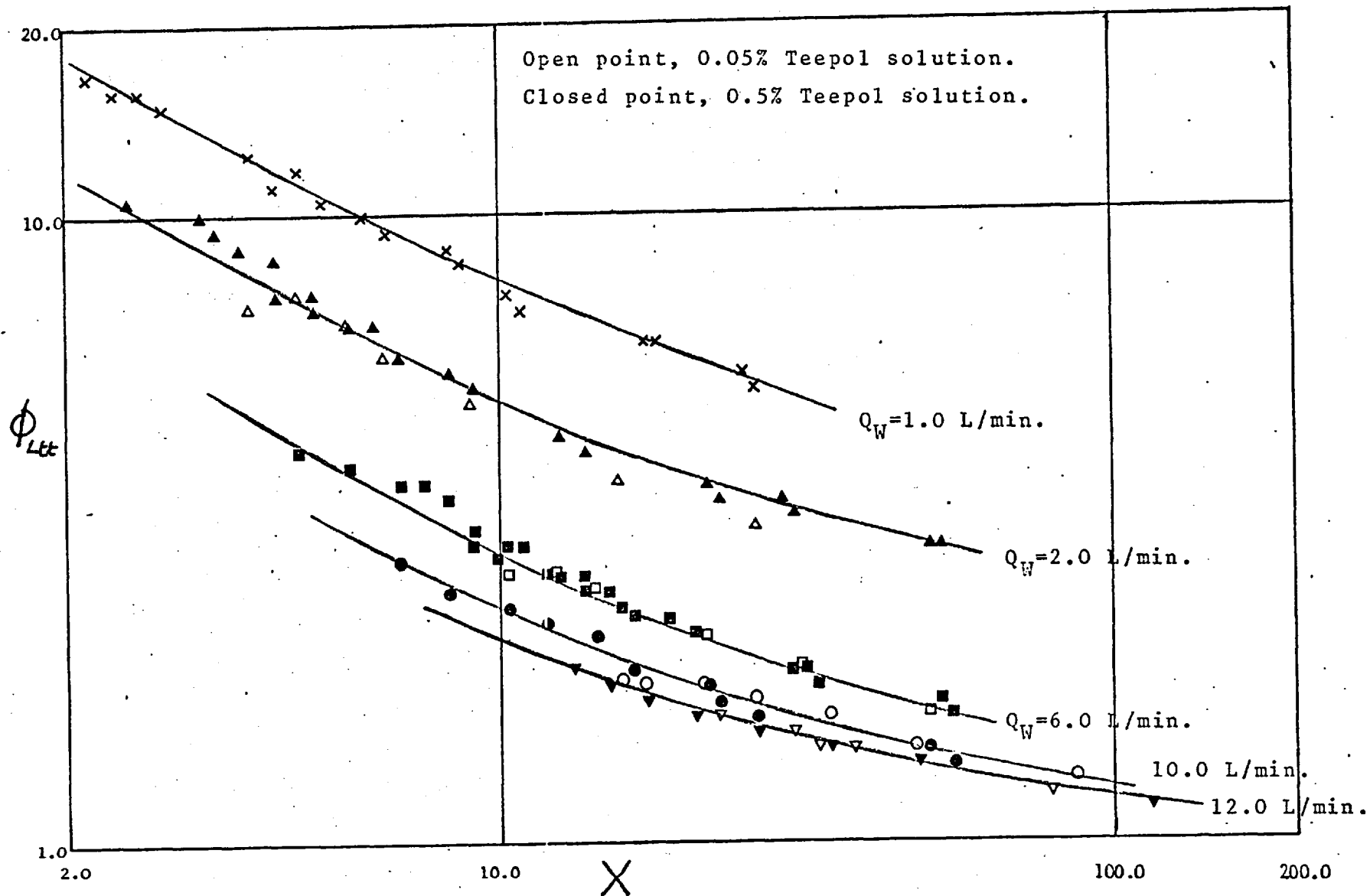
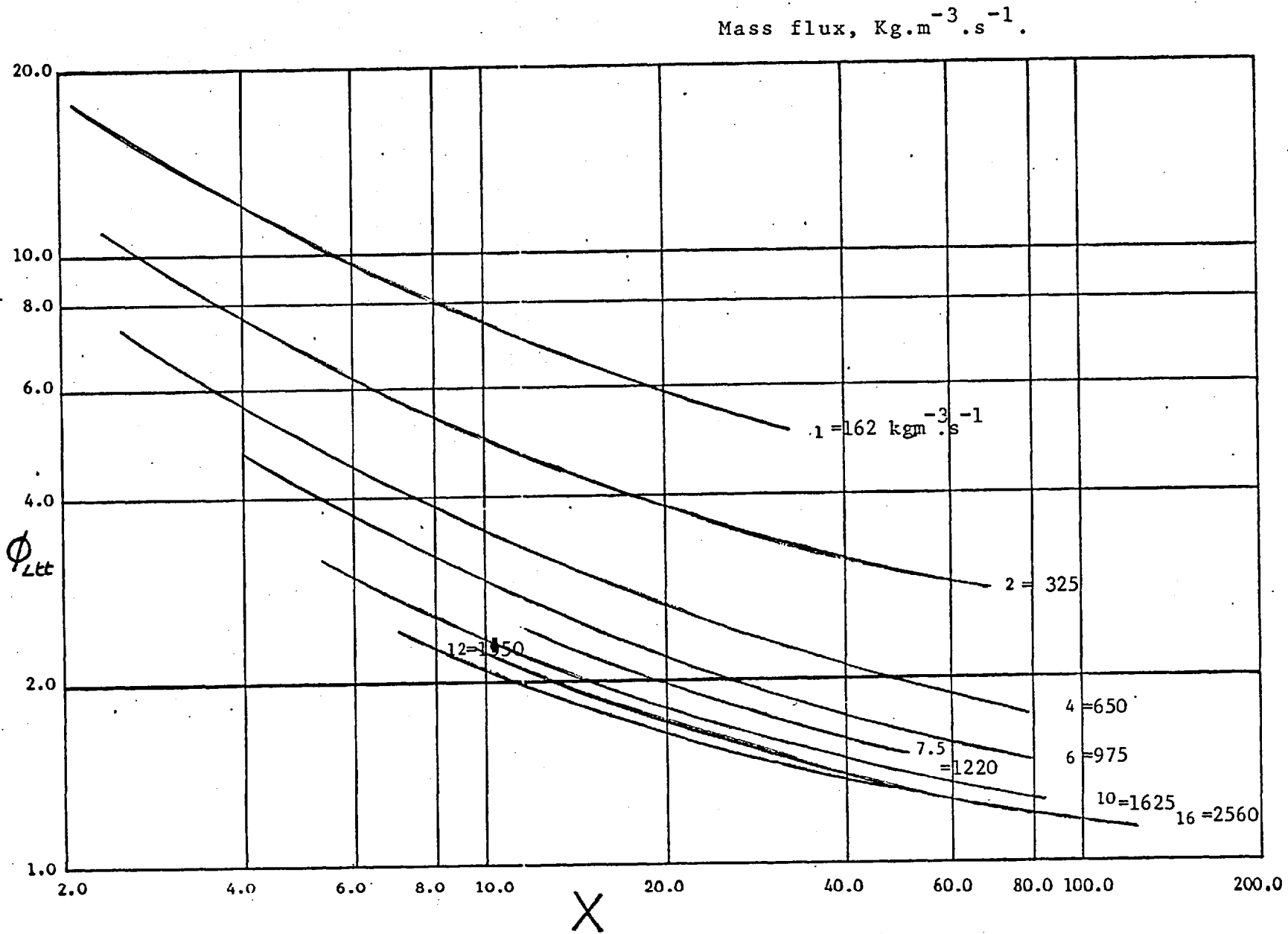


Figure 8.4: Curves obtained from the Correlation of Foam flow in the 11.4 mm pipe, with the Sinter generator.



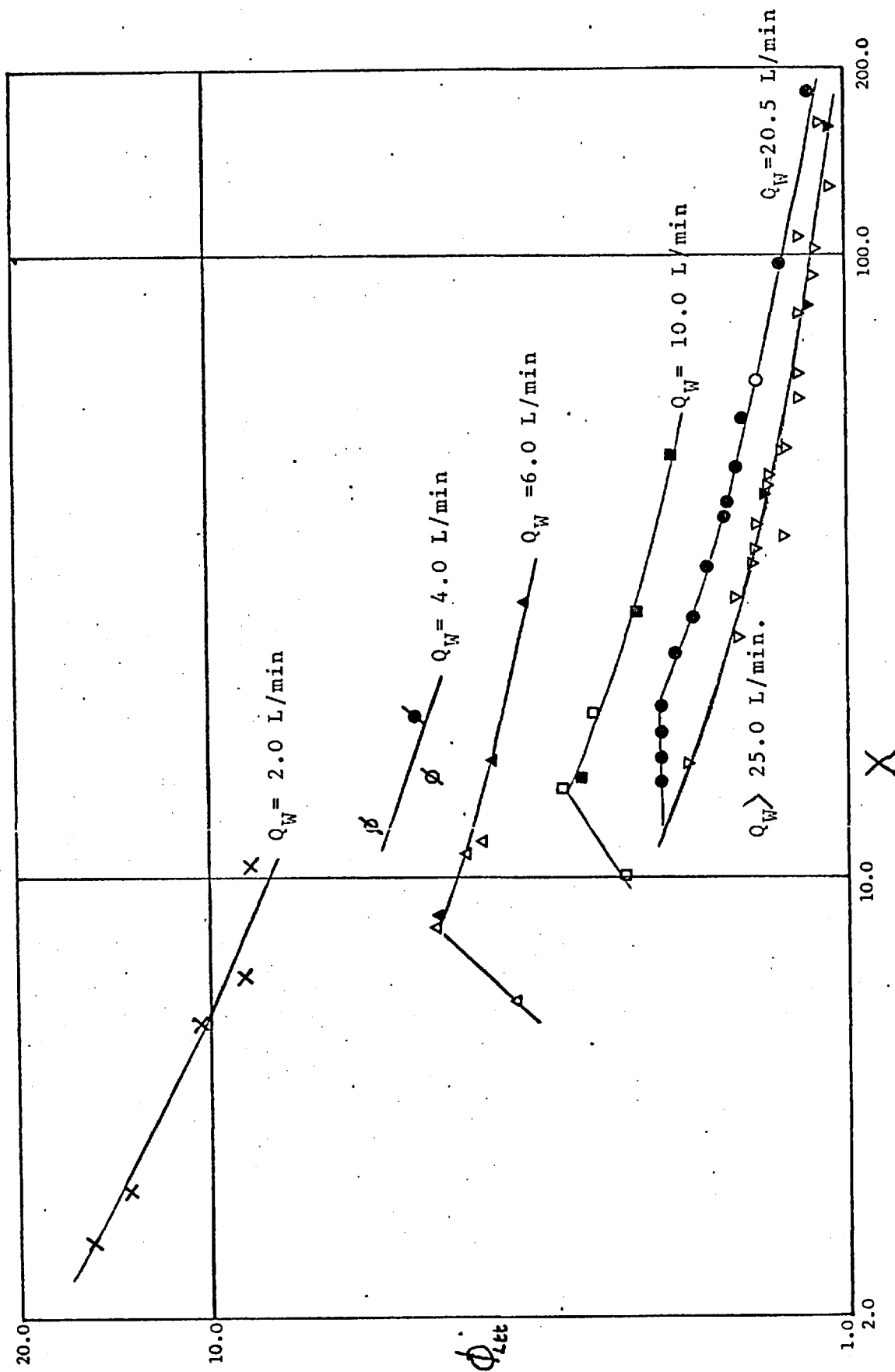


Figure 8.5: Correlation for 0.05% and 0.5% Teepol solutions in the 17.3 mm. pipe, sinter generator.

liquid flow rate, $Q_W > 25$ L/min, the curve coincides with that obtained for the pure water/air flow. At the low flow rates the curves are again displaced, the data for the 11.4 mm pipe and that of the 17.3 mm pipe falling on the same curves if the mass flux is used as a correlating factor.

The curves at low liquid rates tend however to approach the pure water/air curves at small values of X , when the flow pattern is in the slug regime. This can be clearly seen in Figure 8.5, for the data for flow rates of 6, 10 and 20 L/min.

8.4.3) 26.0 and 39.1 mm pipes

The results for the correlation for flow of 0.5% solution in the 26.0 and 39.1 mm pipes was again equivalent to the results obtained with the other pipes. These can be seen in Figure 8.7, where the lines have been plotted with the mass fluxes indicated. The curves for the 11.4 mm pipe are also given, the correspondence of the various mass flux curves being obvious. Results again tend to the pure water line at high water flow rates, the equipment restrictions being such that only comparatively low mass fluxes could be reached in the 26.0 and 39.1 mm pipes.

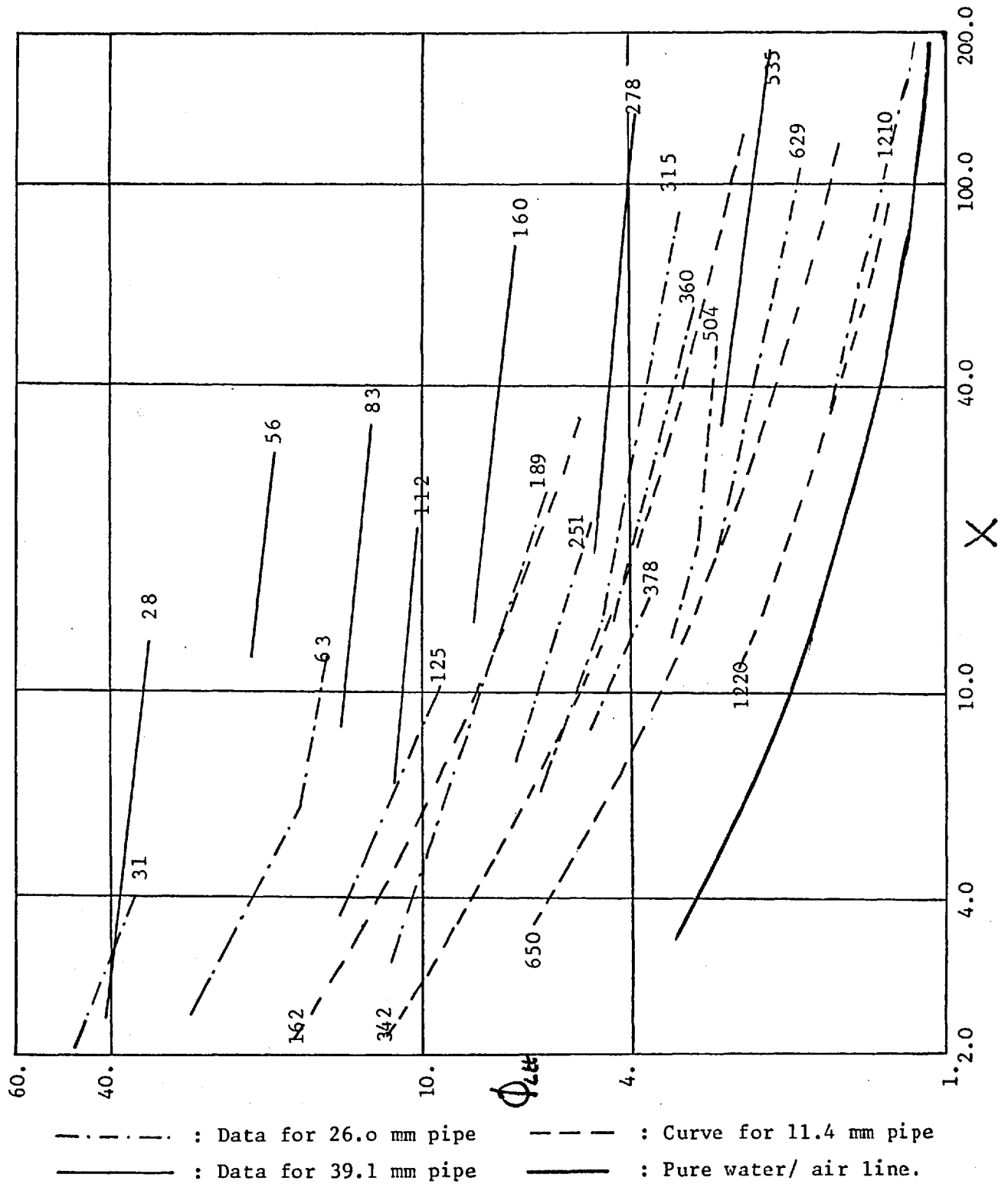


Figure 8.7: Results of Correlation for the 26.0 and 39.1 mm pipes, 0.5% Teepol, Sinter generator. Curves indicate the mass flux , $\text{Kg.m}^{-3}.\text{s}^{-1}$, for each flowrate.

8.5) Detergent Solution with Intense Mixing

The pressure drop results for the intense mixing method, using either a pump or valve generator are given in Section 7 . At low gas to liquid rates, such that X is large, the pressure drop went smoothly to the water alone pressure drop as the gas rate went to zero in a similar manner as the results for air-pure water and unlike the results obtained with the Sinter generator. The flow regime was at all times dispersed, there being no stratified or slugging regime.

8.5.1) 11.4 mm Pipe

The results for the correlation with the 0.5% solution in the 11.4 mm pipe is given in Figure 8.8 The discrete curves for each liquid rate are again present, but there is a tendency for the curves to converge at low air-water rates. At high liquid rates the curves tends to the overall Lockhart and Martinelli correlation curve.

8.5.2) 17.3 mm Pipe

The results for the case of flow in the 17.3 mm pipe with pump generator are shown in Figure 8.9 and are identical to those of the 11.4 mm pipe, with the curves being coincident if similar mass fluxes are considered for each pipe size.

There is however at the lower mass rates a distinct tendency of the curves to approach the overall curve for high flow rates, in a manner analogous to the case of the 17.3 mm pipe with Sinter generator. This occurs at low X values, where air-liquid ratios are high.

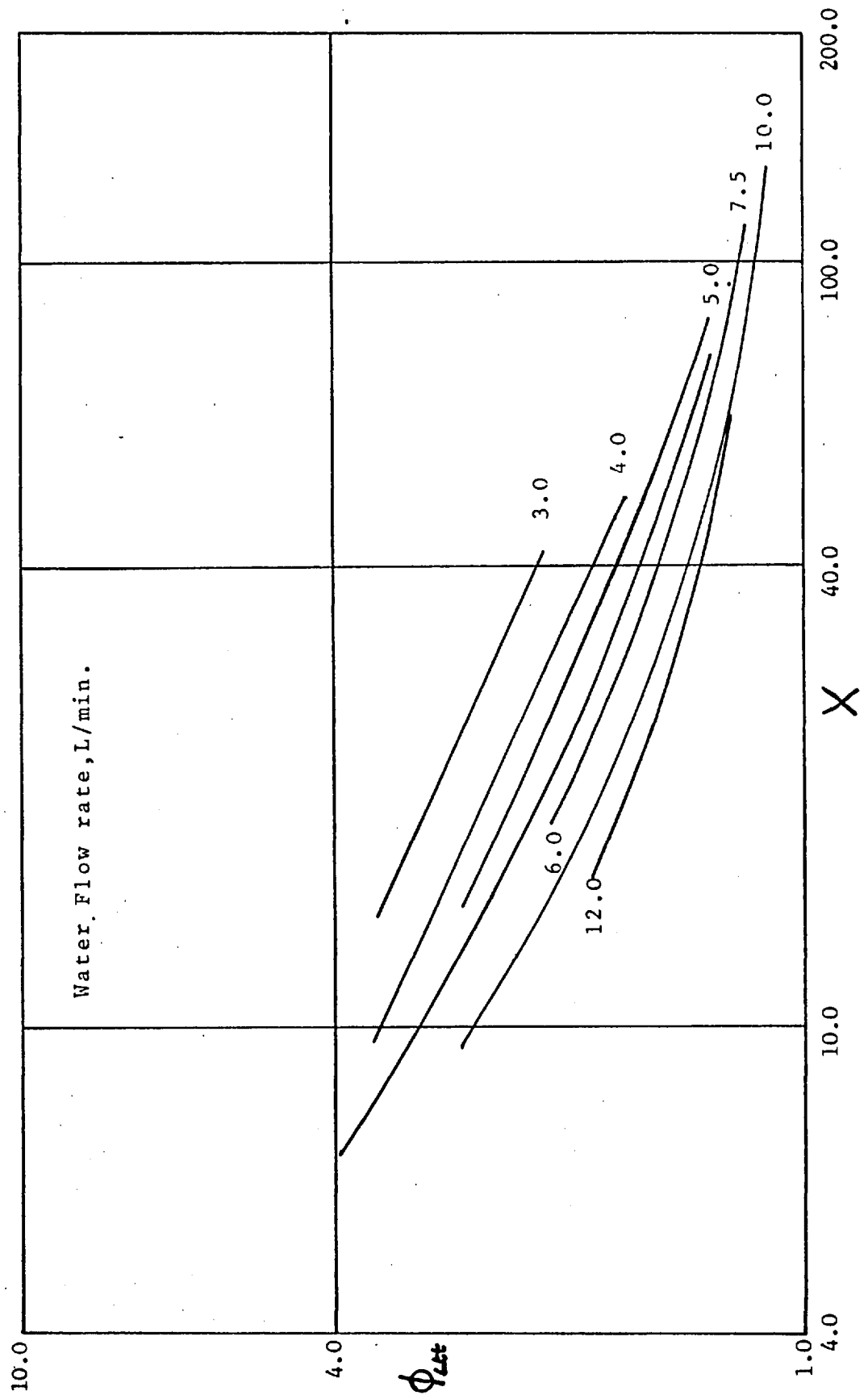


Figure 8.8: curves obtained from the Correlation of Foam flow in the 11.4 mm. pipe, Pump generator.

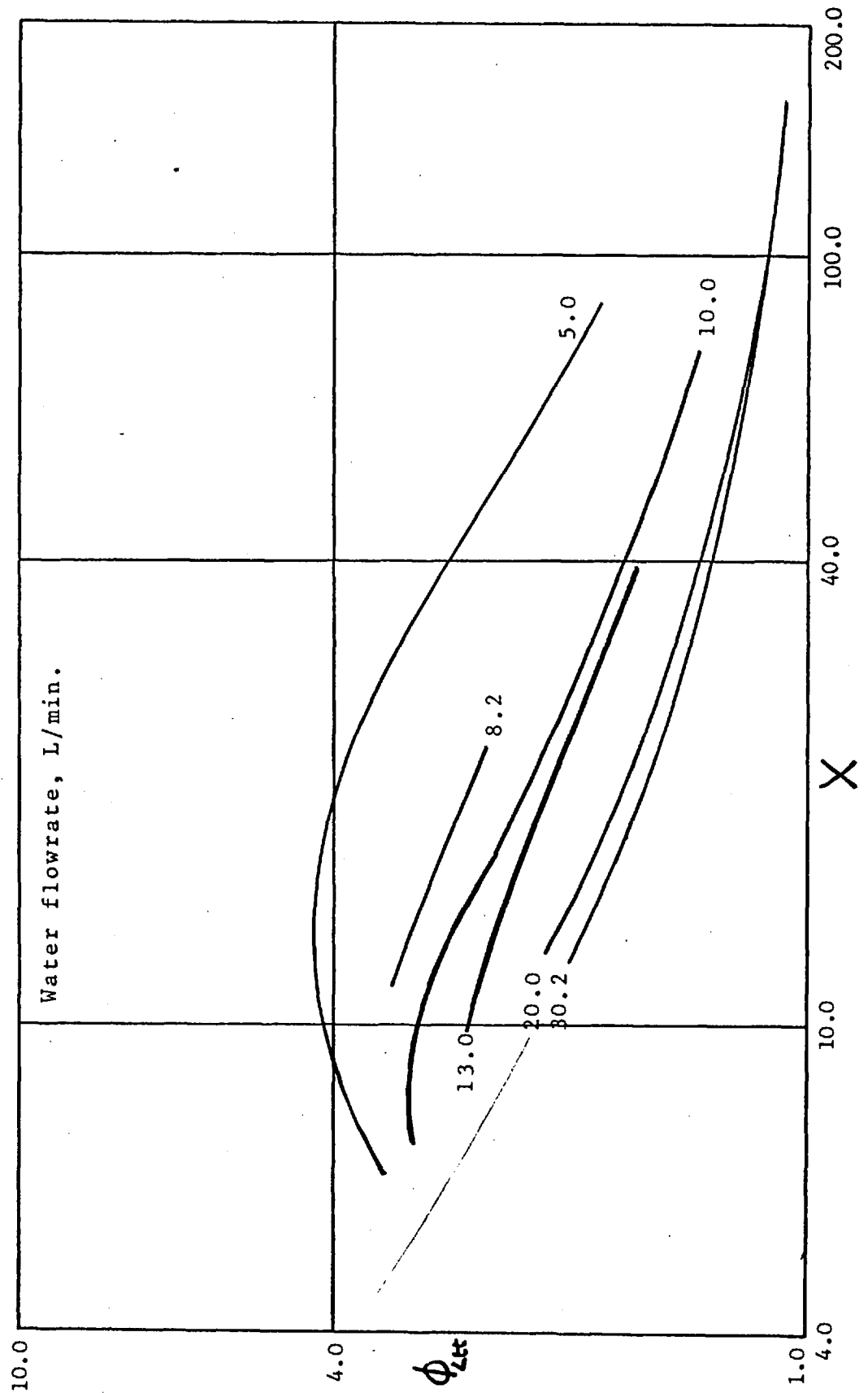


Figure 8.9: Curves obtained from the Correlation of Foam flow in the 17.3 mm. pipe, with Pump generator.

8.6) Discussion and Conclusions

8.6.1) Water-Air Flow in Pipes

The results of the correlation for the pure water-air case, in both pipe sizes and phase mixing methods can be seen in Figure 8.1. The data falls on to a distinct curve which lies just above the original Lockhart-Martinelli correlation. The fit to this correlation can be considered adequate however, as the data scatter is less than that used in the original correlation which included a wider range of fluids and pipe sizes.

The best line through the data was constructed, and is given in the table below, with the co-ordinates of the Lockhart- Martinelli correlation for comparison.

X	L-M curve $\phi_{Ltt}(45)$	This work ϕ_{Ltt}
4.0	2.38	2.95
7.0	1.96	2.29
10.0	1.75	1.98
20.0	1.48	1.60
40.0	1.29	1.36
70.0	1.17	1.23
100.0	1.11	1.15
200.0	--	1.06

The error in pressure drop for the best line through the data for the two pipes gives an average error of approximately 5% for $\frac{\Delta P}{L}_{calc.}$, while only a few points have an error of greater than 15%. These errors are within those acceptable for normal design purposes with two phase systems, and indicate the use of such a correlation for two phase flow situations.

8.6.2) Detergent Solutions with Low Mixing

The use of the T-piece as mixer in which the gas and detergent solution are introduced as separate stratified layers with no phase turbulence has been shown to give equivalent results for the pressure drop and flow pattern maps as the cases with pure water in the pipe. The results when plotted on the correlation are as expected, the points of both detergent solutions in the 11 mm pipe falling on the line obtained with the pure water-air flow system, see Figure 8.2.

Hence it could be expected that any solution capable of foaming which is flowing in a pipe in such a way as to stop the generation of foam, either by virtue of phase velocities, phase input to pipe, or insufficient length of pipe to cause significant foaming to occur over the length of interest, will give pressure drop results which can be correlated by the Lockhart-Martinelli technique.

8.6.3) Detergent Solution with Medium Mixing

The use of the Sinter as phase mixer, in which the gas is dispersed as many small bubbles in the liquid give many changes in the flow regimes, and pressure drop. The resultant foam is non-Newtonian in character, and has a yield stress. This might be expected to result in a poor fit to the correlation as the initial fluids used did not have this property.

The results are typified by Figure 8.4, which shows that each mass flux forms a separate and distinct curve. The data for all the pipes falls approximately on the same curves, and at high mass flux, $M_L > 2000 \text{ Kg.m}^{-3}.\text{s}^{-1}$.,

the curve is coincident with that of the pure water/air correlation. The curves are approximately parallel, there being only a slight tendency for the curves at high M_L to converge to the pure-water/air line at high X values. The gross shift of the curves to higher ϕ_{Ltt} values must be attributed to the non-Newtonian behaviour of the system, and particularly the yield stress. The following case will be used to illustrate the point.

A foam of equal volumes of gas and liquid is flowing in the pipe. From the worked data it can be seen that for the case when $Q_A \approx Q_L$, $X \approx 28.0$

$$\sqrt{\frac{\Delta P_{SL}}{\Delta P_{SL}}} \approx 800 + \sqrt{\frac{\Delta P_{2phase}}{\Delta P_{SL}}} = f(M_L)$$

If the total flow rate is reduced such that the foam is now flowing in it's laminar mode, although Re_G and Re_L may still be > 1000 , due to the Bingham plastic rheological properties of the foam, the pressure drop is still high. The extreme case is when $Q_{Foam} \approx 0$, at this point the pressure drop is due almost entirely to the yield strength, τ_o . But ΔP_{SL} will tend to zero by the model, as $Q_L \rightarrow 0$.

$$\therefore \sqrt{\frac{\Delta P_{Foam}}{\Delta P_{SL}}} \Rightarrow \infty$$

At low mass fluxes the curve will be displaced upwards. The position of the curves for each M_L could also be expected to be a function of the yield stress of the foam, in itself possibly a function of foaming agent, bubble size and pipe size.

At the higher M_L values the foam can be assumed to be fully turbulent and the effect of the yield stress small in relation to the total forces acting, and the results should approximate to the model, which made no assumptions on the rheological properties of the mixture. This is in fact the case, $M_L > 2000$ falling upon the pure water-air correlation line.

At low air-liquid ratios at low mass fluxes the bubbles form a stratified layer of foam at the top of the pipe. This layer maintains the yield stress properties at the very low mass flux, but the amount of data is not sufficient to make any conclusions on it's effect. The use of Figure 8.4 for the estimation of pressure drop for a large bubbled form, 200-800 micron size, is recommended, and can be expected to account for the case of stratified foam flow. However changes to the mass flux curves would be expected if the foam properties, yield stress, drainage times and bubbles size were different from the foam used in the experiments.

8.6.4) Detergent Solution with Intense Mixing

The use of a pump or valve to create intense mixing of the phases and very small bubbles gives an air-liquid mixture which was always in a dispersed state, over the flow rates investigated. The foam produced is non-Newtonian in character and has a yield stress.

The results of the experiments in the 11 mm pipe are given in Figure 8.8, the 17 mm results in Figure 8.9. Once again a family of curves for each mass flux results, but with certain differences to the case of Sinter generated

foams. The general argument used to account for the gross shift in curves is still applicable however.

The curves are no longer parallel but ⁿcoverage to the water-air line at high values of X , ie., low air-water ratios. As the air-water ratio is reduced less bubbles are formed, any bubbles formed are however small enough to be dispersed in the liquid and not form an upper foam layer as in the case of the Sinter generated foams. There comes a point when the bubbles in the liquid are far enough apart that they no longer interact, and the mixture loses its yield stress properties, which are due to the bubble interactions. The results then tend to that of pure water-air, as the effects of the detergent can no longer change the flow mechanism.

At high air-water rates, ie., low X , the curves for low Mass flux, $M_L < 900$, in the 17 mm pipe show a tendency to move back towards the water-air curve. This would indicate that the effect of any yield stress was starting to become unimportant as the gas bubbles started to fill the pipe. At sufficiently high air-water rates the bubbles are no longer spherical and the water present in the lamella between bubbles. In this case bubble wall rupture would occur at high velocities, breakdown of the bubble structure would mean that any yield stress effects of the static foam are reduced in importance and the system approximates to the pure water-air system as the effect of the detergent is no longer significant.

The difference in the curves for the mixing methods at various mass fluxes can be seen in Figure 8.10 where

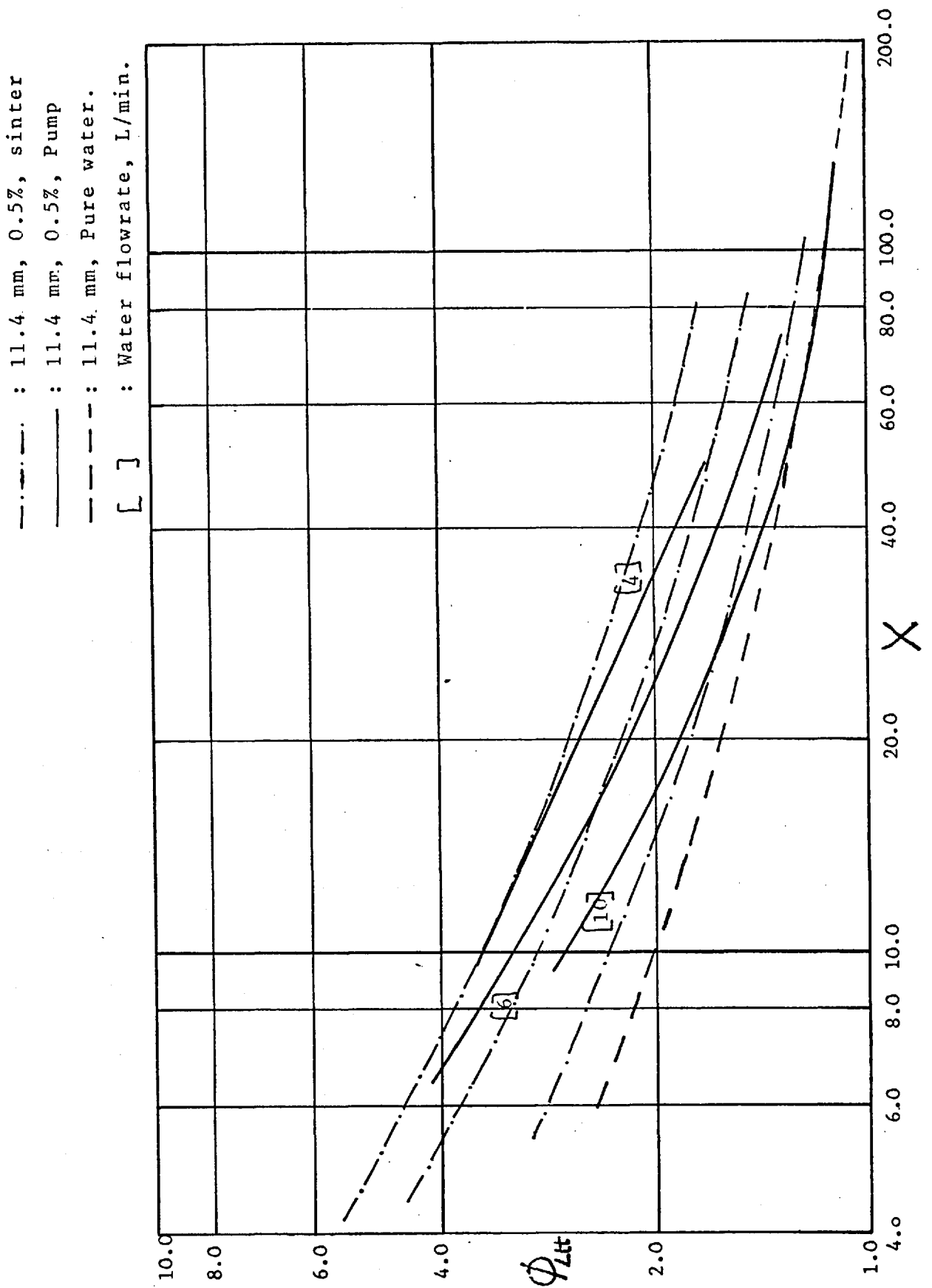


Figure 8.10: Comparison of curves obtained in the Correlation with the detergent solutions, for the Sinter and pump generator. The line for pure water included.

the data from the 11 mm pipe for Sinter and pump generators is shown along with pure water-air curve. The convergence at large X to the overall curve for the pump generator can be clearly seen. At $X > 40$, the curves for the pump are equivalent to the overall curve at $M_L > 1600$, while curves for the Sinter are only equivalent at $M_L > 2000$.

The use of Figure 8.8 is recommended for the estimation of ΔP for very fine bubbled foams which have a high degree of uniformity. At high gas to liquid ratios the curves for the various mass flow rates will tend to the air-water curve, insufficient data being available to indicate the likely behaviour.

9. Foam Flow Mechanisms

9.1) Stratified Foam Flow

The stratified foam flow pattern with the gas-detergent solution, using the Sinter disc air injection system, occurs over a wide range of gas and liquid superficial velocities, and approximately covers the range of the elongated bubble regime in gas-pure liquid flows.

The stratified foam regime corresponds in many ways more to the typical stratified flow regime of normal liquid-liquid systems rather than a gas-liquid system, with the difference between the foam phase and liquid phase accounting for the flow differences.

The foam acts as a psuedo-phase, with definable properties, a density less than that of the liquid phase but of comparable magnitude, and non-Newtonian rheological properties with an apparent viscosity about an order of magnitude greater than the liquid. The system which results from these differing properties can be compared most closely with that of the simultaneous flow of light oil and water phases flowing in stratified flow in a pipe. The situation of two phases flowing in laminar stratified flow, with the upper layer having a greater viscosity, has been studied both theoretically and experimentally and the behaviour of the system adequately explained in relation to the rheological properties of the liquids and the flow geometry, (75, 76).

The theoretical statement of the problem is that of two immiscible liquids flowing in a horizontal pipe. Exact solutions to the problem have been restricted to the case of laminar stratified flow of incompressible

Newtonian liquids in smooth, uniform pipes. With these restrictions the Navier Stokes equations can be employed to describe the behaviour of the fluids, for the situation under discussion, ie., steady uni-directional parallel flow in a horizontal pipe the equations reduce to

$$\frac{dP}{dx} = \mu \left(\frac{\partial^2 u}{\partial y^2} + \frac{\partial^2 u}{\partial x^2} \right) \quad 9.1)$$

for fluids 1,2.

The boundary conditions which apply are

- 1) No slip at the pipe or interface between the two liquids

$$U = 0 \text{ at wall} \quad \text{and} \quad U_1 = U_2 \text{ at interface}$$

- 2) The shear forces which each liquid exerts on the other must be equal at the interface

$$\mu_1 \frac{du_1}{dy} = \mu_2 \frac{du_2}{dy} \quad 9.2)$$

The exact analytical solution of this problem for pipe flow has been carried out by several workers, Teletov (77) and Dumitrescu et al (78), but solutions have only been obtained for the case of the interface on the pipe mid-point due to the complexity of the problem.

To overcome the analytical problems the approximate solution of the equations for two phase stratified flow of Newtonian fluids has been calculated by the use of appropriate numerical techniques for the solution of the equations. Charles and Redberger (75) and Gemmell and Epstein (76) have presented results for the solution of the equations, based on a finite difference approximation over a square mesh superimposed on the pipe, with the use of relaxation techniques and certain boundary approximation

assumptions to speed up solution convergence. They show that for a constant flow rate of the less dense, more viscous phase, ie., upper stratified layer, there is a reduction in pressure drop with addition of the less viscous phase. This reduction going through a maximum at a certain phase ratio. Results from both works are equivalent and diagrams from Gemmel and Epstein are shown which indicate the pressure reduction with less viscous fluid addition, along with the change in the maximum value of this reduction with viscosity ratio,

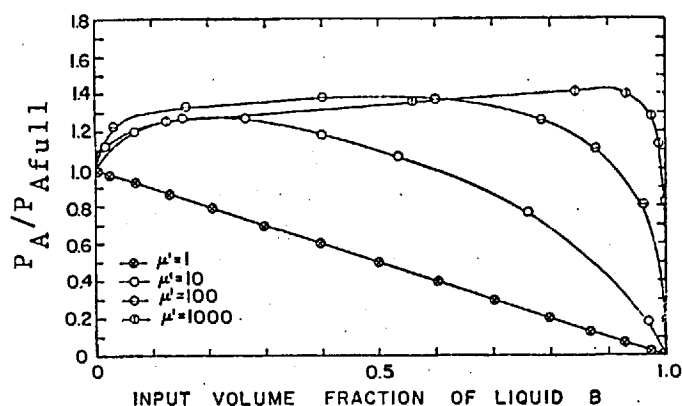


Figure 8—Variation of volumetric flow rate factor with input volume fraction.

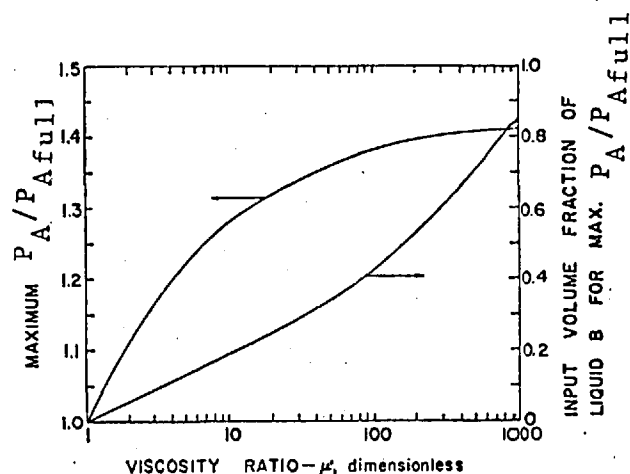


Figure 9—Maximum volumetric flow rate factor and corresponding input volume fraction as a function of viscosity ratio.

Figure 9.1: Typical results of the volumetric ratio and viscosity ratio of the water phase on the pressure gradient in the pipe, for constant flow of more viscous phase (76).

The difference of the velocity profiles which arise between the two layers is also shown in Figure 9.2 for a particular interface position.

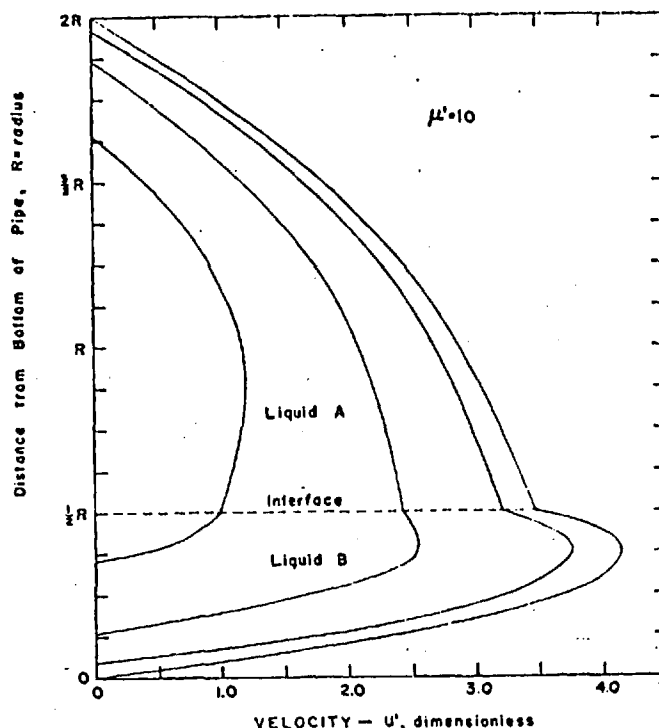


Figure 4—Final vertical velocity profiles for the interface at $y_i = R/2$ and a viscosity ratio of 10.

Figure 9.2: Typical velocity profiles for the upper and lower layers, showing the variation of velocity caused by the phase property differences(76).

Work by Yu et al (79) on the same system but employing a more sophisticated closed form solution, and using more powerful computer techniques has given equivalent results. However all these solutions have been restricted to the case of laminar flows, with a smooth interface. Conditions of more practical interest include the cases when one or the other phase is turbulent and the interface is no longer smooth. Recent work (80, 81) on gas-liquid flow systems in stratified flow has indicated that the solution of the problem, in the case of turbulent flows, can be adequately tackled by the use of a mechanistic model approach, and that results can be extended by these

analysis methods to cover liquid-liquid situations of practical interest.

No attempt has been made in this work to solve the case using the foam as the upper stratified layer, although sufficient investigation into the necessary techniques has indicated that the non-Newtonian nature and nonhomogeneity of the foam will increase the complexity of solution to a major extent. The rheological parameters necessary to describe the foam flow will to a certain degree depend upon the precise flow situation, in a real case the drainage of foam, and hence change in foam properties, will occur along the pipe, and the assumption of constancy of foam properties in both the axial and vertical directions may not be justified, unless short sections are being investigated. The wavy interface formed when the flow was turbulent has a very ordered structure, particularly in the larger pipe diameters, and must depend to a large extent on the foam properties. Waves on the interface were evenly spaced and appeared to move at approximately the same velocity as the lower foam surface. The water velocity was far greater than that of the foam, and hence the mechanistic model approach which is based on the flow in a conduit, consisting of part of the pipe, could be reasonably expected to apply to the liquid layer. The precise boundary conditions necessary to describe the surface will be complicated by the presence of waves, the roughness of the final bubble layer and the inter bubble attractive forces which tend to give any bubble mat an inherent 'strength'.

9.2) Foam Flow

The foam flow pattern which can exist at certain gas and liquid superficial velocities, is a homogeneous mixture of liquid and gas bubbles that fills the pipe. The foam, a psuedo-homogeneous phase, exists in it's precise state due to the mode of generation, the bubble size and distribution depending on the air injection method. Bubble sizes will only change due to foam decay and coalescence mechanisms, which are substantially long term effects when compared with the life-time of the foams investigated in this work.

When forces within the foam are sufficient to cause mechanical breakdown of the bubble the regime will have become slug flow.

The foam can exist in several states with varying degree of stability and homogenety.

9.2.1) Foam Produced by Intense Mixing Method

The foam produced by the intense mixing either in the valve or centrifugal pump, consisted essentially of a very fine dispersion of bubbles which are generated in the liquid and the properties of the mixture measured before the mechanical seperation of the two phases has proceeded to any great extent. The relative amounts of gas and liquid are to a certain extent dictated by the generating process, high liquid/gas ratios being possible with the intense mixers used, but not high gas/liquid ratios. The foam then consists of fine bubbles which are slowly seperating from the liquid, however rise velocities for the bubbles are slow, and hindered rising occurs due

to the bubble density. The generation process creates high turbulence which dies away leaving an initially random spacial distribution of bubbles through the bulk of the foam. Because high water to gas ratios are obtainable bubble separation can be large, and hence the foam properties will to a large extent depend on the bulk liquid properties. The foam next to the wall has a more ordered structure, a thin layer of water against the wall means that few, if any, bubbles will actually touch the wall, and the presence of this anisotropic layer will have a large effect upon the apparent rheological properties of the foam.

9.2.2) Foam Produced by Sinter Disc

The foam produced by the Sinter disc generator consists essentially of a mass of more evenly sized bubbles separated by liquid films. The bubbles have not undergone any size changes due to coalescence, and have the same size distribution as generated by the Sinter. The film thickness between the bubbles may be varying as the foam ages, but the thickness is less than that of the bubble size. Movement of bubbles is low and a static foam has an ordered structure, due to the close packing of the bubbles. Next to the wall there is a liquid layer, but at high gas to liquid ratios, when the bubble size is large and bubbles are deformable, the layer can consist of one wall of the bubble, of comparable thickness to the other water layer thickness between bubbles in the bulk.

If more water is present in the mixture, than can be accommodated in the lamellae between the bubbles, it will drain out and form a lower liquid layer, giving stratified

foam. In the general flow map for Sinter generated foams the foam-stratified layer transition boundary was found to be identical for all pipe sizes, the air-water ratio being a function of mixture velocity.

10. Conclusions.

Foam is a complex mixture of air and detergent solution, with non-Newtonian properties. Foam properties depend on the degree of dispersion of the gas phase in the liquid. Foams with sufficiently small bubbles sizes can be assumed to be a homogeneous fluid. The theory developed for two phase flow systems, with non-foaming liquids, have been found to be applicable for many of the foam properties.

10.1. Pump generated Foams.

This foam has very small bubbles, average diameter of 60 microns, and a long lifetime, t_{30} in the order of several minutes are possible. The foam can be considered homogeneous, and to have a definable set of fluid properties which do not depend on the flow system.

a) The foam behaves as a Bingham fluid, with the yield stress being a function of the gas content, and an apparent viscosity of about 6 centipoise in the voidage range 0.3 to 0.7.

b) No evidence of wall slip due to the two phase nature of the foam was noted. Results were rather inconclusive as insufficient data in the laminar region was available.

c) The foam exhibits a shock wave in high speed venturi flows, and should be an ideal fluid for the detailed examination of such phenomena in two phase mixtures.

10.2. Sinter generated Foams.

This foam had a much larger bubble size, average diameter being about 300 microns, with a very short lifetime, only in the order of seconds. The gas tended to separate out of the mixture, into a dry foam and liquid phases.

a) Fluid properties on the Bingham model were again found, but results were due to the 'lumped' properties of the foam and liquid phases flowing together in the pipe. Any full analysis will require a consideration of the properties of the two layers separately, to determine what the effects of pipe size and phase ratio would be on the total flow.

b) This foam exhibited a complex flow regime pattern map, depending on the detergent concentration, degree of dispersion of the gas phase, and the pipe size. The main changes, to that of pure liquid/air flow, were due to the foam effectively replacing the gas phase. The major effects were to extend the stratified region over the elongated bubble regime, and to suppress the formation of air slugs, greater gas phase velocity being needed to break down the foam bubbles.

Both foams could be correlated by the Lockhart-Martinelli model at high mass flowrates. At low mass flowrates the Bingham fluid properties predominate, and create a family of curves. Each foam investigated gave unique curves for each mass flux, results for all the pipes being equivalent.

Bubble size measurement by photographic methods was used, and the use of the excess pressure in the bubble to obtain an estimate of the bubble size for the pump generated foam indicated promise.

10.3. Suggestions for further work.

1) Additional experimental work with a range of smaller sized pipes with the pump generated foam, to check on the fluid properties, and determine if wall slip effects are present in the foam

2) Experimental work with equipment capable of greater

foam input pressures for nozzle flow investigations. The few results obtained in nozzle flow were hindered by the rather low pressure capability of the designed equipment. The use of foam in the investigation of two phase critical flows is ideal due to the homogeneous nature of the foam.

3) Theoretical and further experimental work on the stratified foam flow regime. This particular flow pattern is the most likely mode of flow in very long pipes, and a better understanding of the flow mechanics will be needed for design purposes.

4) Certain aspects of the work, especially in the flow regime experiments, should be repeated with a different surface active agents, to determine if the results are generally applicable to any system capable of foaming.

The field of two phase flow investigation is made doubly difficult by the wide range of fluid properties which can be chosen for the phases, as well as the extremely difficult problems involved in the description of the general problem. This work on foam flow in pipes, and the investigation into the factors which might be used to characterise the foams, has touched on a topic which up to now has been relatively unexplored. So many aspects of the problem of foam flows, notwithstanding the general case, are as yet so little understood that almost any avenue of investigation will yield new and novel information.

11.

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Appendix A;

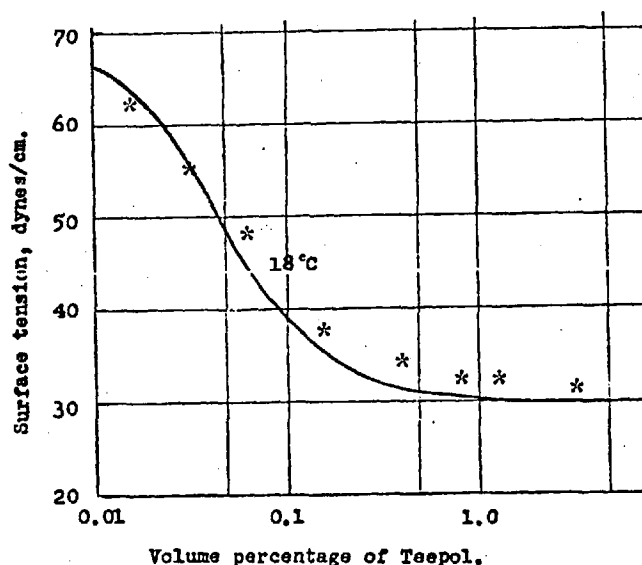
Teepol L; Properties and Composition

The detergent used in the experimental work was Teepol L, a general purpose cleaning agent manufactured by Shell Chemicals Ltd. Detergent used in the experiments was obtained from the standard four gallon drums supplied for industrial use, many different drums over a period of time were used to provide the detergent.

Teepol L is comprised of a solution of secondary alky sulphate, of average molecular weight 305, in water. The detergent used had a 27% active component (84). The volume percentage of Teepol used in the experiments was for the liquid as provided in the drums.

Density of the Teepol L used was 1.035 gm/cm^3 . The viscosity and density of the detergent solutions were taken as equal to the values for water.

The effect on surface tension of the concentration is given in Figure A.1, data collected by use of a Ring Tensometer, and from the literature (73).



* : Data taken at 20°C .
on Teepol L used
in experiments.

Figure A.1 : Variation of surface tension with Teepol L concentration, vol/vol.

Appendix B;

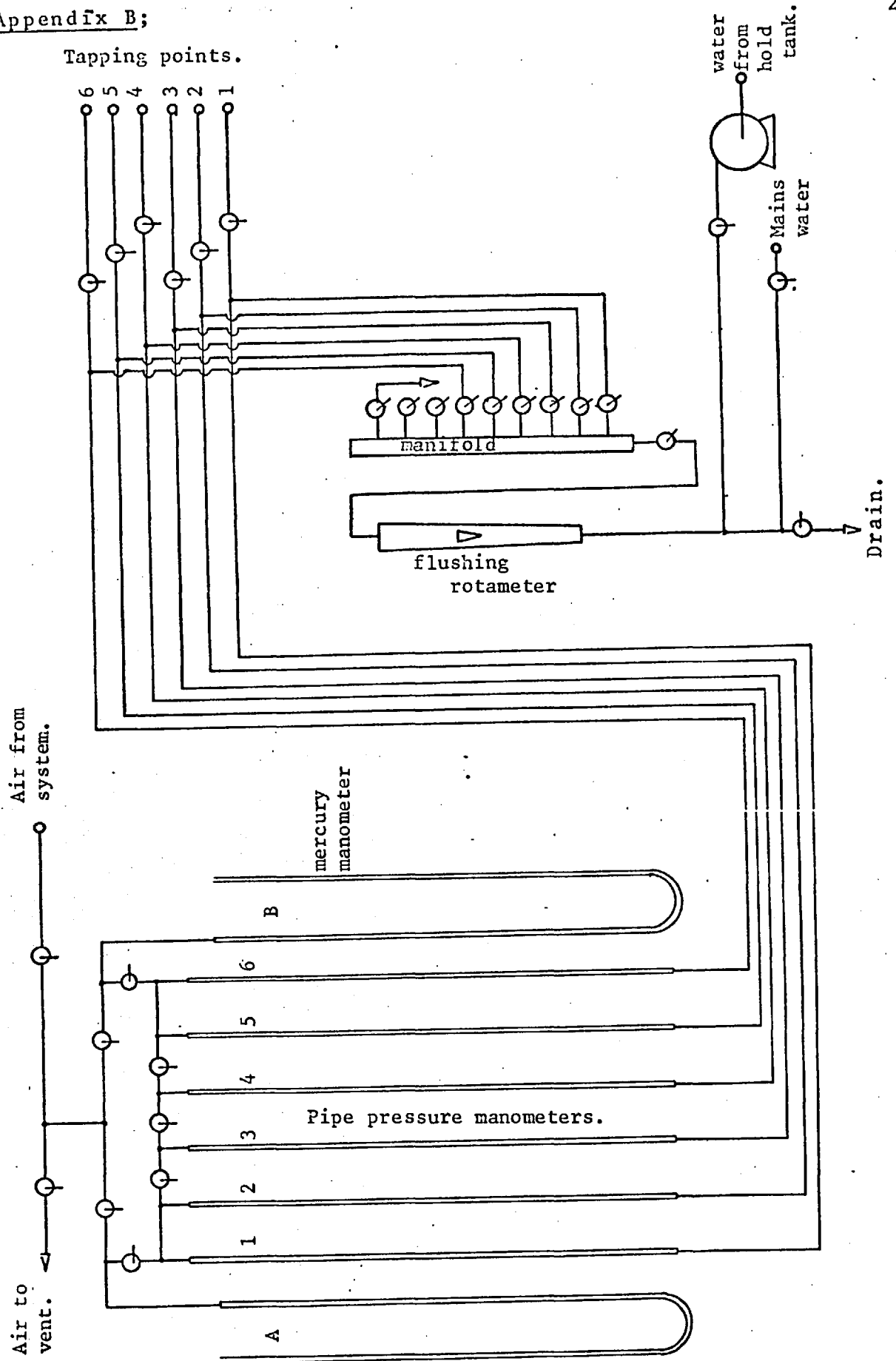


Figure B.1: Full manometer pipework for six tapping points.
Manometers , glass tubing. Lines , $\frac{1}{8}$ " copper.

Appendix C:

Table C.1: Foam Drainage results for pump generated foams.

1) Foams produced in the small scale generator.

a) sample measured in vessel

b) sample measured in graduated cylinder.

Total height = 20.0 cm.

Total volume = 250 cm³.

water height, h = 12.3 cm.

water volume, V = 158.0 cm³.

air/ water ratio = 0.625

air/water ratio = 0.583

Foam half-life, t = 220 s.

Foam half-life, t = 195 s.

h	t	h/h	t/t	V	t	V/V	t/t
cm.	sec.			cm ³	sec.		
0.9*	32	0.07	0.15	25*	60	.16	.31
1.5*	50	0.12	0.23	46*	120	.29	.61
1.9*	70	0.15	0.32	70*	180	.44	.92
2.4*	90	0.19	0.41	97*	240	.62	1.23
2.9*	110	0.23	0.50	115	300	.73	1.54
3.5*	125	0.28	0.57	123	360	.78	1.89
4.1*	150	0.32	0.68	132	480	.83	2.46
4.8*	175	0.38	0.80	140	600	.88	3.08
5.6*	205	0.44	0.93	142	660	.90	3.39
6.9*	240	0.54	1.09	144	720	.91	3.69
7.7*	270	0.61	1.23	146	780	.93	4.00
8.6	300	0.68	1.37	148	840	.94	4.31
9.1	340	0.72	1.55	151	900	.95	4.62
9.8	395	0.77	1.80	152	960	.96	4.92

* : Interface between liquid and foam layers indistinct.

2) Foams produced in the test pipe, samples measured in 1000 cm³ graduated cylinder.Total volume = 1000 cm³Total volume = 1000 cm³Water volume, V = 475 cm³Water volume, V = 374 cm³

air/water ratio = 1.10

air/water ratio = 1.70

Foam half-life, t = 172 s.

Foam half-life, t = 140 s.

V	t	V/V	t/t	V	t	V/V	t/t
cm ³	sec.						
100*	90	0.21	0.52	70*	77	0.19	0.55
210*	154	0.44	0.89	100*	100	0.27	0.71
300*	215	0.63	1.25	210*	155	0.56	1.11
350	253	0.73	1.47	240*	173	0.64	1.24
380	318	0.80	1.85	250	195	0.67	1.39
				270	240	0.72	1.71
				280	265	0.75	1.89
				300	325	0.80	2.32
				325	440	0.87	3.14

Table C.2: Foam Drainage results for Stratified Foam flow .

Foam flowing in the 39.1 mm. pipe in the stratified flow regime, generated with the sinter disc, was collected and the volume drained from the sample with time measured.

A constant water flow of 10.4 L/min. was used.

Foam sample = 1000 cm³.

Q_A L/min	V cm ³	t sec.	V/V_∞	R_L	V_∞ cm ³	$V - R_L \cdot V_\infty$	$R_F \cdot V_\infty$	$\frac{V - R_L \cdot V_\infty}{R_F \cdot V_\infty}$	R_F	$\frac{Q_A}{Q_L}$
8.1	480	10	0.84	0.72	572	52	162	0.32	0.28	0.78
	510	15	0.89			82		0.51		
	530	21	0.93			102		0.63		
	540	27	0.94			112		0.69		
	550	37	0.96			122		0.75		
	560	60	0.98			132		0.82		
16.3	300	10	0.77	0.65	390	46	136	0.34	0.35	1.57
	330	17	0.85			76		0.56		
	350	26	0.90			96		0.71		
	360	32	0.92			106		0.78		
	370	44	0.95			116		0.85		
23.4	150	5	0.49	0.34	307	45	202	0.22	0.66	2.25
	190	10	0.62			85		0.42		
	220	16	0.72			105		0.52		
	240	22	0.78			135		0.66		
	250	27	0.81			145		0.72		
	260	32	0.85			155		0.77		
	270	42	0.88			165		0.82		
	280	56	0.91			175		0.86		
30.3	100	5	0.36	0.22	282	38	220	0.17	0.78	2.91
	140	10	0.50			78		0.35		
	170	15	0.60			108		0.49		
	190	20	0.67			128		0.58		
	210	27	0.75			148		0.67		
	220	30	0.78			158		0.72		
	230	37	0.82			168		0.76		
	240	43	0.85			178		0.81		
	250	54	0.89			188		0.85		

Appendix D:

Typical bubble Histogram for Sinter generated foams

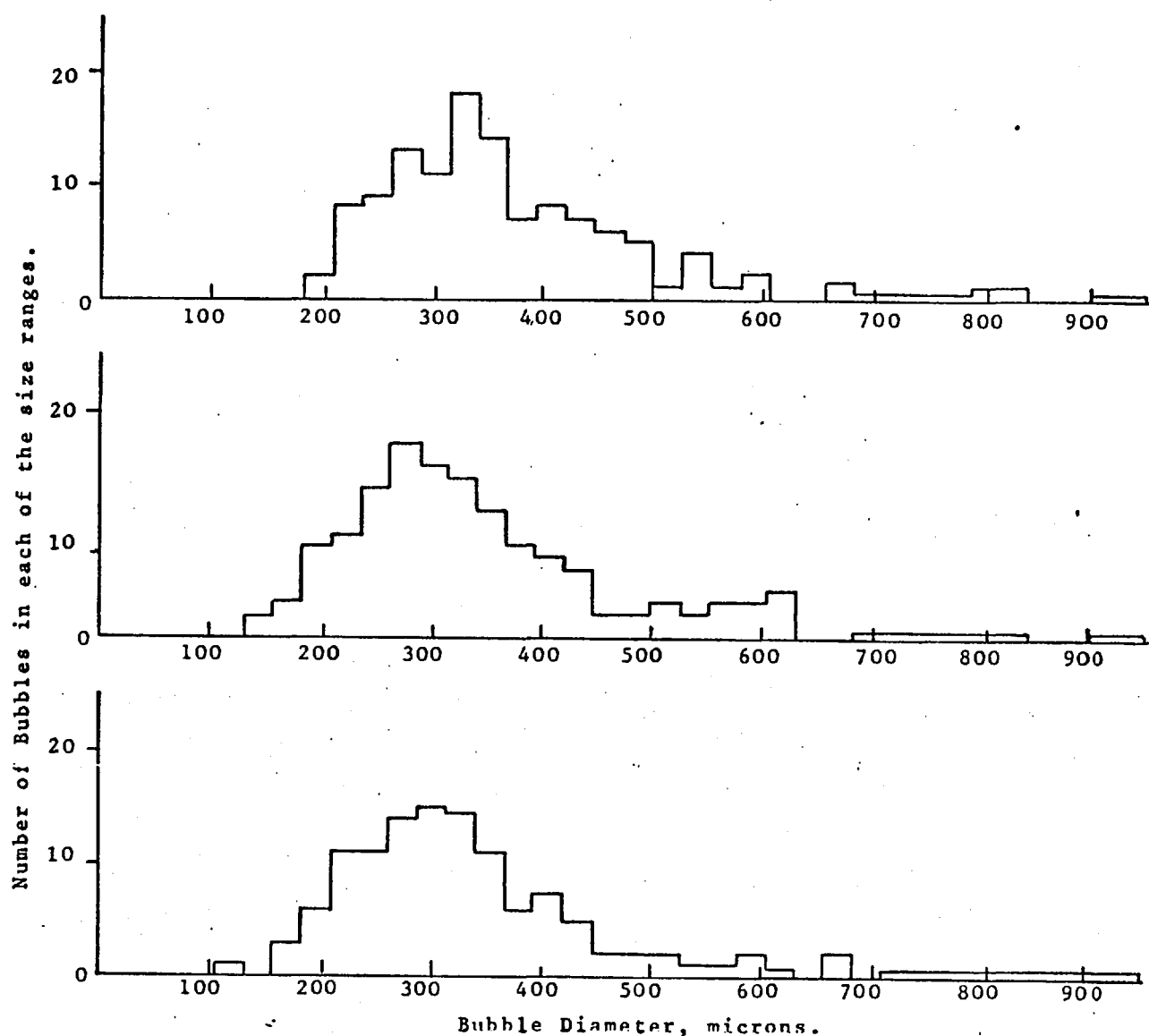


Figure D.1: Histograms taken from the photographs of Sinter generated foams, Prints used to obtain these histograms are given in Figure E.1.

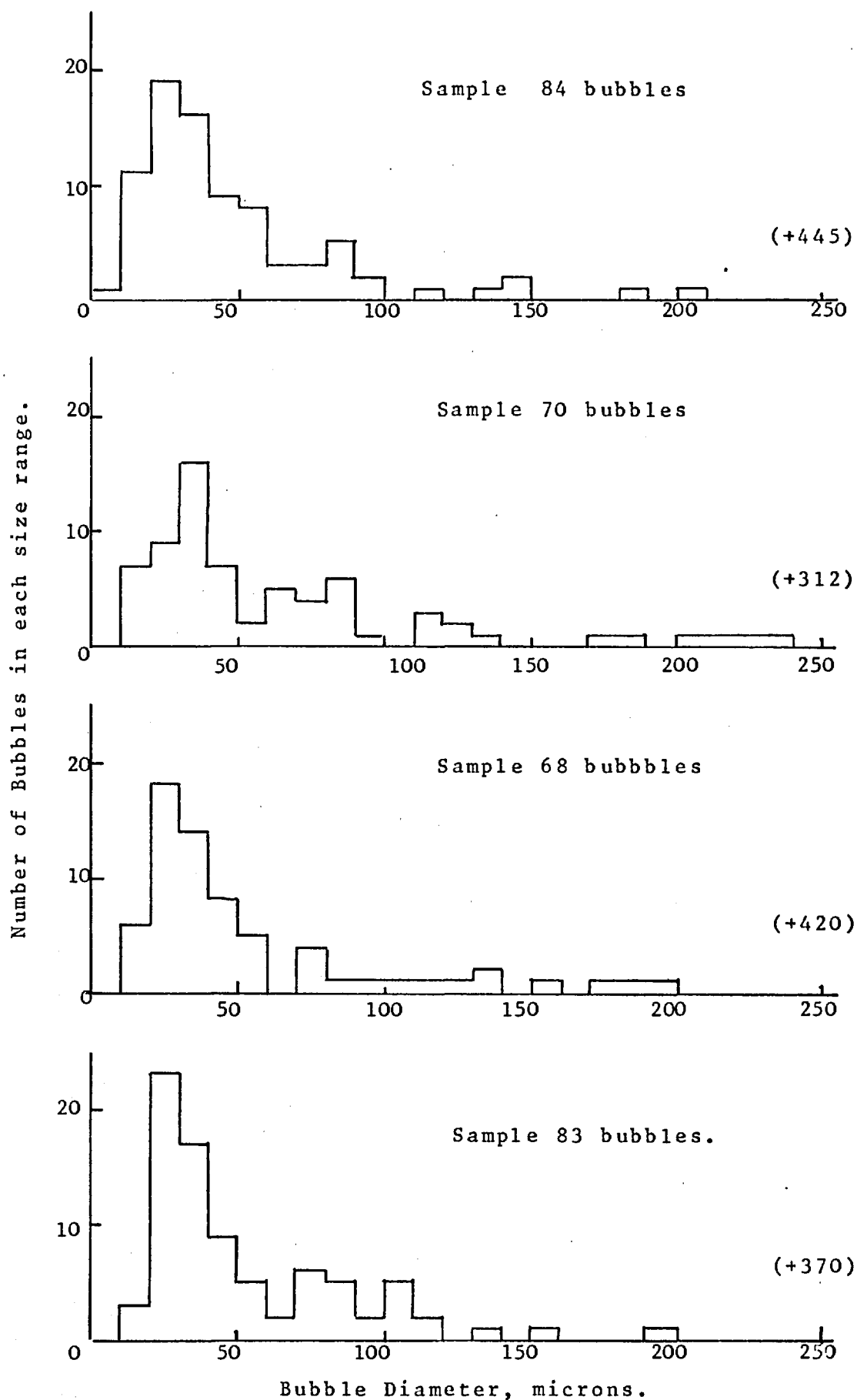


Figure D.2.: Histograms taken from the Photographs of Pump generated foams, prints used to obtain these histograms are given in Figure E.2.

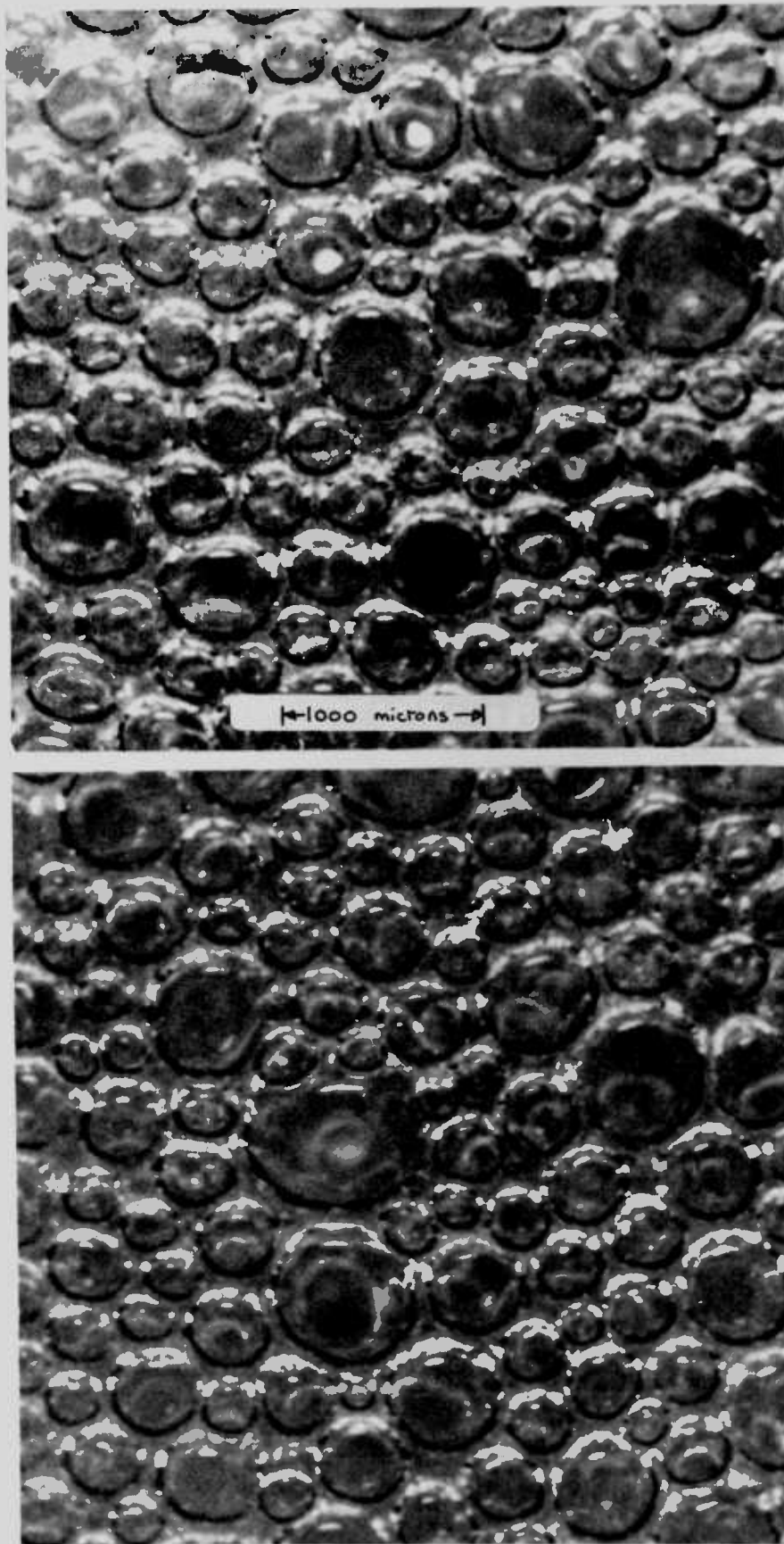
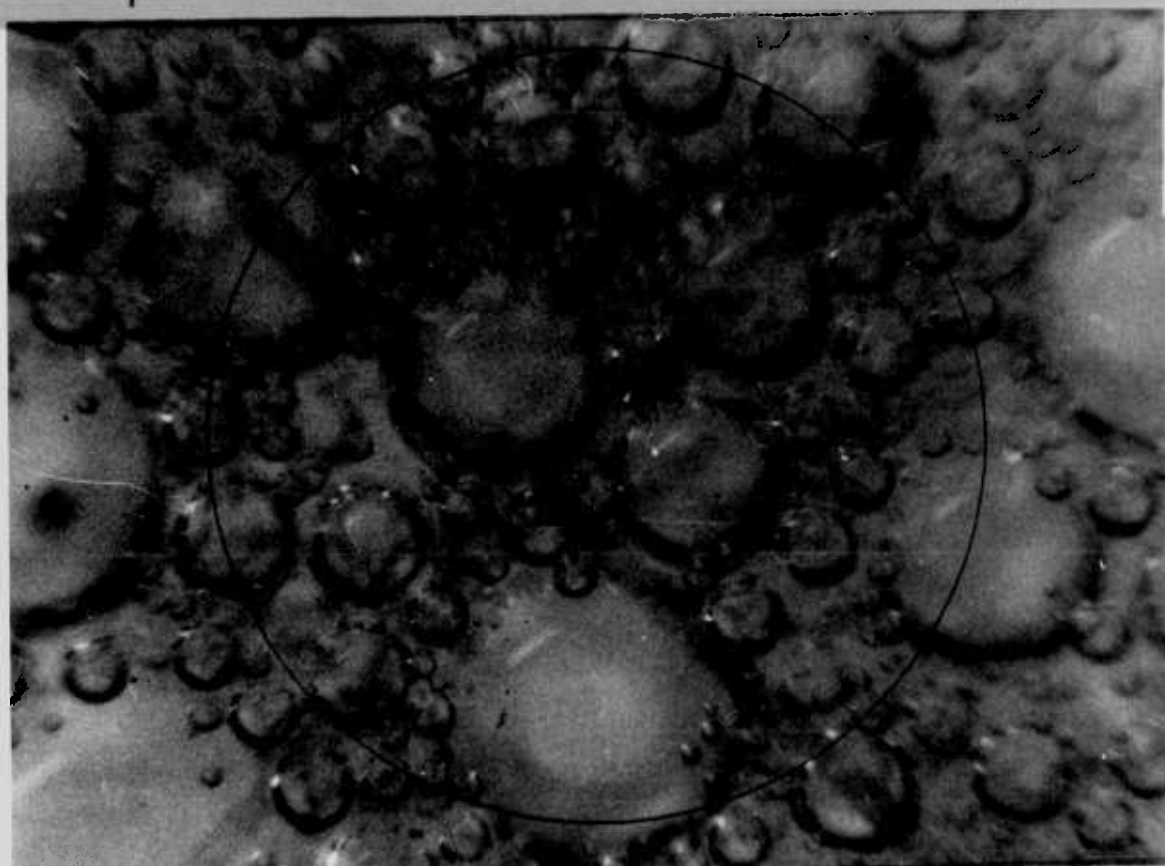
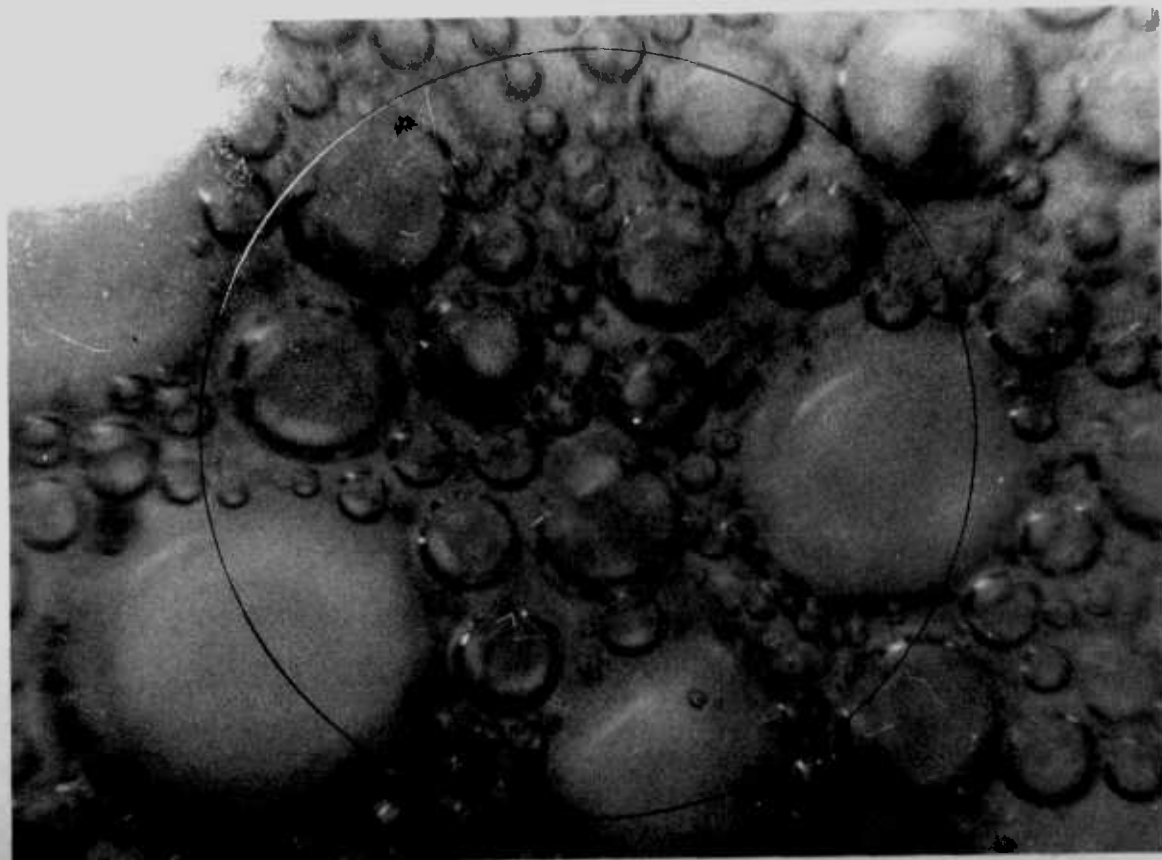
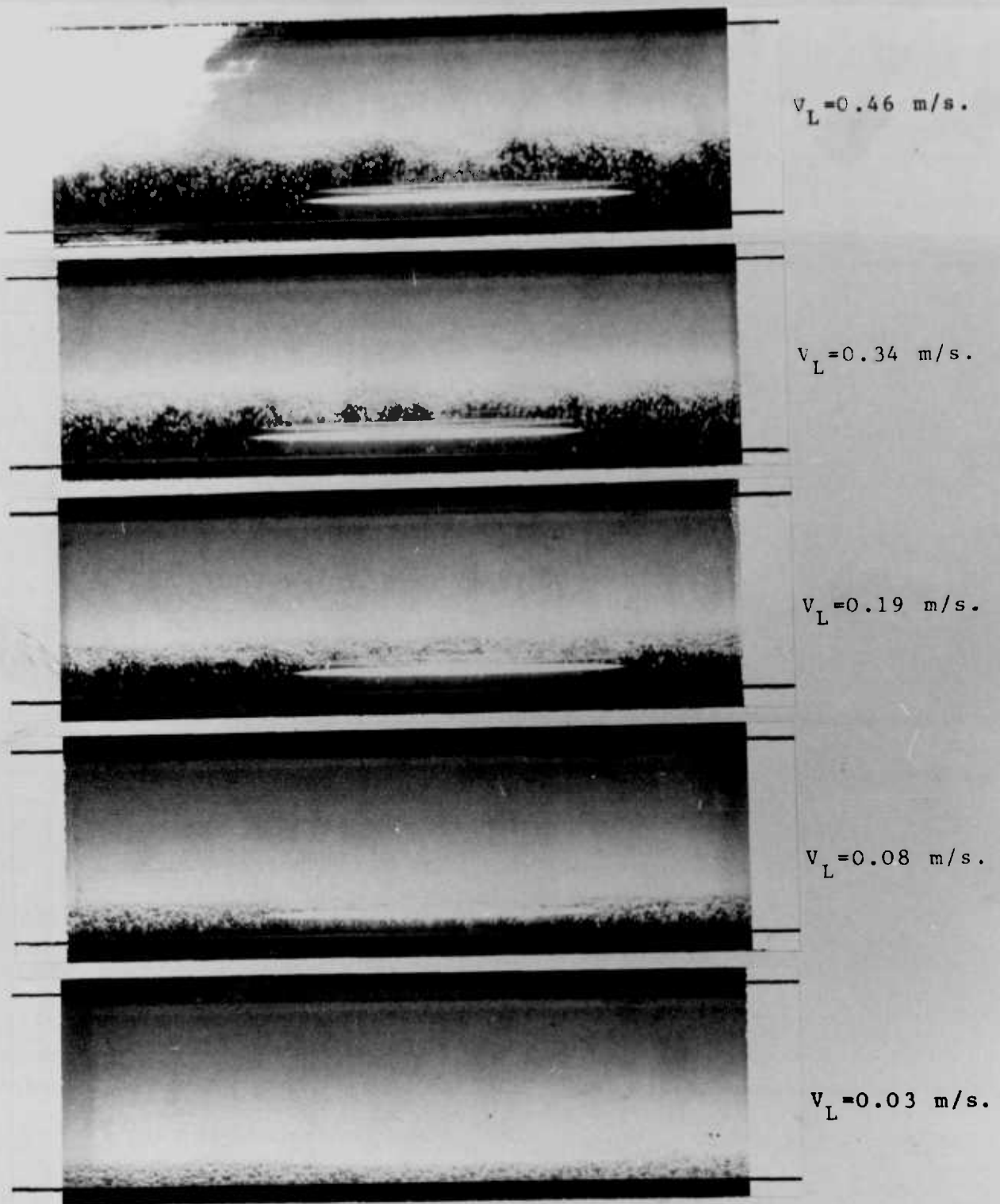


Figure E.1: Photographs of typical Sinter generated foam in the 17.3 mm pipe. $Q_A = 15.4$ L/min, $Q_W = 7.4$ L/min. 0.5% Teepol.



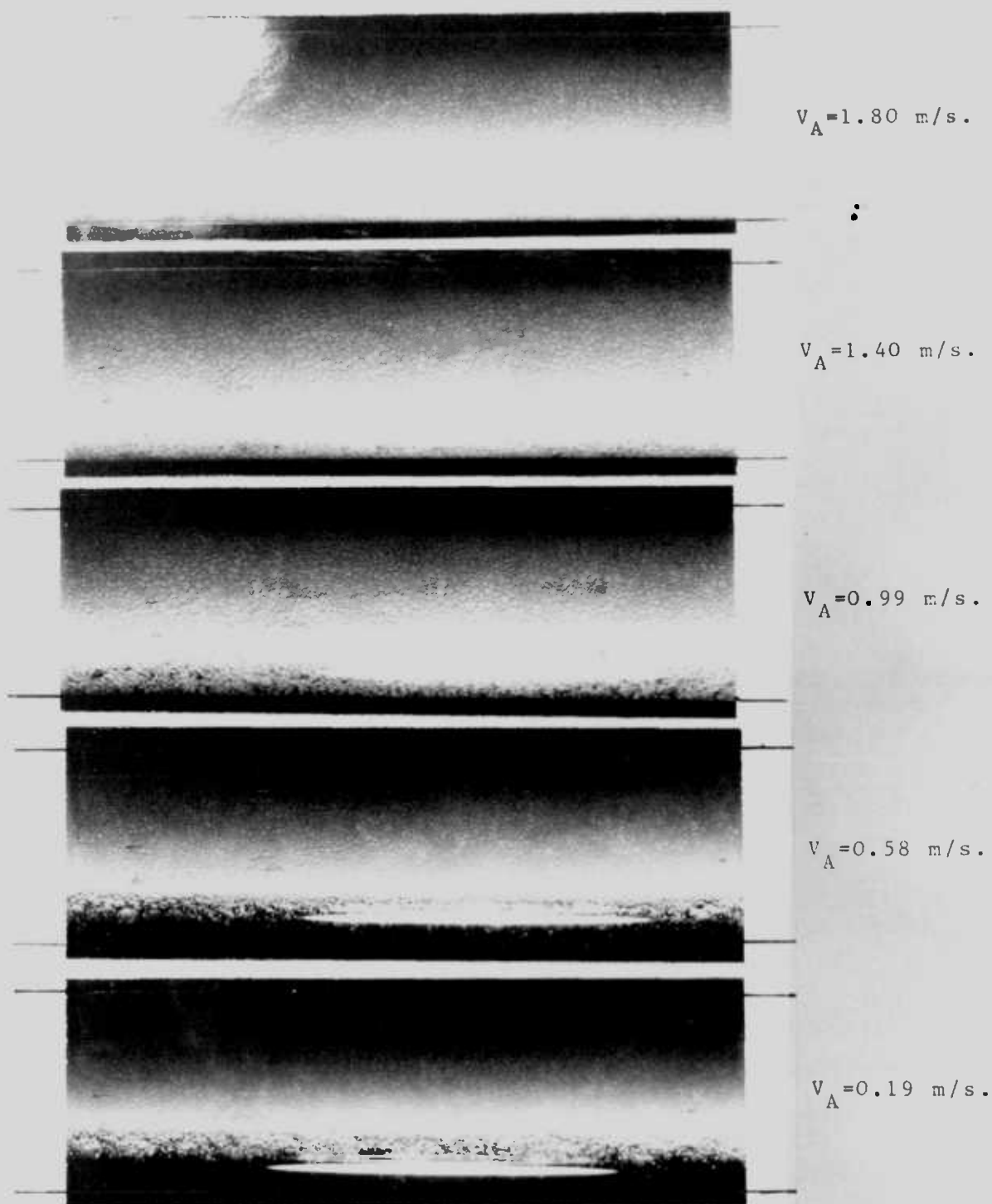
Bubbles inside circles measured.

Figure E.2: Photographs of typical Pump generated foam in the 17.3 mm pipe. $Q_A = 10.2$ L/min, $Q_W = 10.2$ L/min. 0.5% Teepol.



Flow Regime changing from Smooth stratified foam flow to Wavy stratified foam flow as the superficial liquid velocity increases. Photograph of complete 26.0 mm pipe, scale = x 1.3.

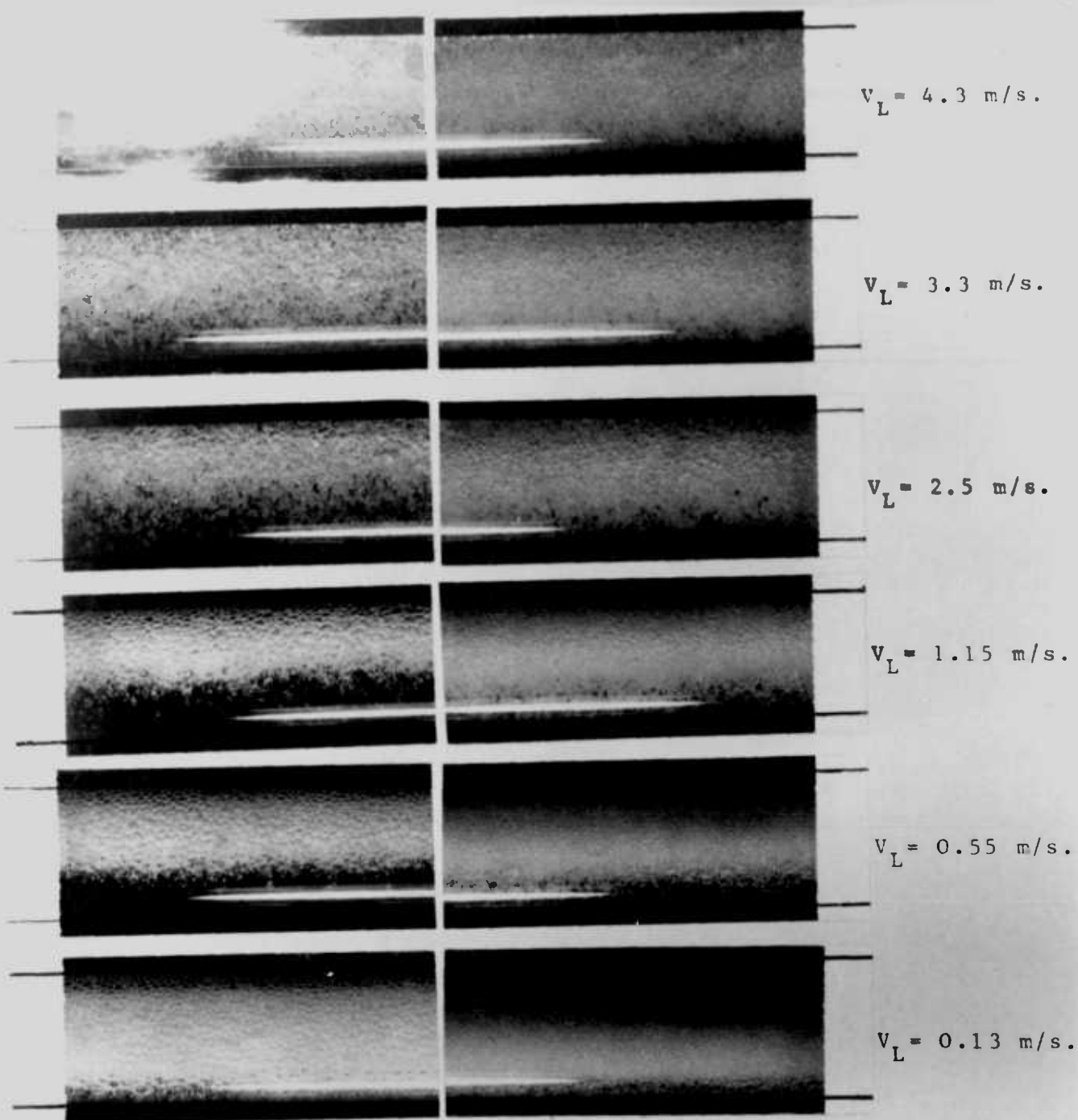
Figure E.3: Photograph of 26.0 mm pipe, with a constant air flowrate, $V_A = 0.19$ m/s, Sinter generator, 0.5% Teepol.



Flow regime changing from Smooth stratified foam flow to foam flow as the superficial gas velocity increases.

Photograph of complete 26.0 mm pipe, scale = x 1.3.

Figure E.4: Photograph of 26.0 mm pipe, with a constant water flowrate, $V_L = 0.19 \text{ m/s.}$ Sinter generator, 0.5% Teepol.



0.05 % Teepol solution

0.5% Teepol solution

Photographs of the 17.3 mm pipe, showing the Stratified foam to Dispersed flow regimes, for 0.05% and 0.5% Teepol solutions. Change in bubble sizes are due to detergent effect, the same sinter generator and pipe layout being employed. Scale = x 1.3

Figure E.5: Photographs of 17.3 mm pipe, with a constant air flowrate, $V_A = 0.59 \text{ m/s.}$ Sinter generator, 0.05% and 0.5% Teepol solutions.

Appendix F:

Flow regime pattern maps, Data symbol notation.

A range of symbols were used on the flow regime maps to represent the various flow patterns which could exist in the pipe. The symbols were of two main types,

a) A solid symbol, which indicated that the particular flow pattern had been definitely identified at that point.

b) An open symbol, which indicated that the flow pattern which had been defined at that point was of some intermediate regime. The symbol used is that of the prime regime, if the flow was defined as stratified with slugging tendency, the open symbol for the stratified regime is plotted.

<u>Flow regime</u>	<u>Abbreviation</u>	<u>Symbol employed.</u>	
		<u>Definite</u>	<u>Indeterminate</u>
Dispersed	DS	▼	▽
Stratified-wavy	ST-W	■	□
Stratified-smooth	ST-S	▣	▢
Foam	FM	▲	△
Slug	SL	●	○
Elongated bubble	EB	⬤	□

The abbreviations are used in the tabulated data in Appendix I, where the flow regime at the various phase flowrate are indicated.

The accuracy of the flow regime identification depends on the particular flow conditions. Identification of the Slug regime with detergent solutions was positive, the case with pure water flows was subject to a much greater degree of uncertainty. The stratified to dispersed regimes was also similarly difficult to define uniquely .

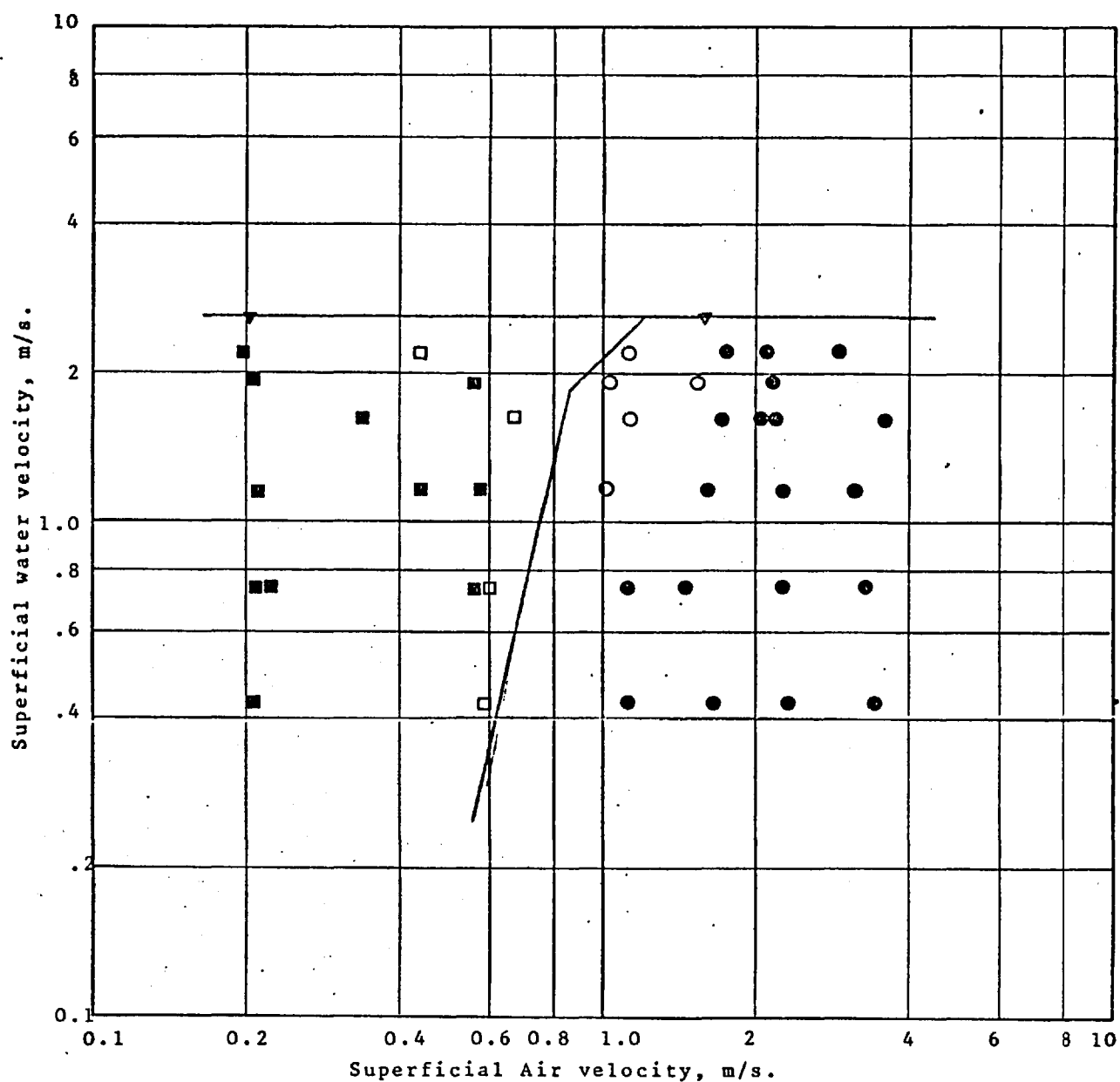


Figure F.1: Flow regime map for 17.3 mm pipe, with pure water/air flows. Sinter generator.

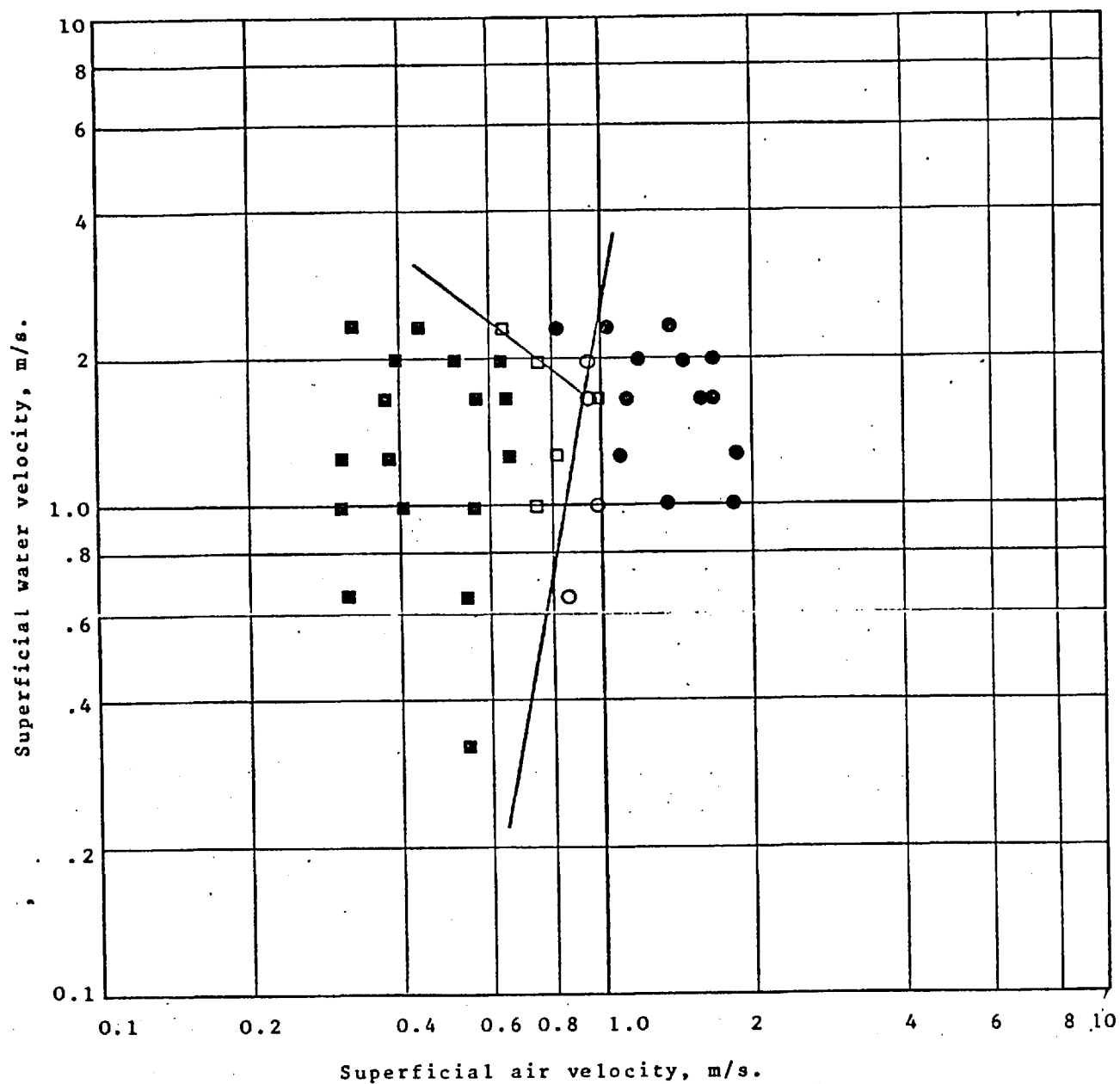


Figure F.2: Flow regime map for 11.4 mm pipe, with pure water/air flows, Sinter generator.

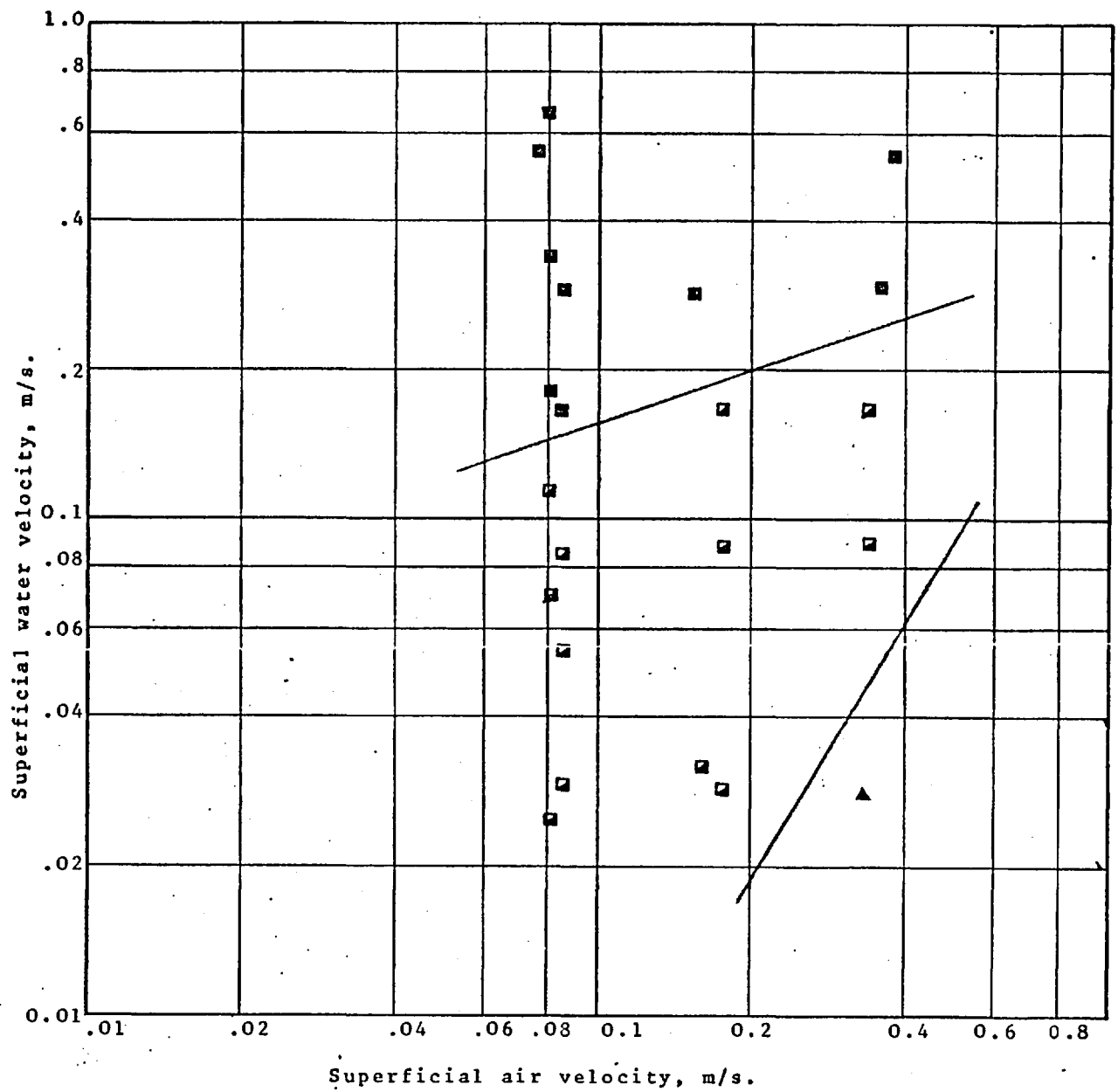


Figure F.3: Flow regime map for 39.1 mm pipe, with 0.5% Teepol solution, Sinter generator.

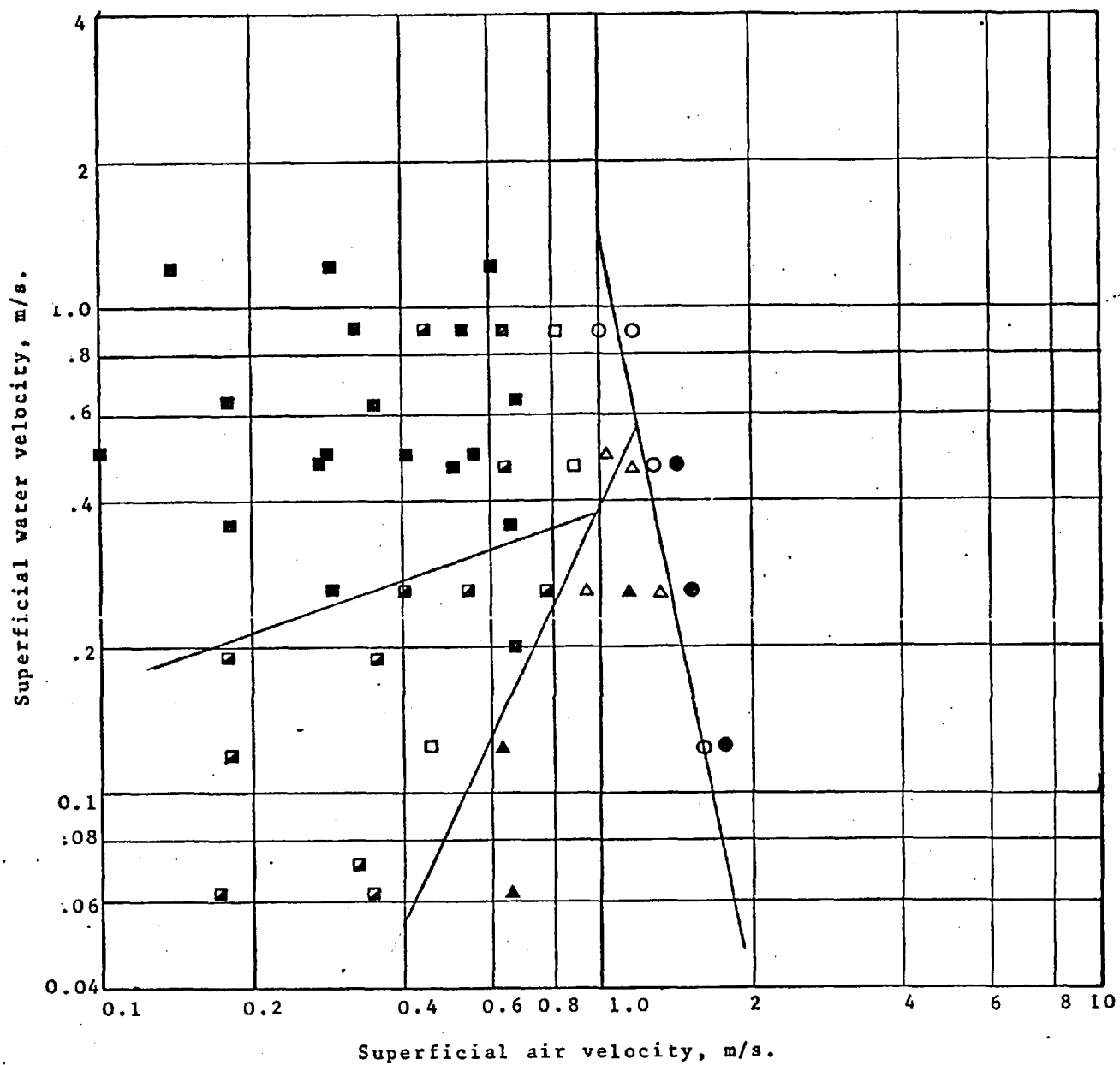


Figure F.4: Flow regime map for 26.0 mm pipe, 0.5% Teepol solution, Sinter generator.

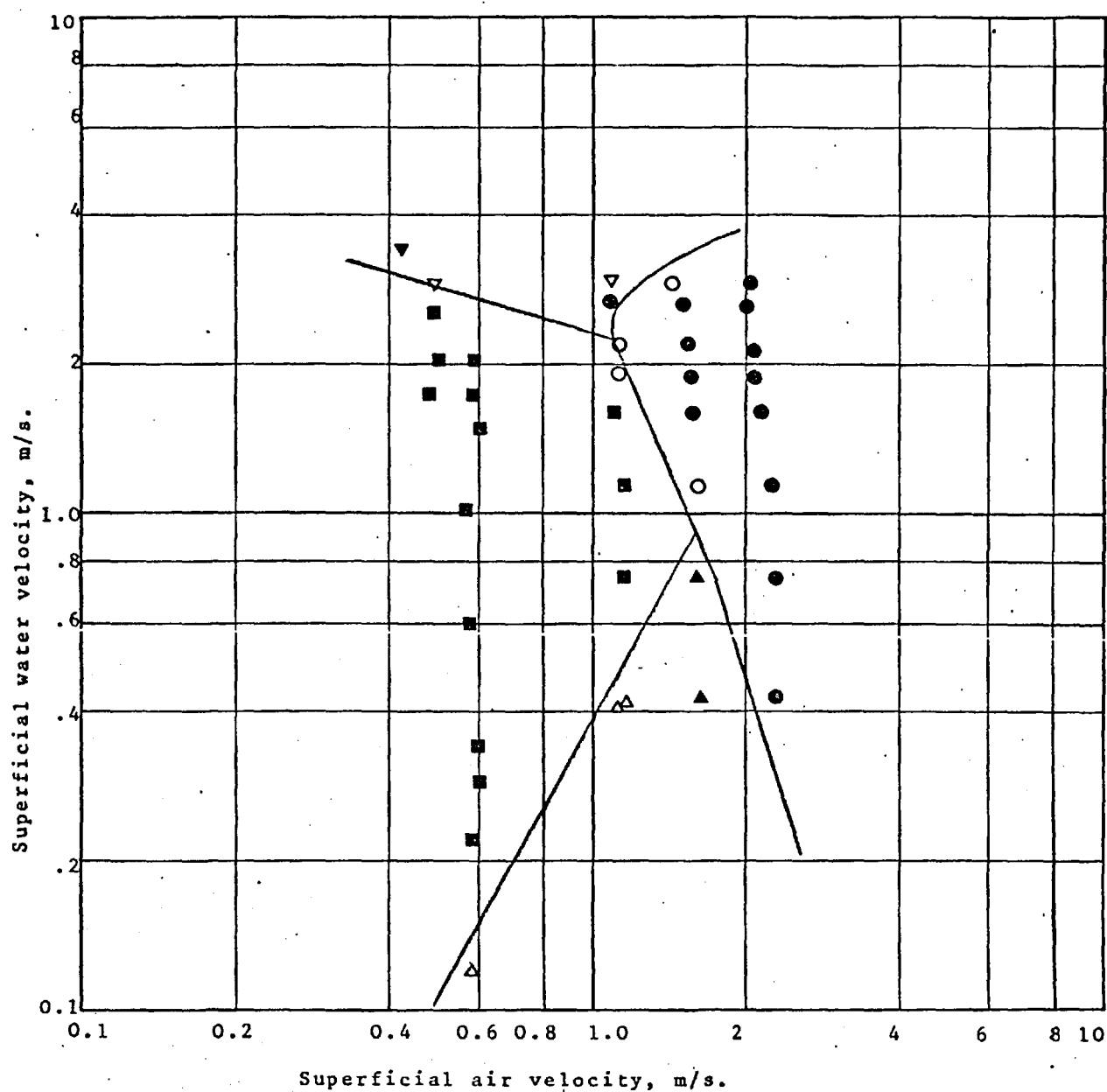


Figure F.5: Flow regime map for 17.3 mm pipe, with 0.05% Teepol solution, Sinter generator.

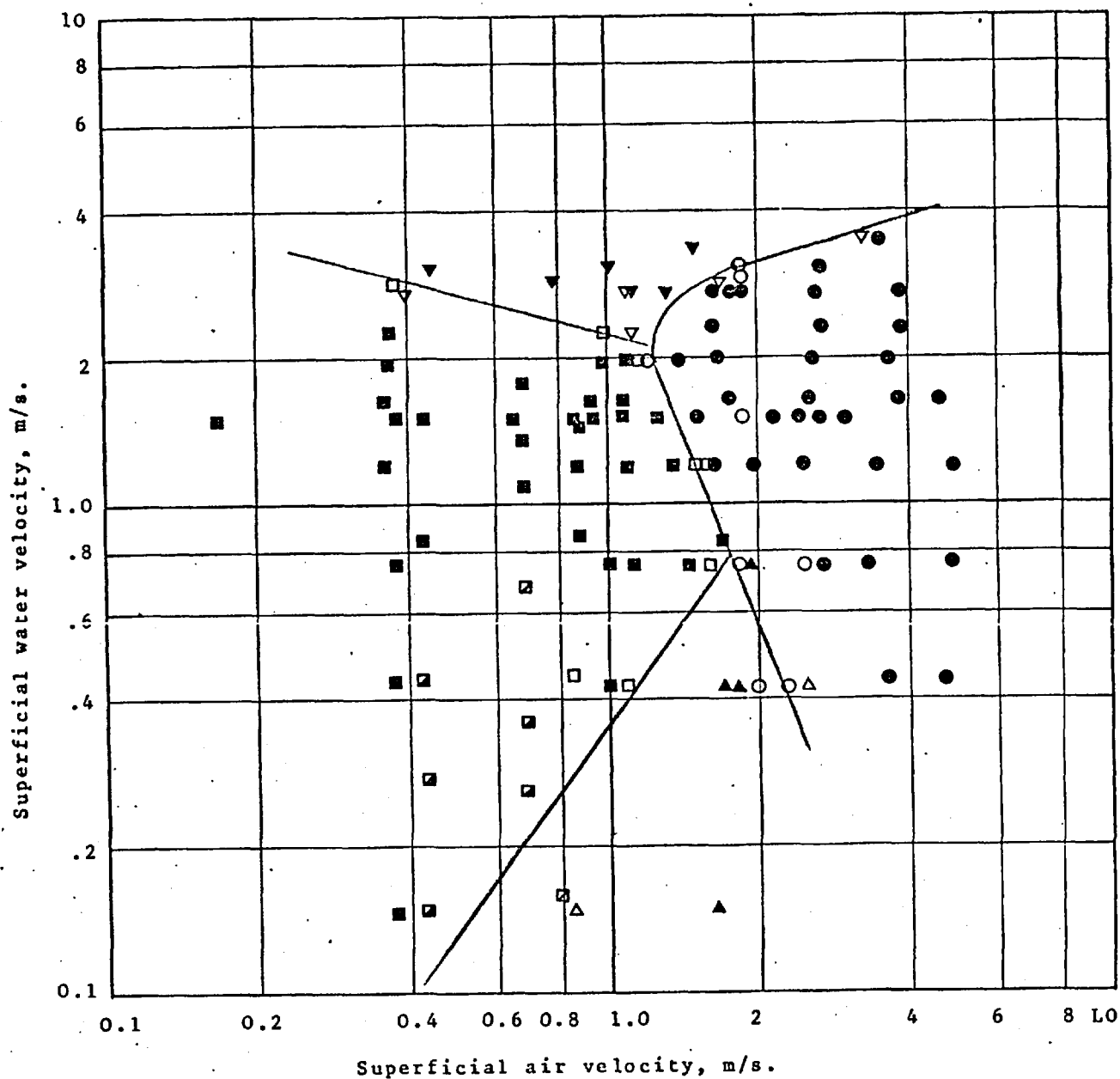


Figure F.6: Flow regime map for 17.3 mm pipe, with 0.5% Teepol solution, Sinter generator.

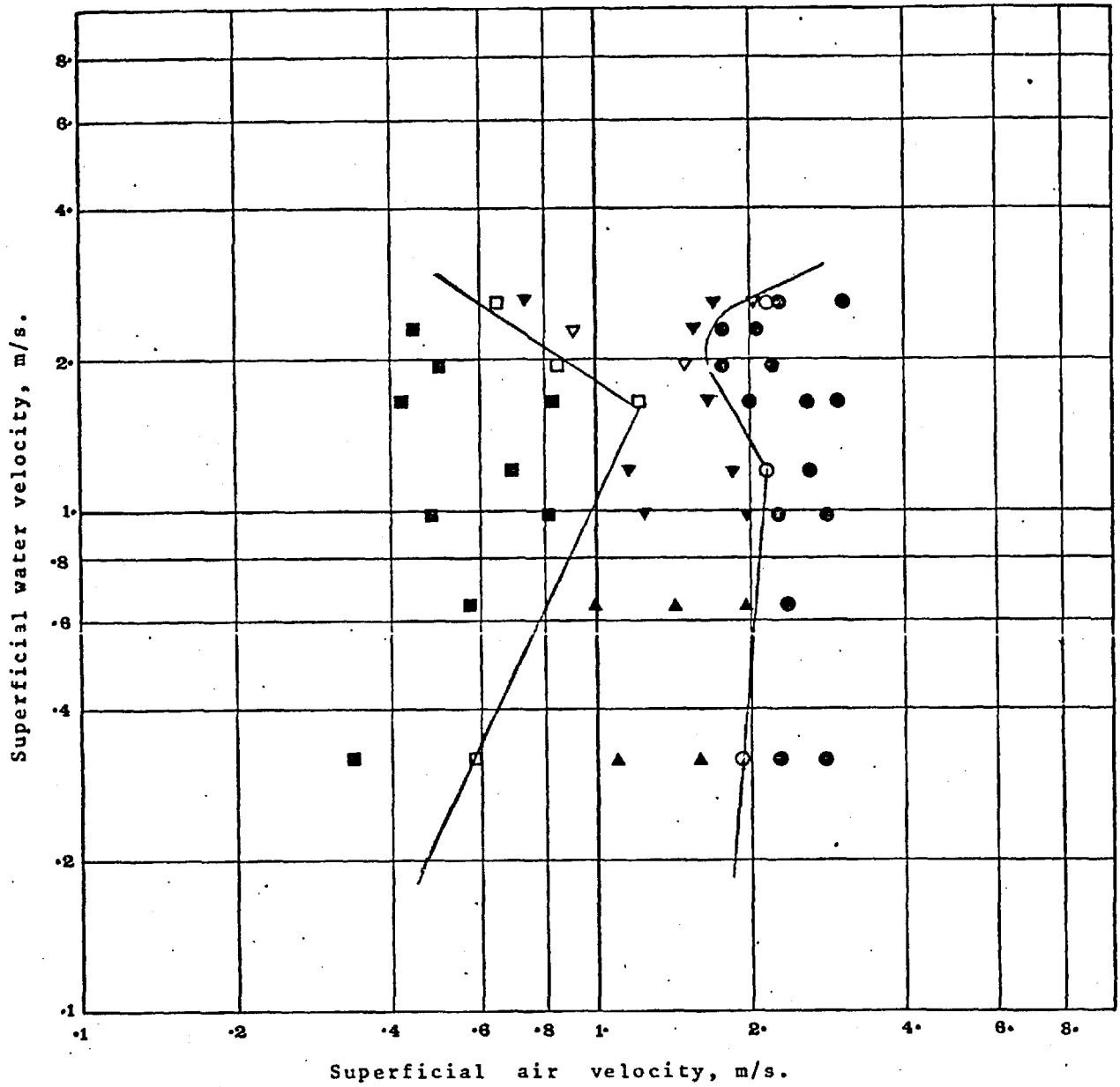


Figure F.7: Flow regime map for 11.4 mm pipe, with 0.05% Teepol solution, Sinter generator.

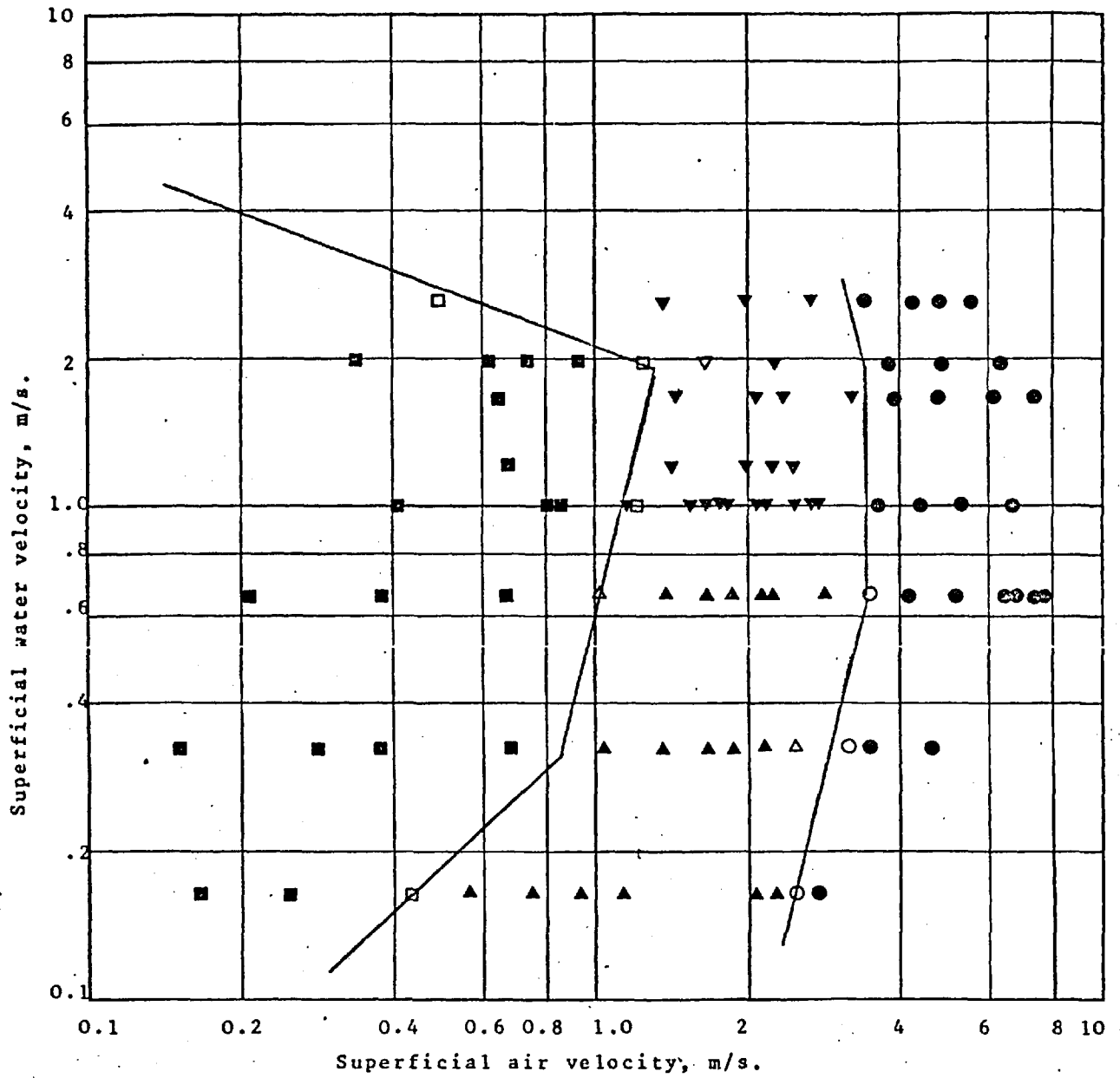


Figure F.8: Flow regime map for 11.4 mm pipe, with 0.5% Teepol solution, Sinter generator.

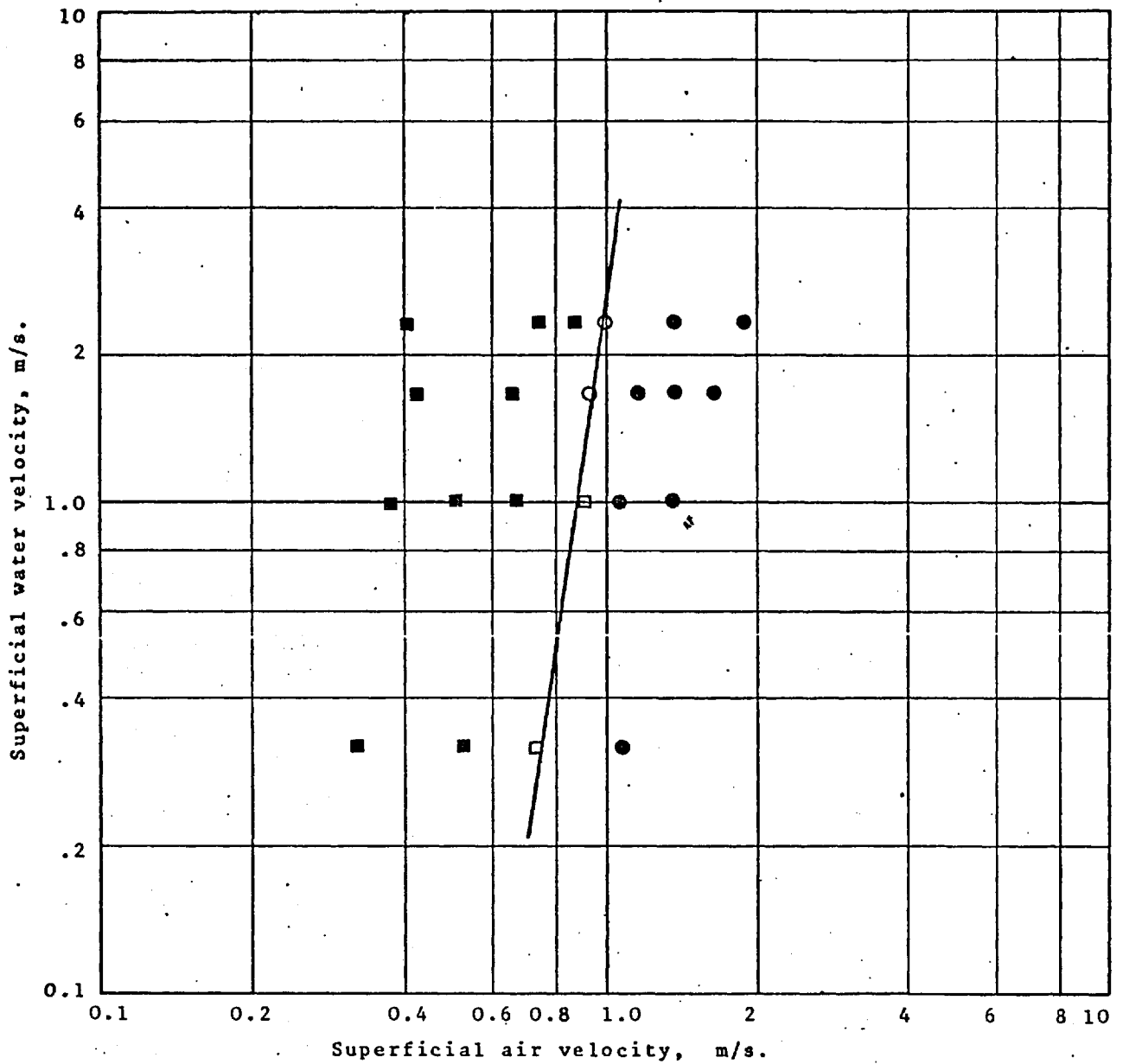


Figure F.9: Flow regime map for 11.4 mm pipe, with 0.05% Teepol Solution, T-piece generator.

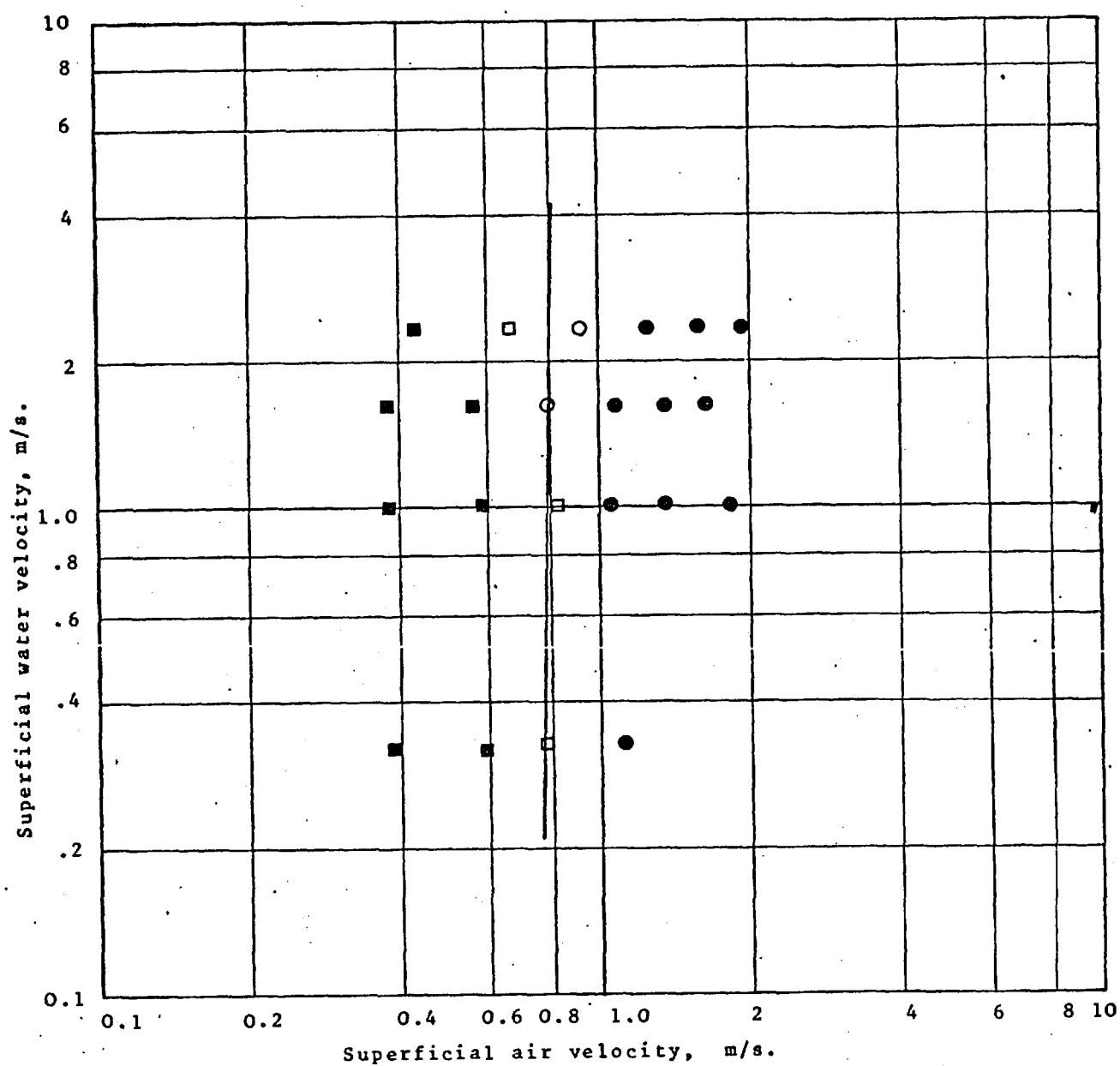


Figure F.10: Flow regime map for 11.4 mm pipe, with 0.5% Teepol solution, T-piece generator.

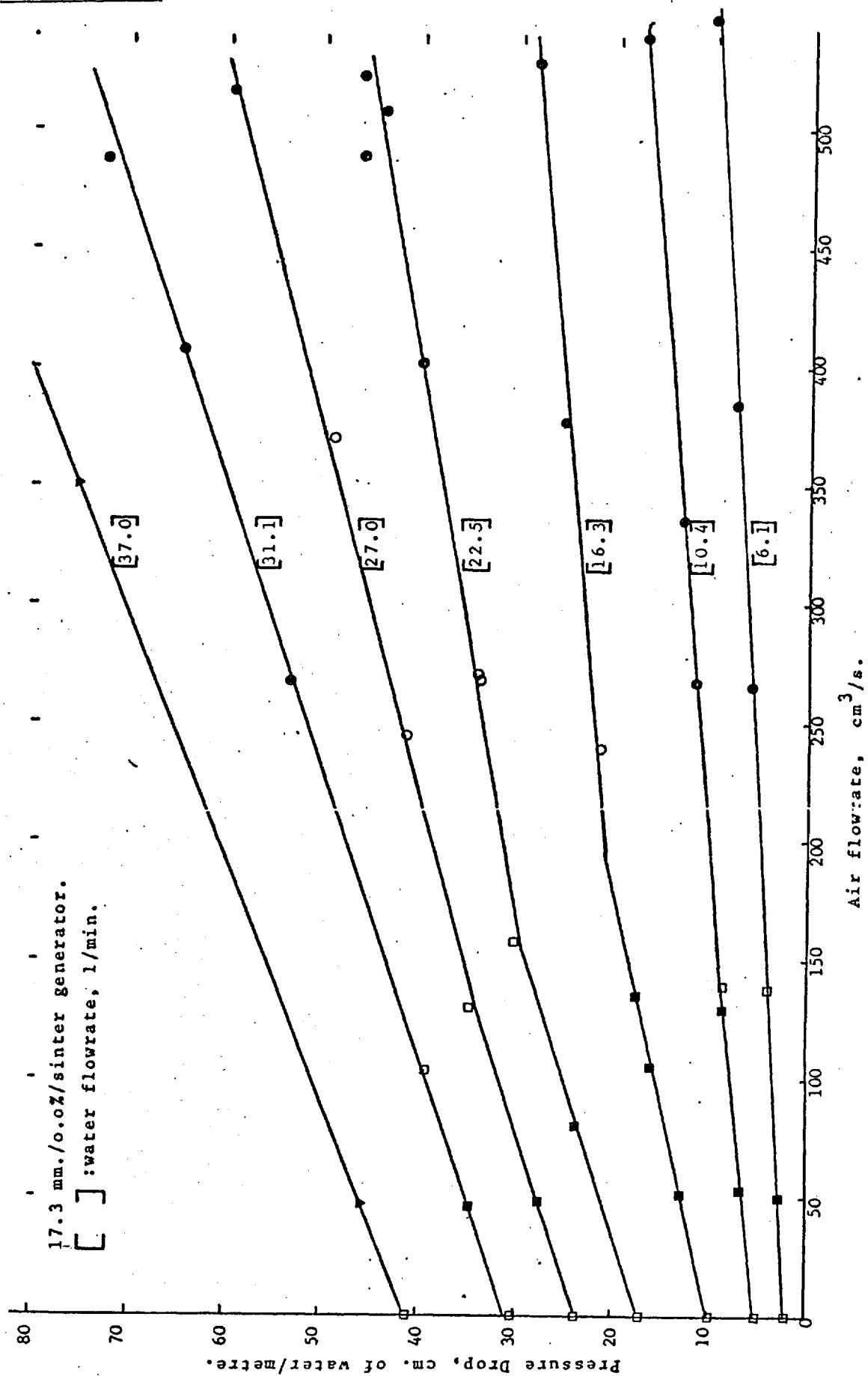


Figure G.1: Pressure drop results at constant water flowrate for the 17.3 mm pipe, with pure water/air flows.

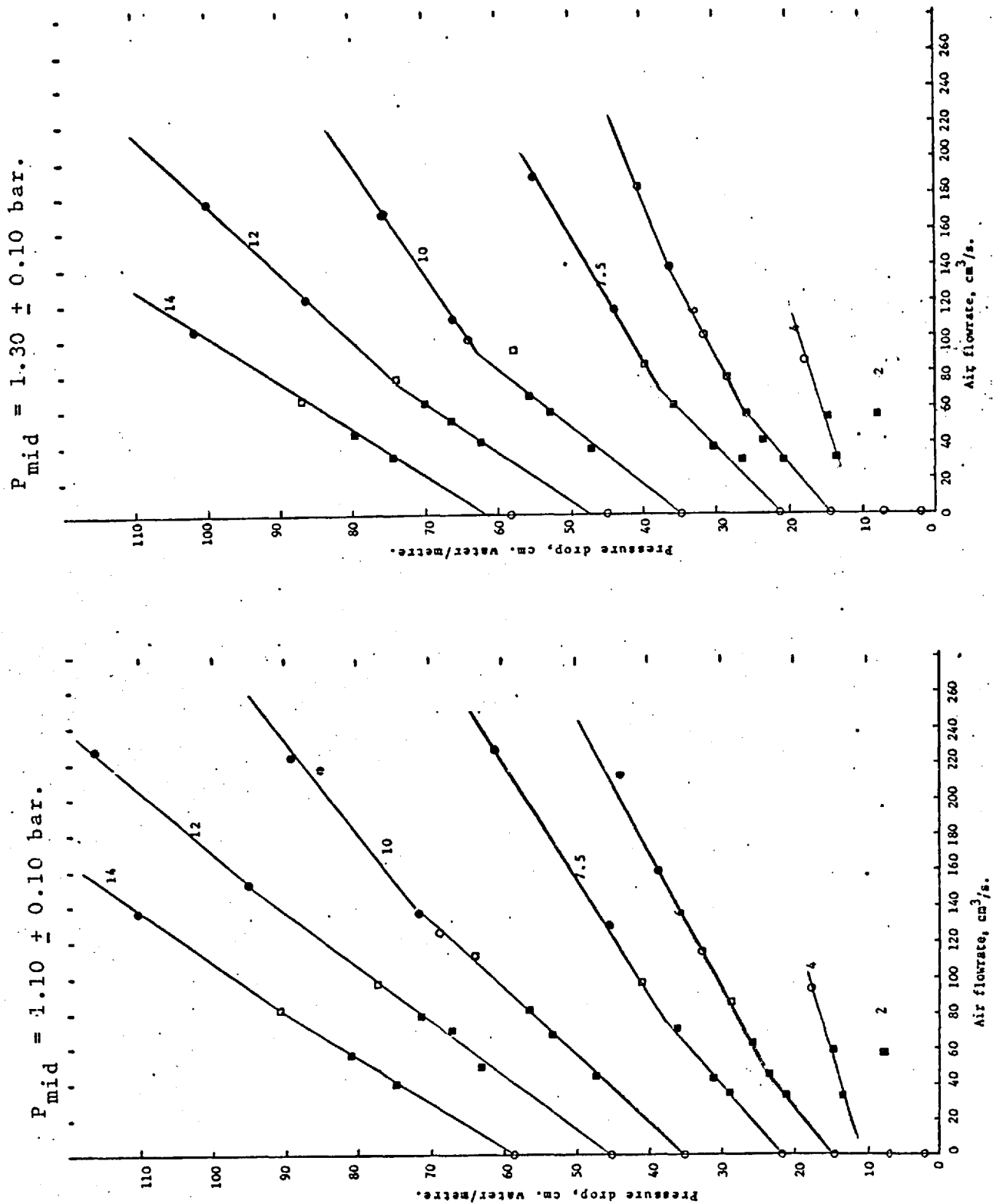


Figure G.2: Pressure drop results at constant water flowrate for the 11.4 mm pipe, with pure water/air flows. The results collected at P_{mid} of 1.10 and 1.30 bar.

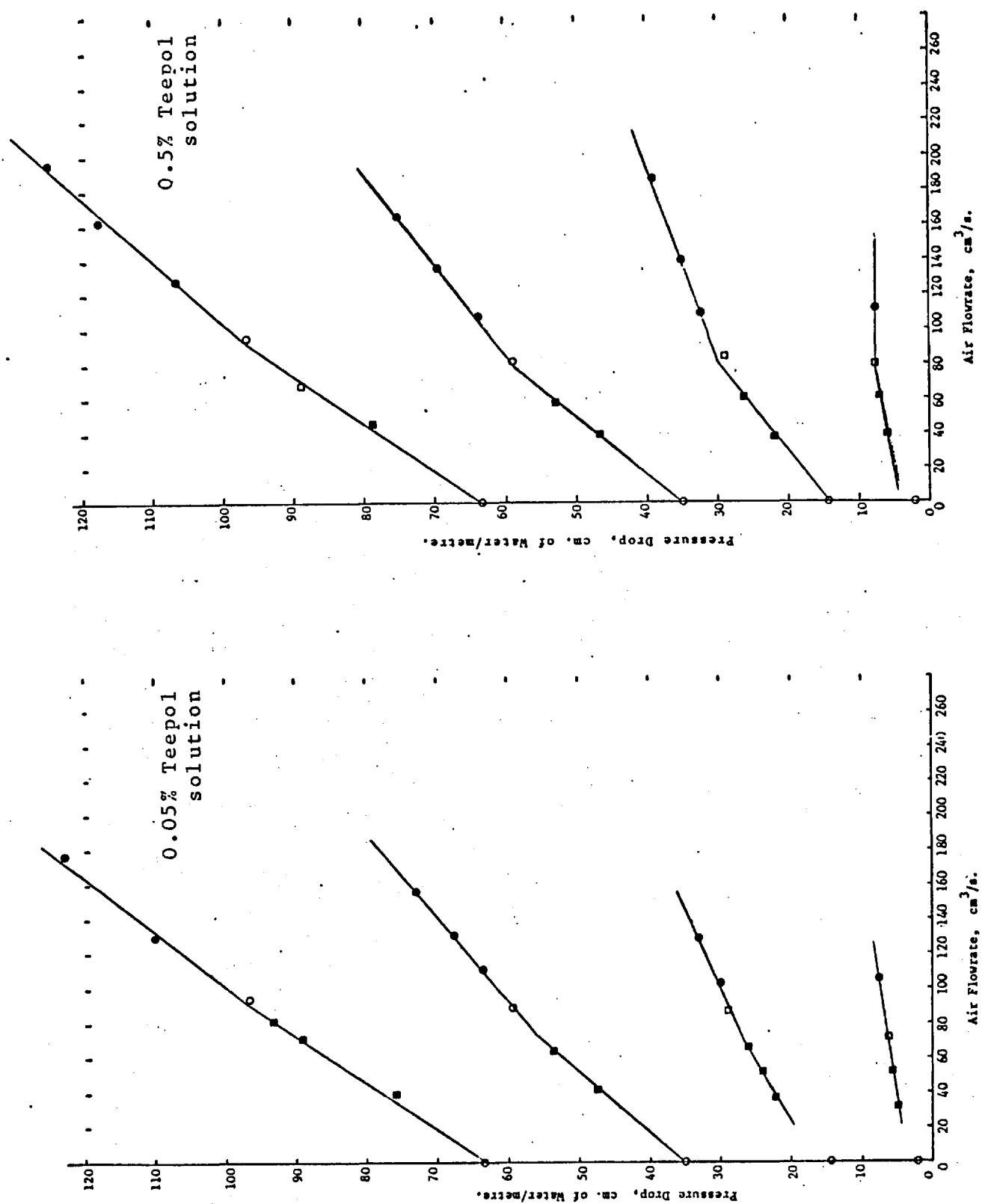


Figure G.3: Pressure drop results at constant water flowrate, for the 11.4 mm pipe, with 0.05% and 0.5% Teepol solutions, Using the T-piece generator.

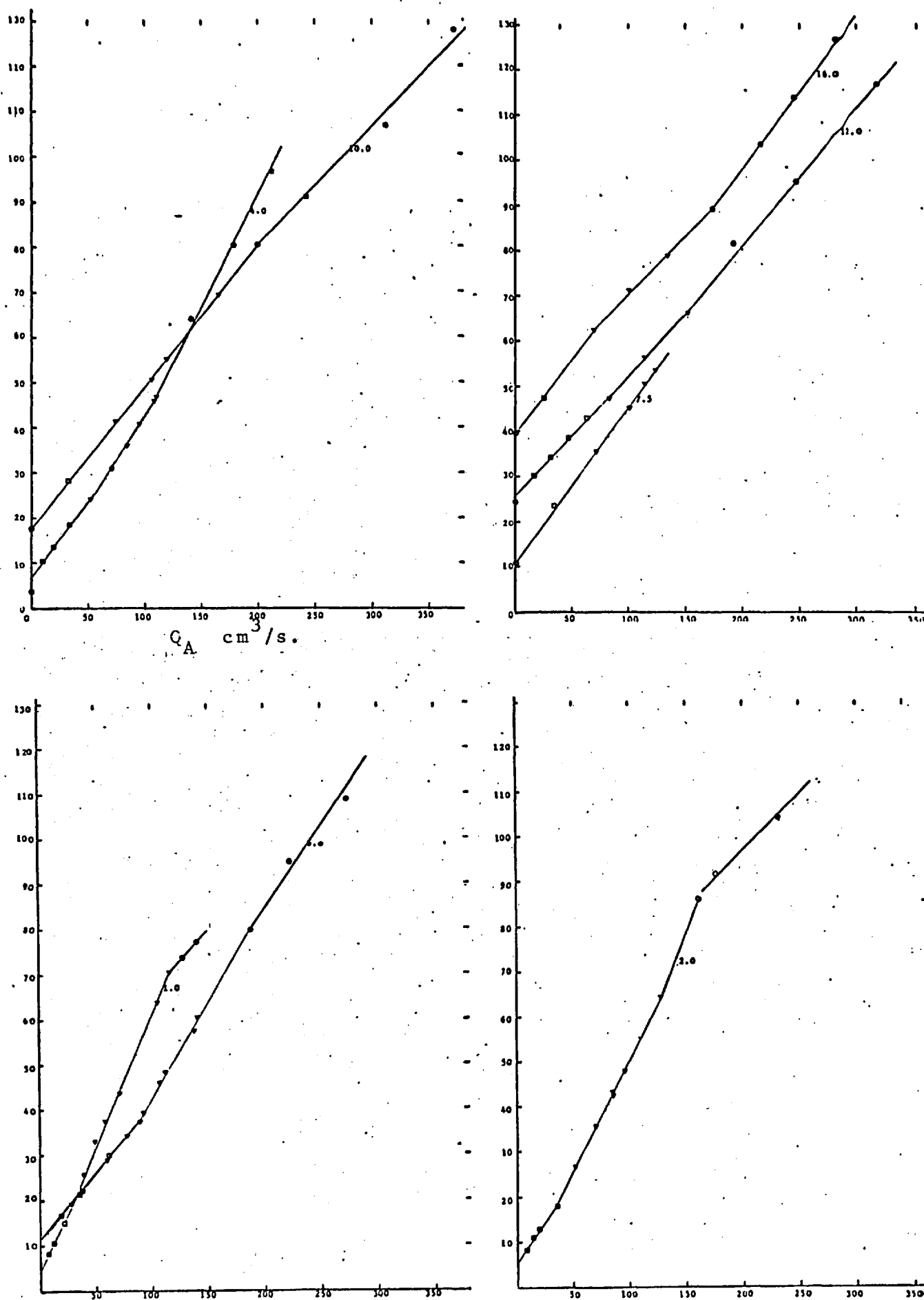


Figure G.4: Pressure drop results at constant water, 11.4 mm.
with 0.5% Teepol solution, Sinter generator.

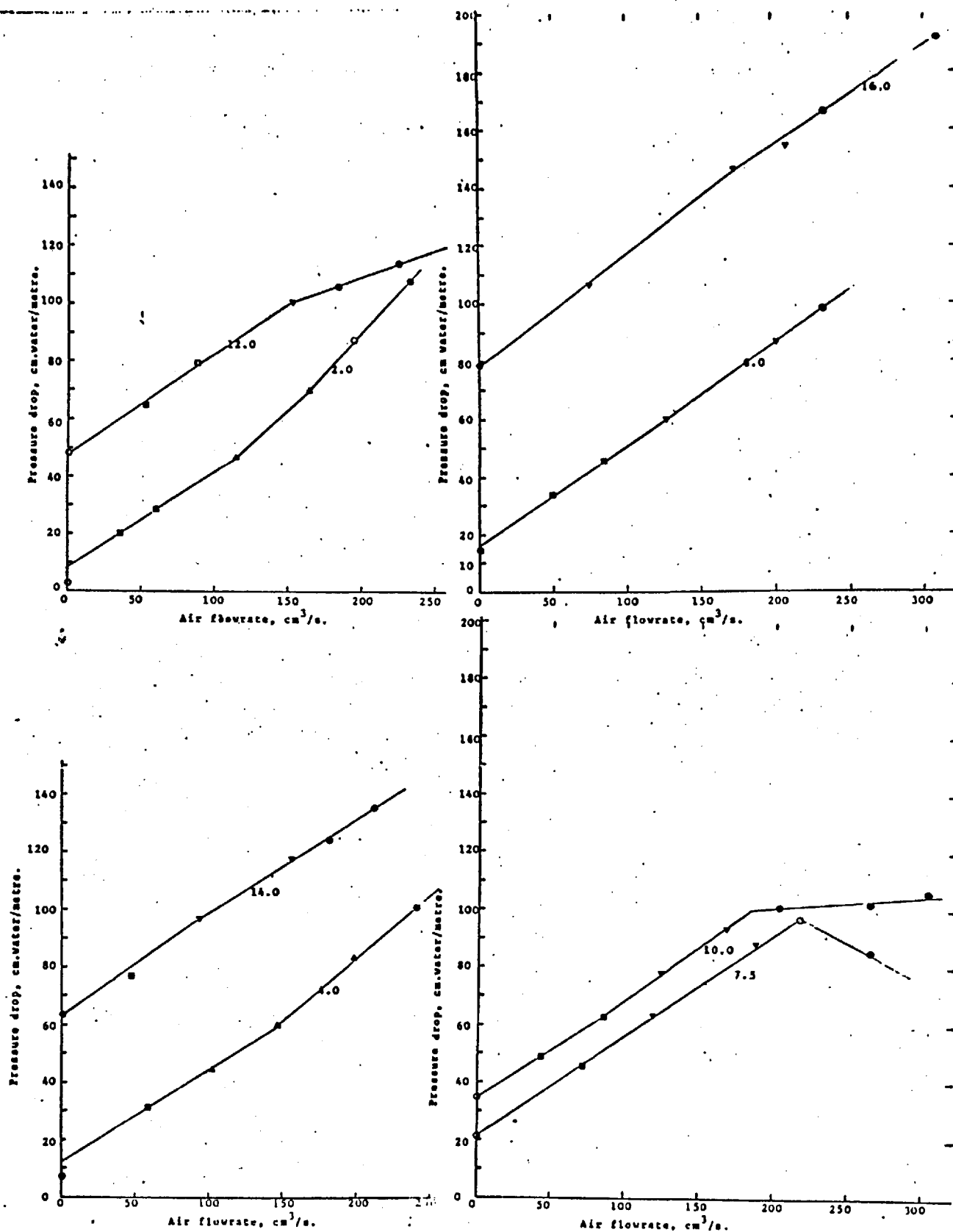


Figure G.5: Pressure drop results at constant water flowrate,
for the 11.4 mm pipe, with 0.05% Teepol solution,
sinter generator.

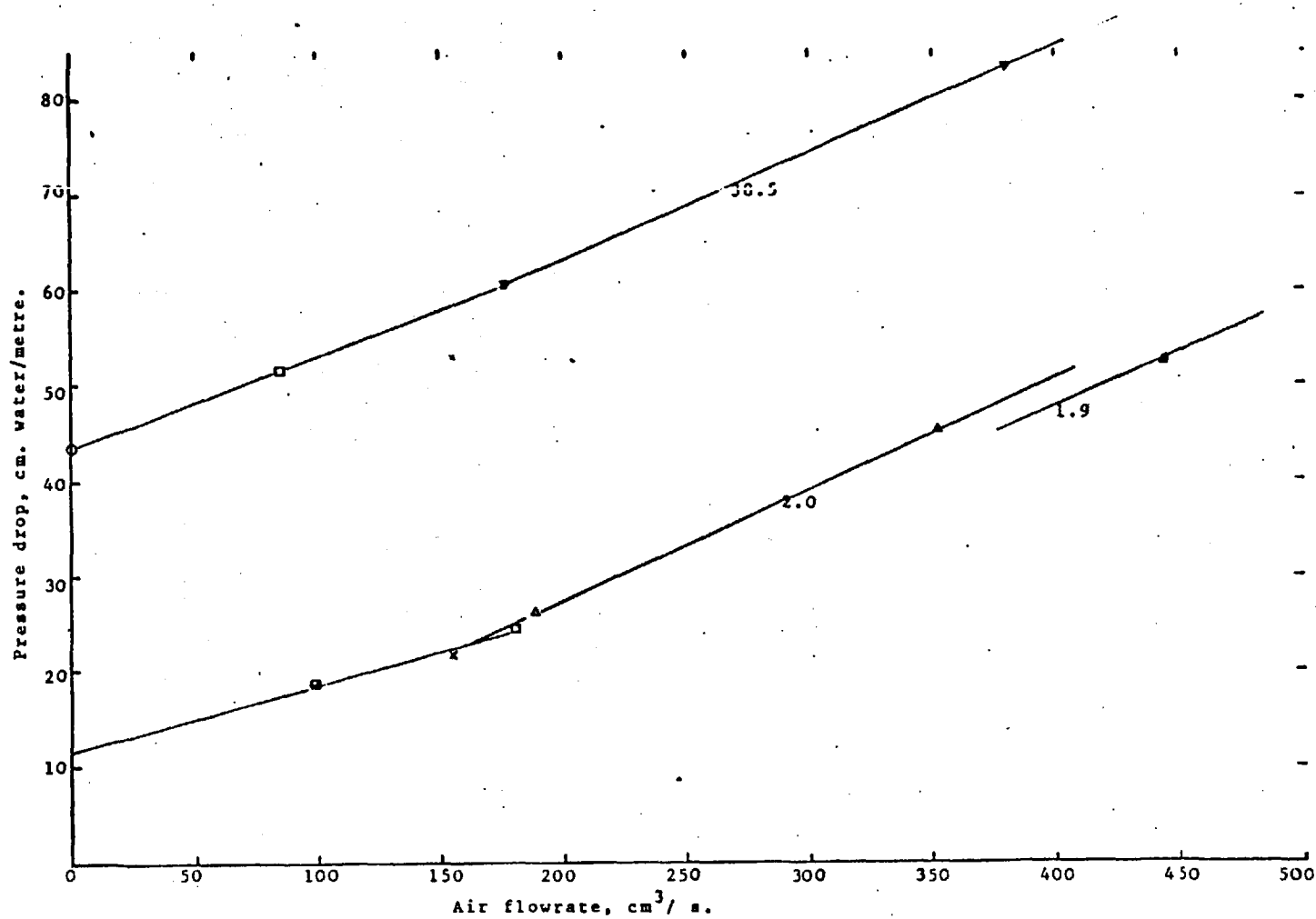
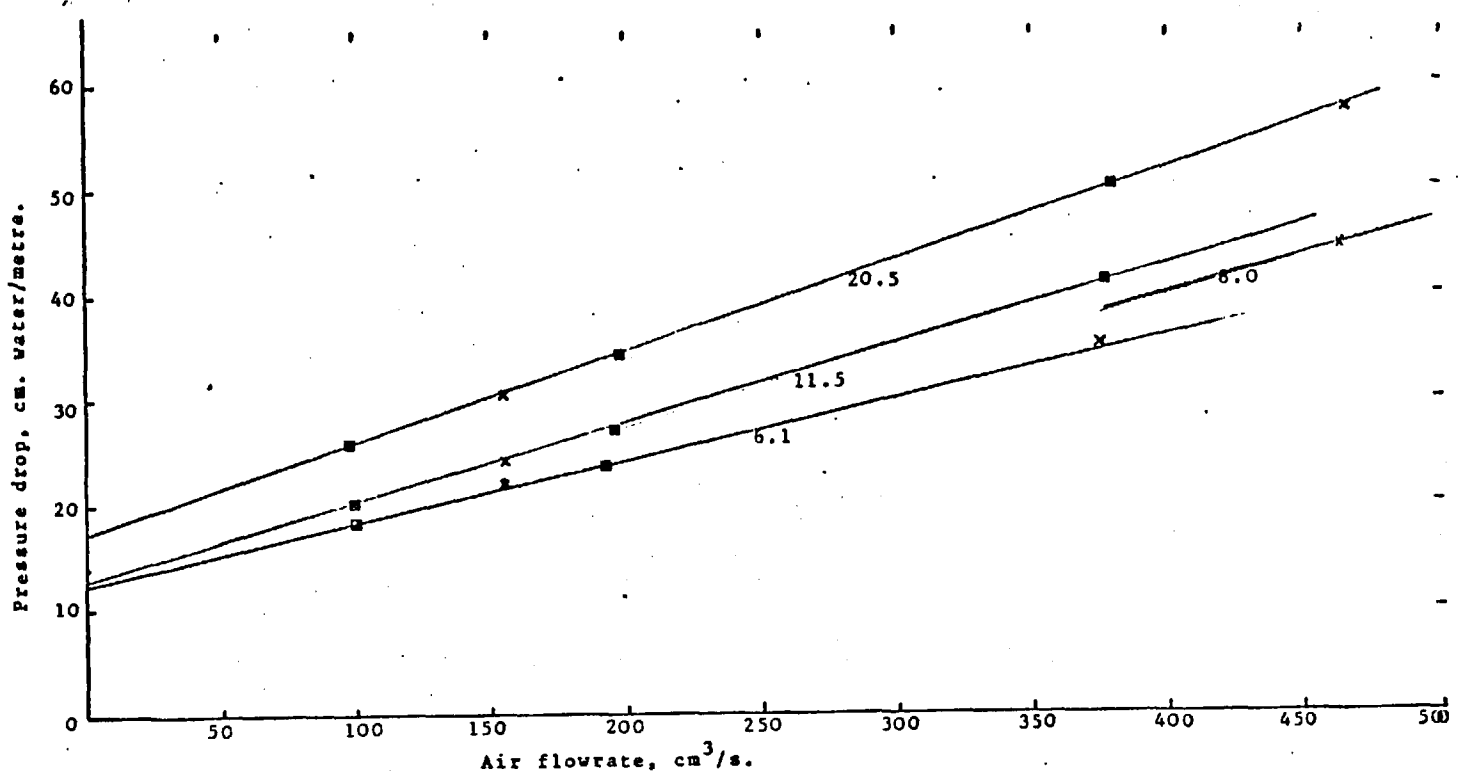


Figure G.6: Pressure drop results at constant water flowrate;

for the 17.3 mm pipe, with 0.5% Teepol solution,

Sinter generator.

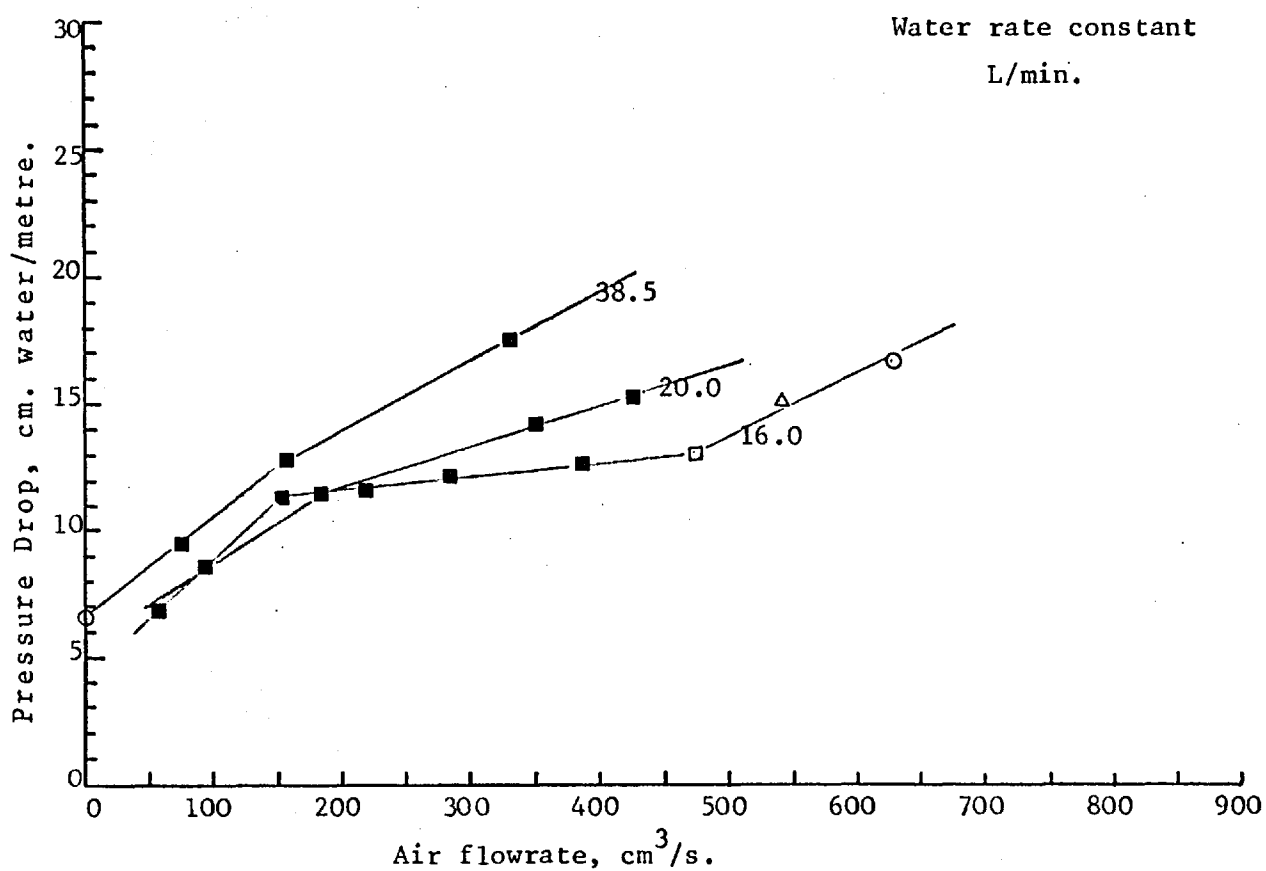
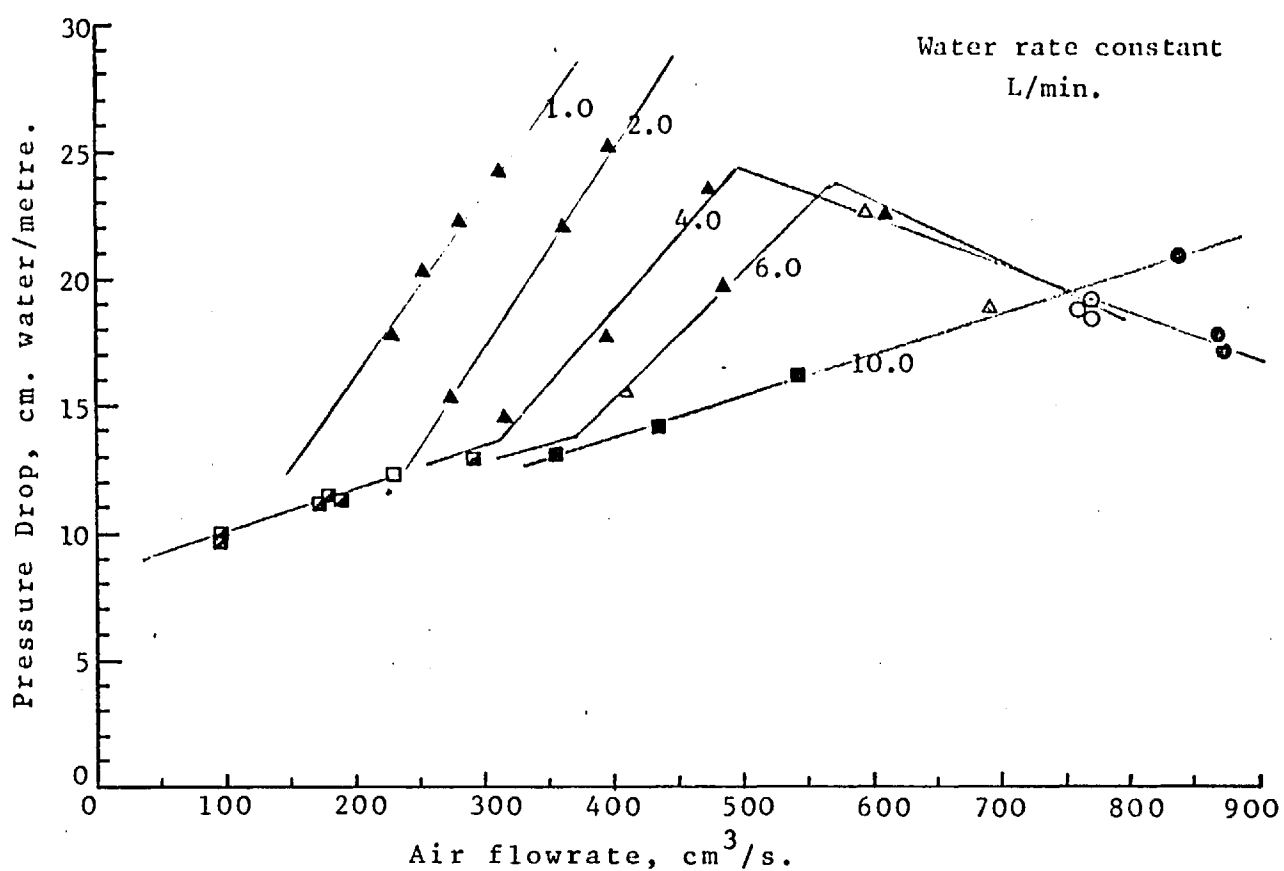


Figure G.7: Pressure drop versus Air flowrate, with water rate constant. 26.0 mm. pipe, 0.5% Teepol, Sinter generator.

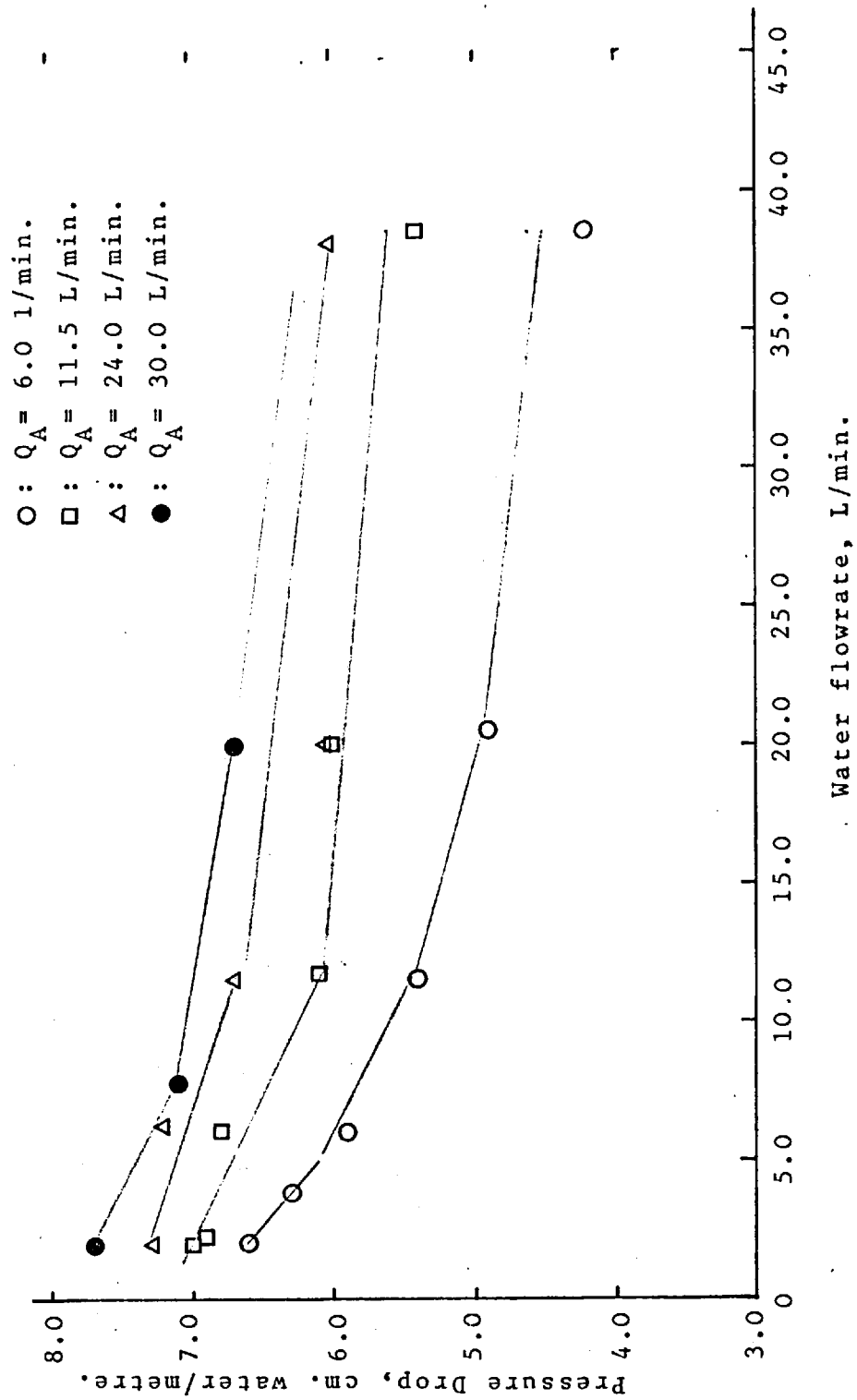


Figure G.8: Pressure drop results at constant air flowrates for the 39.1 mm pipe, with 0.5% Teepol solution, Sinter generator.

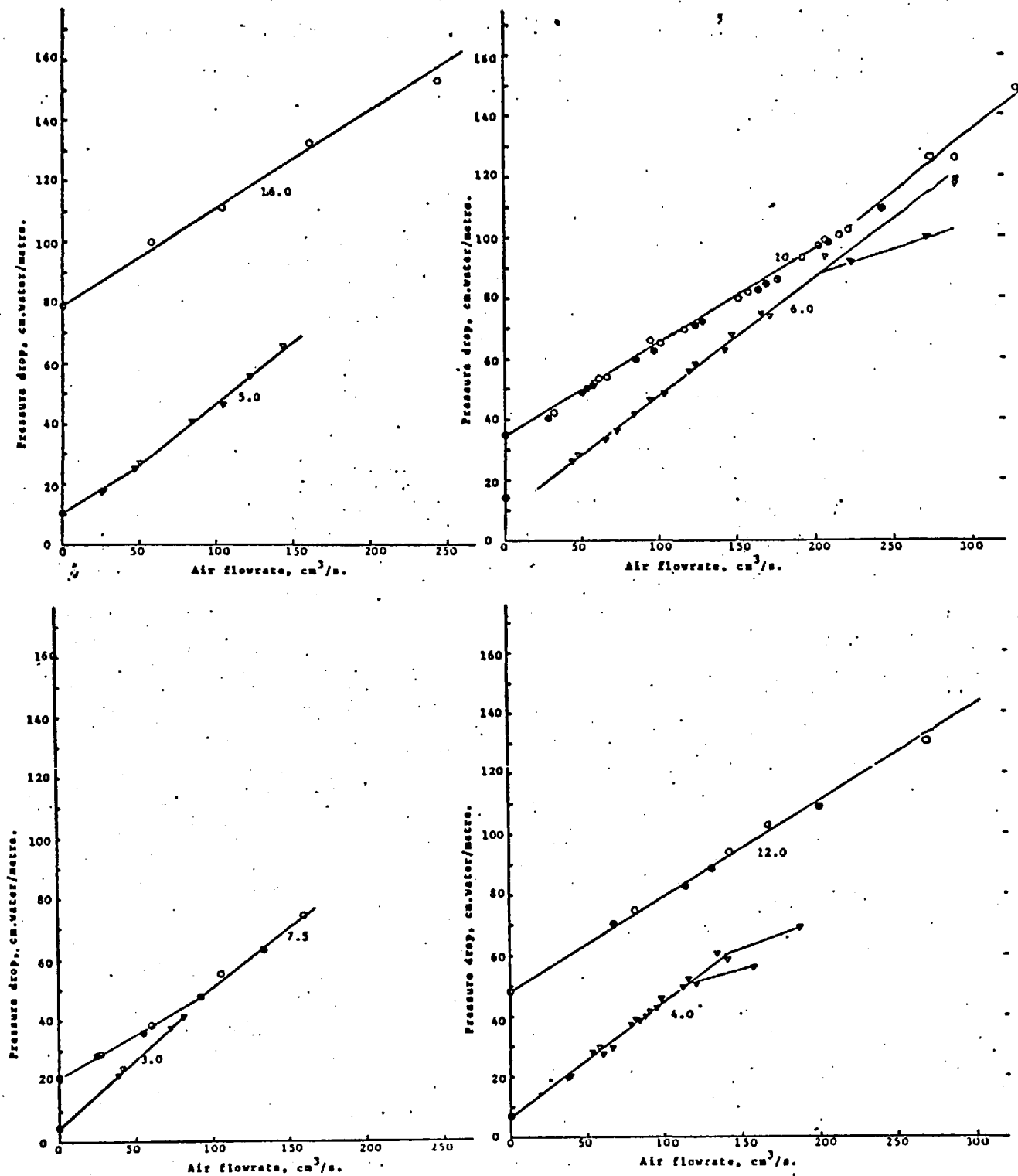


Figure G.9: Pressure drop results at constant water flowrate,
for the 11.4 mm pipe, with 0.5% Teepol solution,
Pump generator.

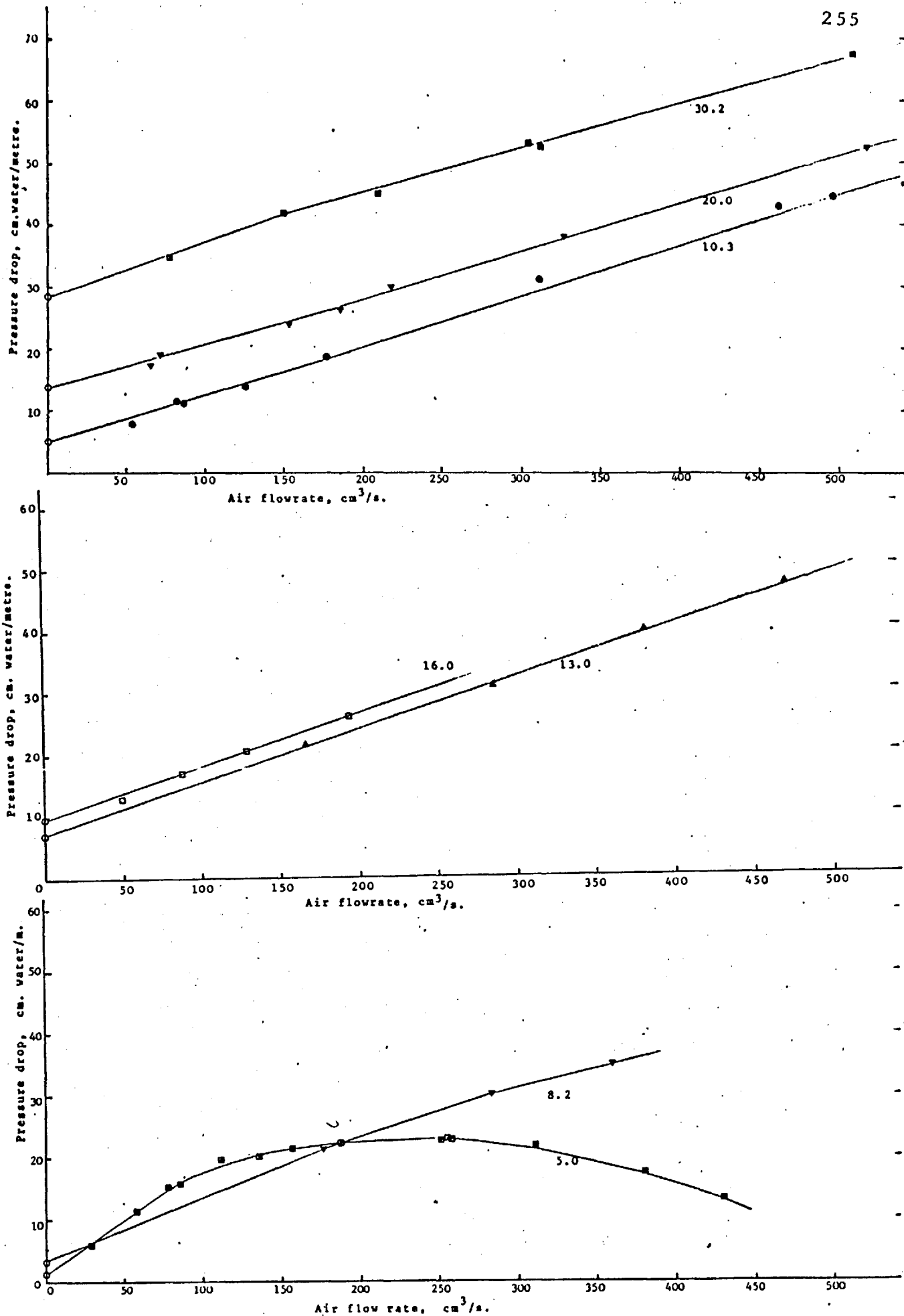


Figure G.10; Pressure drop results at constant water flowrate,
for the 17.3 mm pipe, with 0.5% Teepol solution,
Pump generator.

Table G.1: Air-pure water flows in 17.3 and 11.4 mm. pipes, at constant air/water ratios. Data taken from Figures G.1 and G.2 . (* : exrapolated value.)

1) Data for 17.3 mm. pipe, pure water, sinter generator.

Water flowrate l/min. cm^3/s .		Pressure Drop, cm. water/metre. at Air flowrates of,			
		$\frac{1}{2}Q_w$	$1Q_w$	$1\frac{1}{2}Q_w$	$2Q_w$
6.1	101.6	3.0	3.5	4.3	5.0
10.4	173.5	7.5	9.4	11.5	13.3
16.3	272.	17.6	22.5	25.7	28.8
22.5	375.	30.8	38.9	47.6	55.8*
27.0	450.	40.4	54.8	69.7*	
31.1	518.	52.4	73.9	100.4*	
37.0	617.	71.2	101.3*		

2) Data for 11.4 mm. pipe, pure water, sinter generator.

Water flowrate l/min. cm^3/s .		Pressure Drop, cm. water/metre. at Air flowrates of,			
		$\frac{1}{2}Q_w$	$1Q_w$	$1\frac{1}{2}Q_w$	$2Q_w$
2.0	33.3	3.6*	5.3*	7.0*	8.8
4.0	66.7	13.1	15.5	17.9	20.4*
6.0	100.0	23.9	30.4	37.1	43.7
7.5	125.	34.8	45.2	54.1	64.2*
10.0	167.	56.6	77.4	93.5*	
12.0	200.	78.0	108.8		
14.0	233.	102.8			

3) Data for 11.4 mm. pipe pure water, sinter generator.

Water floerate l/min. cm^3/s .		Pressure Drop, cm. water/metre. at Air flowrates of,			
		$\frac{1}{2}Q_w$	$1Q_w$	$1\frac{1}{2}Q_w$	$2Q_w$
4.0	66.6	13.6	16.0	18.4	20.9*
6.0	100.0	24.4	31.8	37.3	42.1
7.5	125.	35.8	45.4	54.3	63.0*
10.0	167.	60.3	75.0	87.6*	
12.0	200.	80.4	106.6		
14.0	233.	105.7			

Ambient pressures in pipe at midpoint of tappings were,

1.10 $\pm$ 0.10 bar for 1) and 2).

1.30 $\pm$ 0.20 bar for 3).

Table G.2: Air-Detergent flows in pipes, at constant air/liquid ratios. Sinter generated foams, 0.5% Teepol solution. (* : extrapolated value).

1) Data for 11.4 mm pipe, 0.5% solution, sinter generator.

Water flowrate l/min. cm ³ /s.		Pressure drop, cm. water/metre at Air flowrates of,			
		$\frac{1}{2}Q_W$	$1Q_W$	$1\frac{1}{2}Q_W$	$2Q_W$
1	16.7	13.5	17.6	20.8	25.0
2	33.3	18.2	22.9	28.7	34.3
4	66.7	24.5	35.8	47.2	57.6
6	100.	36.9	51.8	67.1	84.8
7.5	125.0	42.1	64.3	86.0	107.6
10.0	167.0	61.2	87.1	113.2	139.6
12.0	200.	77.4	103.9	131.7	160.4
16.0	267.	122.0	157.	195.6	241.3

Value of } $\Delta P @ Q_F$	0	10.2	12.9	17.5	15.0
	80	22.0	25.2	30.8	28.0

2) Data for 17.3 mm pipe; 0.5% solution, sinter generator.

Water flowrate l/min. cm ³ /s.		Pressure drop, cm. water/metre at Air flowrates of,			
		$\frac{1}{2}Q_W$	Q_W	$1\frac{1}{2}Q_W$	$2Q_W$
2.0	33.3	12.7	13.8	15.1	16.3
6.1	102.	15.5	18.6	21.6	24.6
8.0	133.3	16.8*	21.4*	26.2*	30.8*
11.5	192	19.8	26.7	34.1	41.5
20.5	342	32.1	46.9	61.8	76.6*
38.5	642	76.8	92.5*	-	-

Value of } $\Delta P @ Q_F$	0	11.5	11.9	12.1	12.4
	160	15.8	17.2	18.0	19.1

3) Data for 26.0 mm pipe, 0.5% solution, sinter generator.

Water flowrate l/min. cm ³ /s.		Pressure drop, cm. water/metre at Air flowrates of,			
		$\frac{1}{2}Q_W$	$1Q_W$	$1\frac{1}{2}Q_W$	$2Q_W$
2	34	-	-	9.3	9.5
4	67	7.7*	8.7	9.5	10.2
6	100	8.2 ⁺	9.4	10.4	11.5
8.2	137	-	-	11.3	
10.0	167	8.1	10.5	12.1	13.0
11.5	192	-	11.2	13.4	-
16.0	267	10.5	12.1	13.0	14.5
20.0	334	11.1	13.8	16.5	19.0*
38.5	642	17.5	-	-	-

Value of ΔP at Q_F	0	6.8	7.4	8.1	8.5
	500	11.1	12.3	13.0	13.6

4) Data for 39.1 mm pipe; 0.5% solution; sinter generator.

Air flowrate l/min cm ³ /s.		Pressure drop, cm of water/metre at water flow of,			
		$\frac{1}{2}Q_A$	$.66Q_A$	$1Q_A$	$2Q_A$
6	100	6.45	6.3	6.0	5.4
11.5	192	6.65	6.5	6.15	5.9
30.0	500	6.90	6.70	6.45*	

Value of ΔP at Q_F	0	6.40	6.27	5.91	5.25
	500	6.71	6.50	6.20	5.53

Note: data was plotted P_v 's Q_W for 39.1 mm pipe, results have been quoted at constant Q_A .

To obtain values of Q_F at any $\frac{Q_A}{Q_W}$ ratio,

$$Q_F = Q_W + \frac{Q_A \cdot Q_W}{Q_W}$$

and the corresponding ΔP value read off the chart at $Q_W; \frac{Q_A \cdot Q_W}{Q_W}$

Table G.3: Air detergent flows in pipes at constant air/liquid ratios. Sinter generated foams, 0.05% Teepol solution.

Data for 11.4 mm pipe- 0.05% solution- sinter generated.

Q_W l/min	Q_W $\text{cm}^3/\text{s.}$	Pressure drop, at Q_A equal to			
		$\frac{1}{2}Q_W$	Q_W	$1\frac{1}{2}Q_W$	$2Q_W$
2	33.3	14.0	19.4	24.9	30.4
4	66.7	22.7	33.9	44.1	55.3
6	100.	33.0	50.1	67.9	94.1
7.5	125.	43.5	64.8	87.1	88.8
10.0	167.	63.9	93.8	102.5	108*
12.0	200.	82.6	109.2	127.5*	-
14.0	233.	103.8	142.1	-	-
16.0	267.	130.	178.2	-	-

Value of ΔP at Q_F	0	5.1	6.4	7.9	10.0
	100	22.1	26.5	28.9	30.5

Table G.4: Air detergent flows in pipes, at constant air/liquid ratios. Pump generated foams, 0.5% Teepol solution.
(* extrapolated value).

1) Data for 11.4 mm pipe, 0.5% solution, pump generator.

Water flowrate l/min. cm ³ /s.		Pressure drop, cm. water/metre at Air flow rates of			
		$\frac{1}{2}Q_W$	Q_W	$1\frac{1}{2}Q_W$	$2Q_W$
3	50.0	15.8*	27.0	38.5	49.5*
4	66.7	19.1	31.9	44.7	57.2
5	83.5	23.2	39.7	56.4	73*
6	100.	28.4	47.9	67.4	86.8
7.5	125.	39.0	60.3	82.7	-
10.0	167.	60.3	86.4	115.	150*
12.0	200.	79.5	110.7	142.5*	-
16.0	267.	122.	165*	-	-

Value } $\Delta P @ Q_F$	0	6.8	14.2	20.0	25.5
	100	19.1	27.0	35.0	41.5

2) Data for 17.3 mm pipe, 0.5% solution, pump generator.

Q_W l/min. cm ³ /s.		Pressure drop, at Q_A equal to		
		$\frac{1}{2}Q_W$	Q_W	$1\frac{1}{2}Q_W$
5.0	83.3	8.3	15.5	19.6
8.2	137	10.2	17.2	23.6
10.3	172.	11.6	18.0	35.0
13.1	218.	16.4	25.6	34.5
16.0	267.	20.8	32.2	43.4*
20.0	333.	25.4	38.3	51.4
30.1	502	49.0	67.0	85.1**

Value of } $\Delta P @ Q_F$	0	5.5	13.3	14.5
	200	10.3	16.0	19.6

- 3) Foam flow in 17.3 mm pipe, flow division at T-piece,
0.5% solution, pump generator.

$\frac{Q_A}{Q_{Wfeed}}$	$\frac{Q_A}{Q_{Wexit}}$	Q_F	$\Delta P / L$
.50	.57	76.7	8.1
.51	.42	82.6	8.8
.51	.51	115.	9.3
.50	.48	160.	10.1
.50	.51	157.	10.5
.50	.49	232.	12.7
.52	.50	330.	16.0
1.00	1.00	93.4	13.5
1.01	.97	110.	14.3
1.01	.95	168	15.4
1.04	.99	172	15.8
1.03	.99	214.	17.4
1.04	.97	251.	17.5
1.03	.98	292.	19.2
1.03	.98	300.	19.5
1.02	.96	318	19.7

Value of ΔP at Q_F	.5	0	5.9
	.5	200	11.8
	1.0	0	10.9
	1.0	200	16.6

Table C.5: Friction factor v's Reynolds number for foams.

 Q_f in cm^3/s . $\Delta P/L$ in (cm. of water) / metre.
A) Pump generated foams.1) 11.4 mm pipe, 0.5% detergent solution.

a) $\frac{Q_A}{Q_W} = 0.5$; $\beta_f = 0.67 \text{ g.cm}^{-3}$

$$\mu_a = 0.050 \text{ poise}; \tau_o = 1.91 \text{ Nm}^{-2}$$

$$\text{He} = 1663$$

b) $\frac{Q_A}{Q_W} = 1.0$; $\beta_f = 0.5 \text{ g.cm}^{-3}$

$$\mu_a = 0.052 \text{ poise}; \tau_o = 3.98 \text{ Nm}^{-2}$$

$$\text{He} = 2391$$

Q_f	$\Delta P/L$	f	Re
75	15.8	.0249	1106
100	19.1	.0169	1474
125	23.2	.0131	1843
150	28.4	.0112	2211
183	39.0	.0101	2698
250	60.3	.00854	3686
300	79.5	.00782	4423
400	122.	.00675	5897

Q_f	$\Delta P/L$	f	Re
100	27.0	.0316	1074
133	31.9	.0211	1428
167	39.7	.0166	1793
200	47.9	.0140	2148
250	60.3	.0113	2685
333	86.4	.00911	3576
400	110.7	.00809	4297
533	165.0	.00679	5724

c) $\frac{Q_A}{Q_W} = 1.5$; $\beta_f = 0.4 \text{ g.cm}^{-3}$

$$\mu_a = 0.061 \text{ poise}; \tau_o = 5.61 \text{ Nm}^{-2}$$

$$\text{He} = 1960$$

d) $\frac{Q_A}{Q_W} = 2.0$; $\beta_f = 0.33 \text{ g.cm}^{-3}$

$$\mu_a = 0.065 \text{ poise}; \tau_o = 7.15 \text{ Nm}^{-2}$$

$$\text{He} = 1814$$

Q_f	$\Delta P/L$	f	Re
125	38.5	.0360	915
167	44.7	.0234	1223
209	56.4	.0189	1531
250	67.4	.0158	1831
313	82.7	.0123	2292
417	115.	.00964	3058
500	142.5	.00833	3662

Q_f	$\Delta P/L$	f	Re
150	49.5	.0390	850
200	57.2	.0253	113
250	73	.0207	1418
300	86.8	.0171	1701
500	150	.0106	2835

2) 17.3 mm pipe; 0.5% detergent solution.

a) $\frac{Q_A}{Q_W} = 0.5$; $\rho_f = 0.67 \text{ g.cm}^{-3}$
 $\mu_o = 0.052 \text{ poise}$; $\tau_o = 2.34 \text{ Nm}^{-2}$
 He = 4338

Q_f	$\Delta P/L$	f	Re
125	8.3	.0373	1185
205	10.2	.0170	1949
258	11.6	.0122	2447
324	16.4	.0110	3072
400	20.8	.00913	3793
500	25.4	.00713	4741
753	49.0	.00607	7140

b) $\frac{Q_A}{Q_W} = 1.0$; $\rho_f = 0.50 \text{ g.cm}^{-3}$
 $\mu_o = 0.062 \text{ poise}$; $\tau_o = 5.66 \text{ Nm}^{-2}$
 He = 5509

Q_f	$\Delta P/L$	f	Re
167	15.5	.0523	991
274	17.2	.0215	1626
344	18.0	.0143	2042
436	25.6	.0127	2588
535	32.2	.0106	3175
667	38.3	.00810	3959
1004	67.0	.0063	5959

c) $\frac{Q_A}{Q_W} = 1.5$; $\rho_f = 0.4 \text{ g.cm}^{-3}$
 $\mu_o = 0.055 \text{ poise}$; $\tau_o = 6.17 \text{ g.cm}^{-2}$
 He = 6105

Q_f	$\Delta P/L$	f	Re
208	19.6	.0533	1113
342	23.6	.0237	1830
430	25.0	.0159	2302
542	34.5	.0138	2901
668	43.5	.0115	3576
833	51.4	.00871	4459
1255	85.1	.00635	6717

Table G.6: Friction factor v's Reynolds Number for foams

B) Sinter generated foams; $\frac{Q_A}{Q_W} = 1.0$; $\rho_f = 0.5 \text{ g.cm}^3$.

1) 11.4 mm pipe.

$\mu_a = 0.063 \text{ poise}$; $\tau_o = 3.62 \text{ Nm}^{-2}$
He = 5927

Q_f	$\Delta P/L$	f	Re
34	17.6	.186	296
67	22.9	.0596	593
133	35.8	.0237	1179
200	51.8	.0151	1773
250	64.3	.0120	2216
333	87.1	.00919	2952
400	103.9	.00759	3546
533	157.	.00646	4725

2) 17.3 mm pipe.

$\mu_a = 0.058 \text{ poise}$; $\tau_o = 5.07 \text{ Nm}^{-2}$
He = 22,550

Q_f	$\Delta P/L$	f	R
67	13.8	.289	425
204	18.6	.0421	1294
267	21.4	.0283	1694
384	26.7	.0170	2436
684	46.9	.00943	4340
1284	92.5	.00528	8146

3) 26.0 mm pipe.

$\mu_a = 0.108 \text{ poise}$; $\tau_o = 4.74 \text{ Nm}^{-2}$
He = 13,740

Q_f	$\Delta P/L$	f	RE
133	8.7	.355	301
200	9.4	.170	453
333	10.5	.0683	755
384	11.2	.0548	871
534	12.1	.0306	1211
668	13.8	.0223	1514

4) 39.1 mm pipe.

$\mu_a = 0.033 \text{ poise}$; $\tau_o = 5.69 \text{ Nm}^{-2}$
He = 393,400

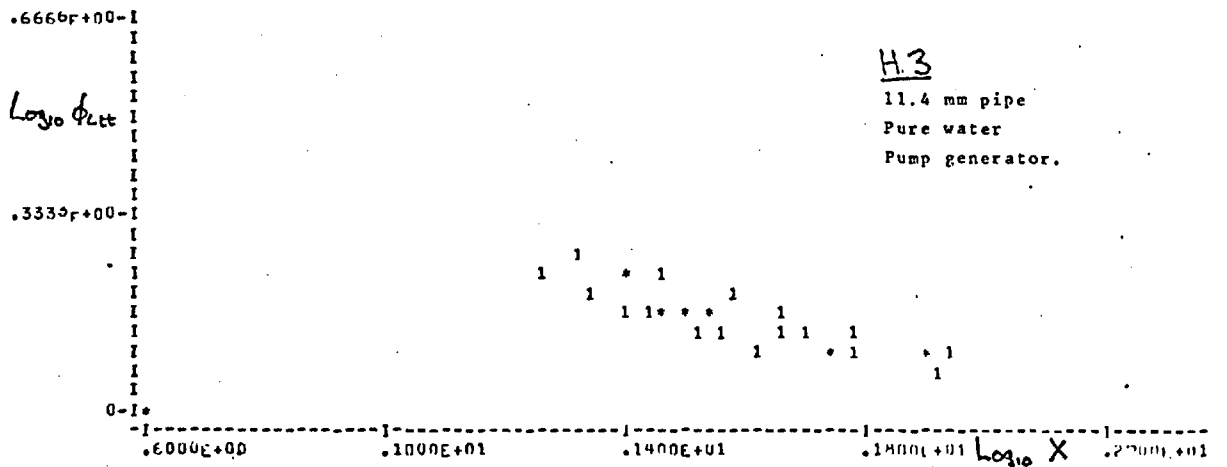
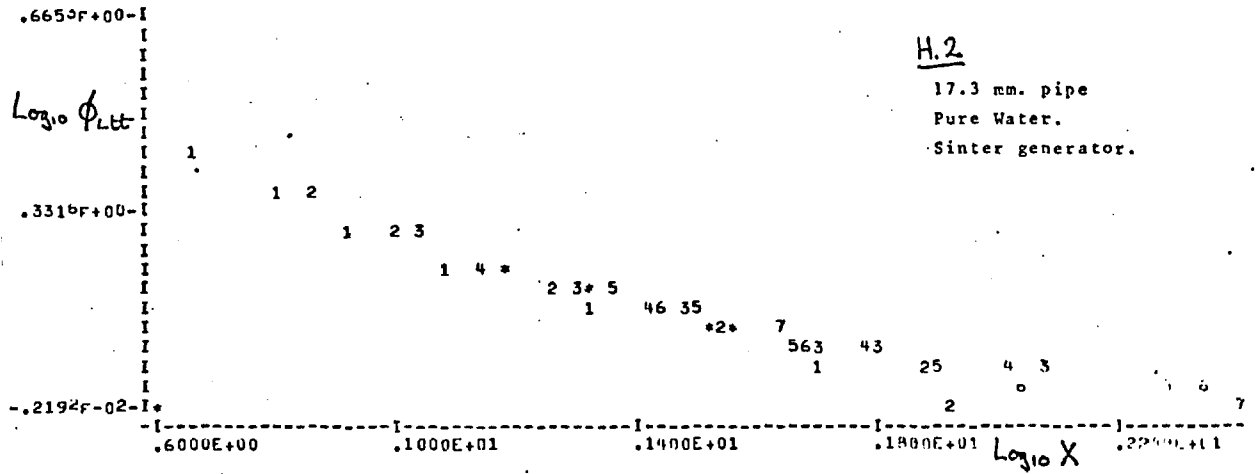
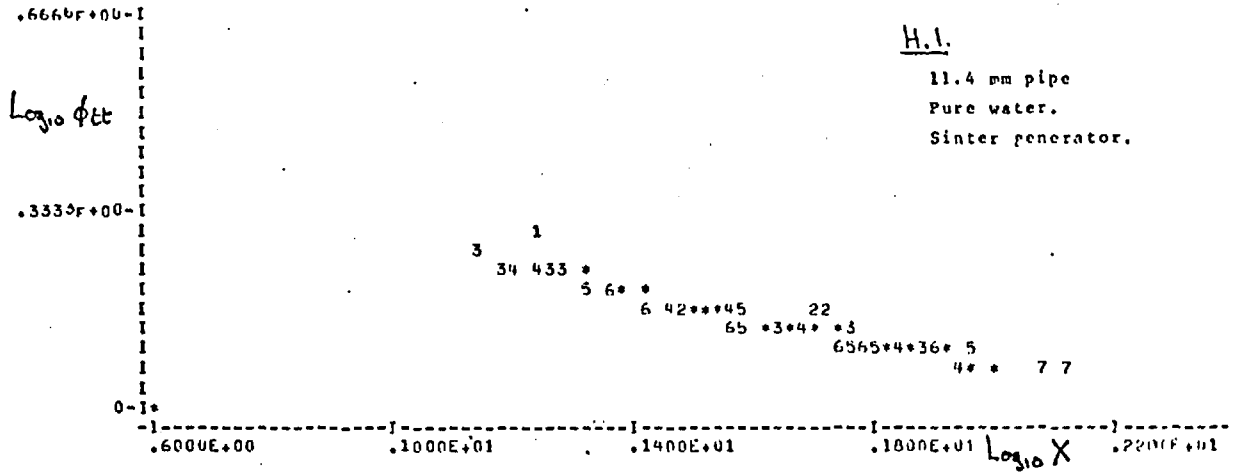
Q_f	$\Delta P/L$	f	Re
200	6.0	.833	987
384	6.15	.232	1895
800	6.35	.0551	3947
1000	6.45	.0358	4934

Table G.7: Results of pressure drop experiments with the venturi nozzle.

Throat diameter = 4.5 mm.

Q_{Ao}	Q_W	P_1	P_2	ΔP_f	ΔP_L	Q_{A2}	Q_{A2}/Q_W	$\Delta P_f/\Delta P_L$	U_2	a	M
L/min	L/min	bar	bar	bar	bar	L/min			m/s	m/s	
3.2	6	1.16	0.95	0.21	0.146	3.35	0.56	1.43	9.8	20.3	0.48
5.7	6	1.25	1.00	0.25		5.7	0.95	1.71	12.3	20.0	0.62
8.7	6	1.34	1.06	0.28		8.2	1.34	1.92	14.9	20.8	0.72
11.8	6	1.41	1.10	0.31		10.75	1.80	2.12	17.5	21.9	0.80
3.2	8	1.27	0.92	0.35	0.259	3.7	0.46	1.35	12.3	20.6	0.60
5.0	8	1.34	0.95	0.39		5.25	0.66	1.51	13.9	19.9	0.70
7.1	8	1.44	1.02	0.42		6.95	0.87	1.62	15.7	20.3	0.77
10.7	8	1.59	1.12	0.47		9.5	1.19	1.81	18.3	21.2	0.86
3.7	10	1.36	0.81	0.55	0.405	4.6	0.46	1.36	15.3	19.3	0.79
5.6	10	1.53	0.90	0.63		6.25	0.63	1.56	17.0	19.5	0.87
8.1	10	1.69	1.05	0.64		7.75	0.78	1.58	18.6	20.6	0.90
10.7	10	1.83	1.16	0.67		9.4	0.94	1.65	20.3	21.6	0.94
12.7	10	1.93	1.24	0.69		10.3	1.03	1.71	21.3	22.3	0.96
2.3	12	1.38	0.68	0.70	0.579	3.4	0.28	1.21	16.1	19.9	0.81
3.9	12	1.58	0.82	0.76		4.8	0.40	1.31	17.6	20.0	0.93
6.8	12	1.84	1.02	0.82		6.7	0.56	1.42	19.6	21.0	0.93
9.4	12	1.98	1.14	0.84		8.3	0.69	1.45	21.3	21.7	0.98
10.2	12	2.15	1.27	0.88		8.1	0.68	1.52	21.1	23.0	0.92

Appendix H; Lockhart-Martinelli Charts for the flow in the
11.4, 17.3, 26.0 and 39.1 mm pipes, at 0.5% and 0.05% Teepol.
Foams generated with the sinter and pump.



.6666F+00-I

$\log_{10} \phi_{LH}$

.3333F+00-I

0-I*

.6000E+00

.1000E+01

.1400E+01

.1800E+01

.2200E+01

$\log_{10} X$

1

1

1

2

1

2 32 34

*

*2

4*4

4

3

4

.6666F+00-I

$\log_{10} \phi_{LH}$

.3333F+00-I

0-I*

.6000E+00

.1000E+01

.1400E+01

.1800E+01

.2200E+01

$\log_{10} X$

1

1

1

2

2

1

23

4*

4

2

3

4

3

4 23

4

3

4

H.5.

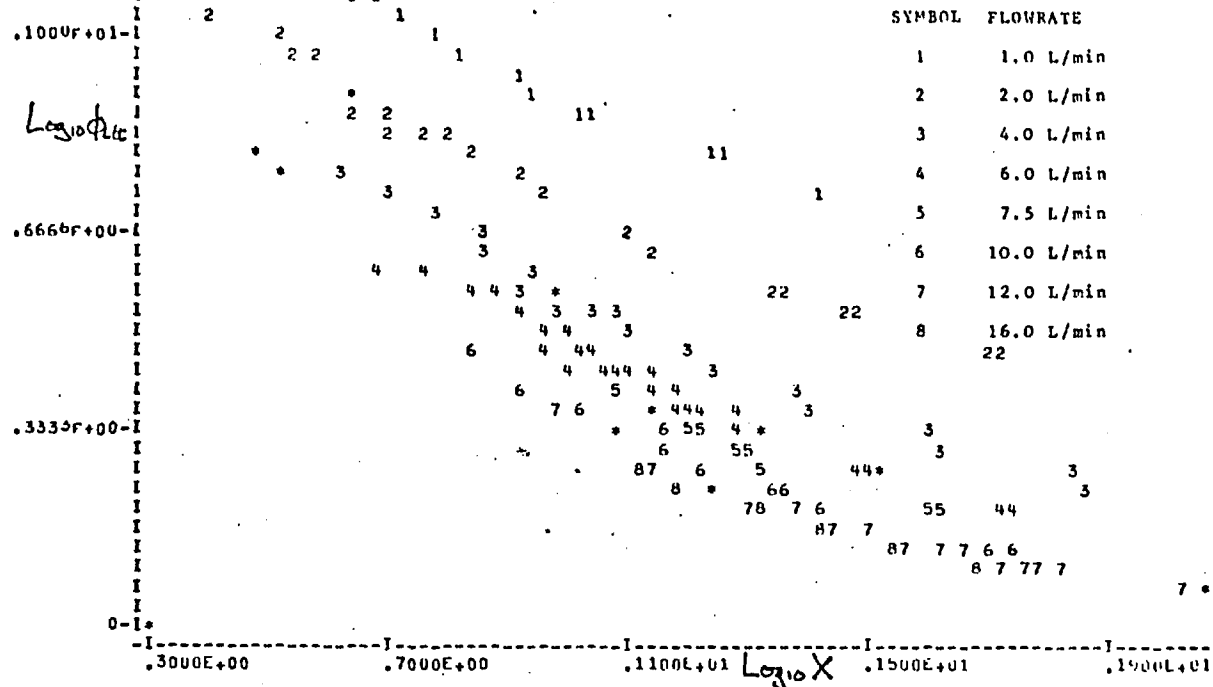
11.4 mm pipe

0.5% Teepol L

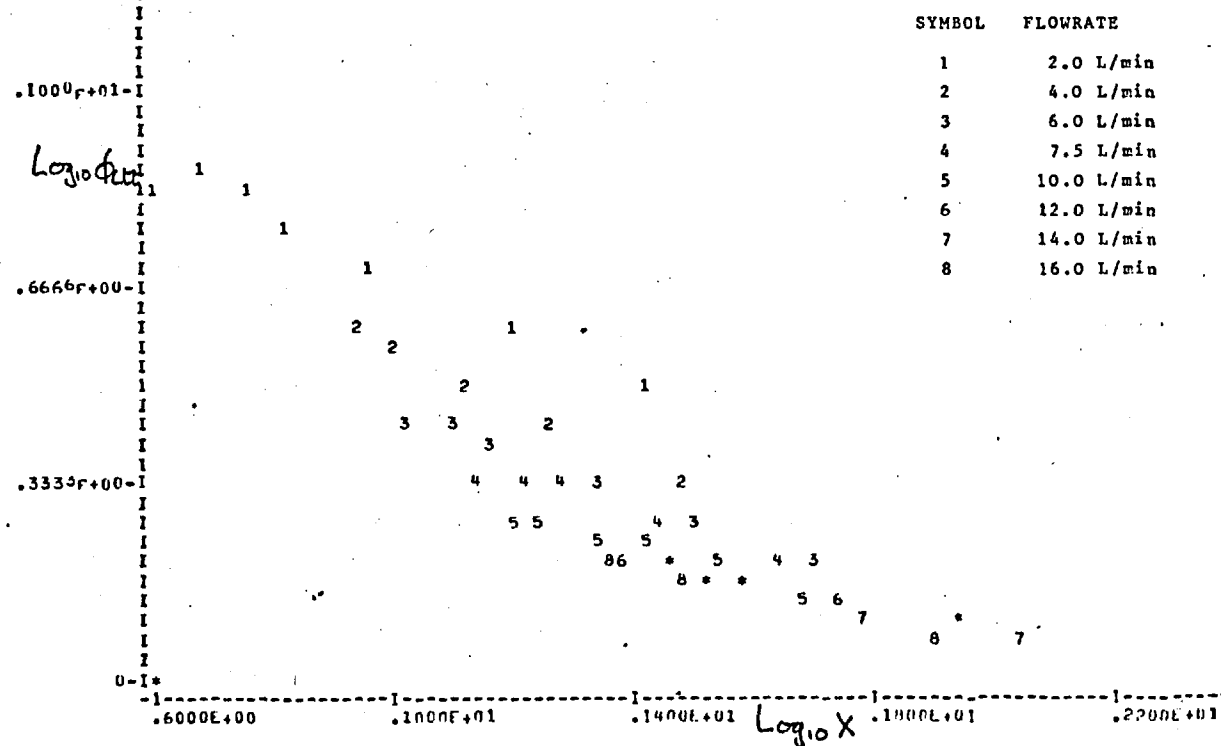
T-piece mixer.

H.611.4 mm pipe
0.5% Teepol

Sinter generator.

H.711.4 mm pipe
0.05% Teepol

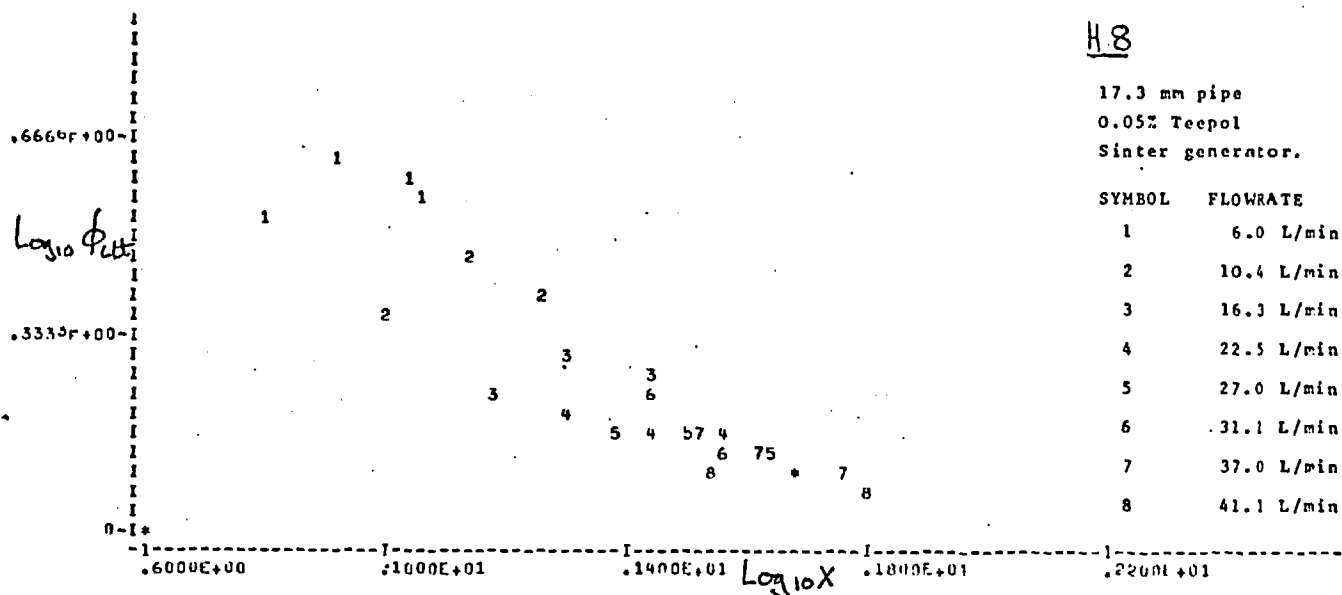
Sinter generator.



H.8

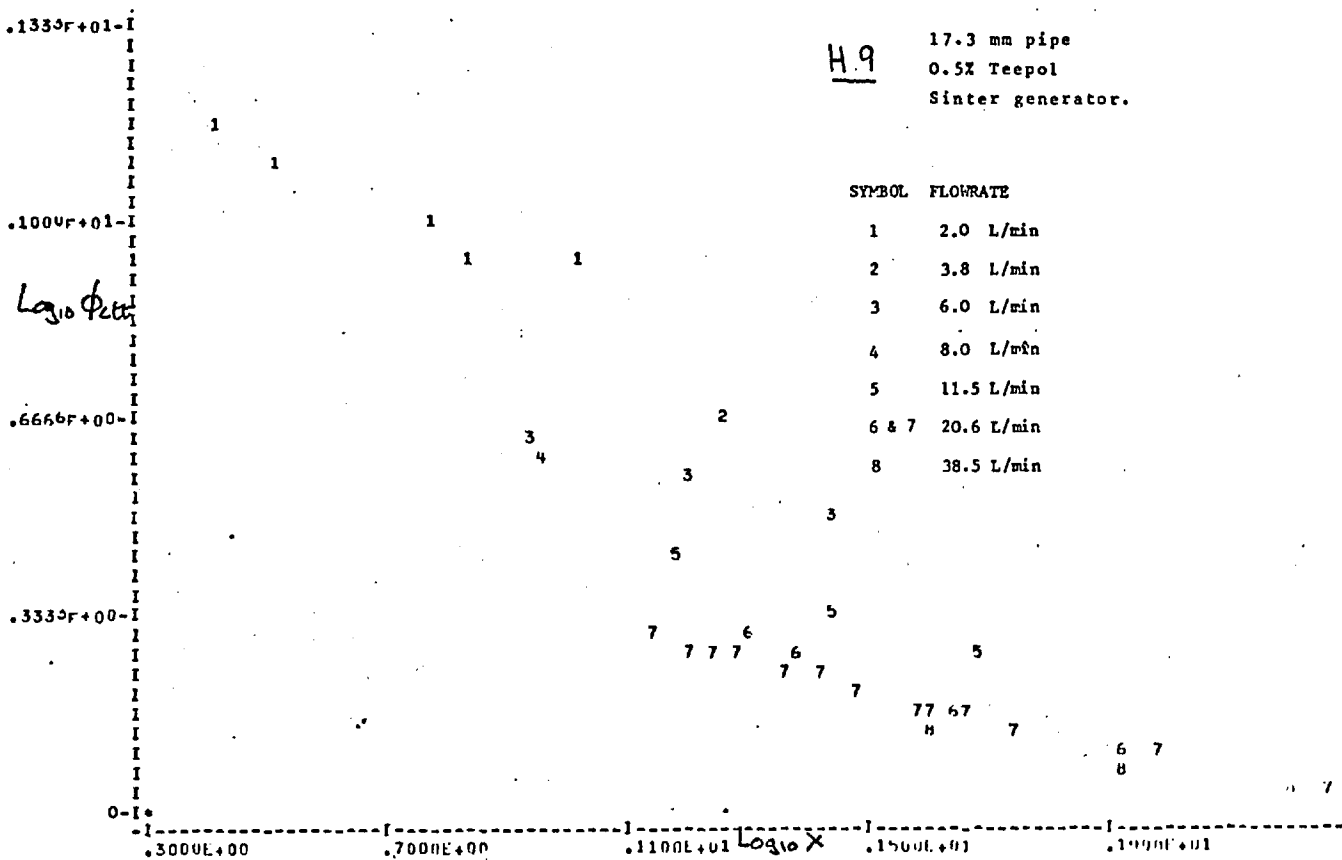
17.3 mm pipe
0.05% Teepol
Sinter generator.

SYMBOL	FLOWRATE
1	6.0 L/min
2	10.4 L/min
3	16.3 L/min
4	22.5 L/min
5	27.0 L/min
6	31.1 L/min
7	37.0 L/min
8	41.1 L/min

H.9

17.3 mm pipe
0.5% Teepol
Sinter generator.

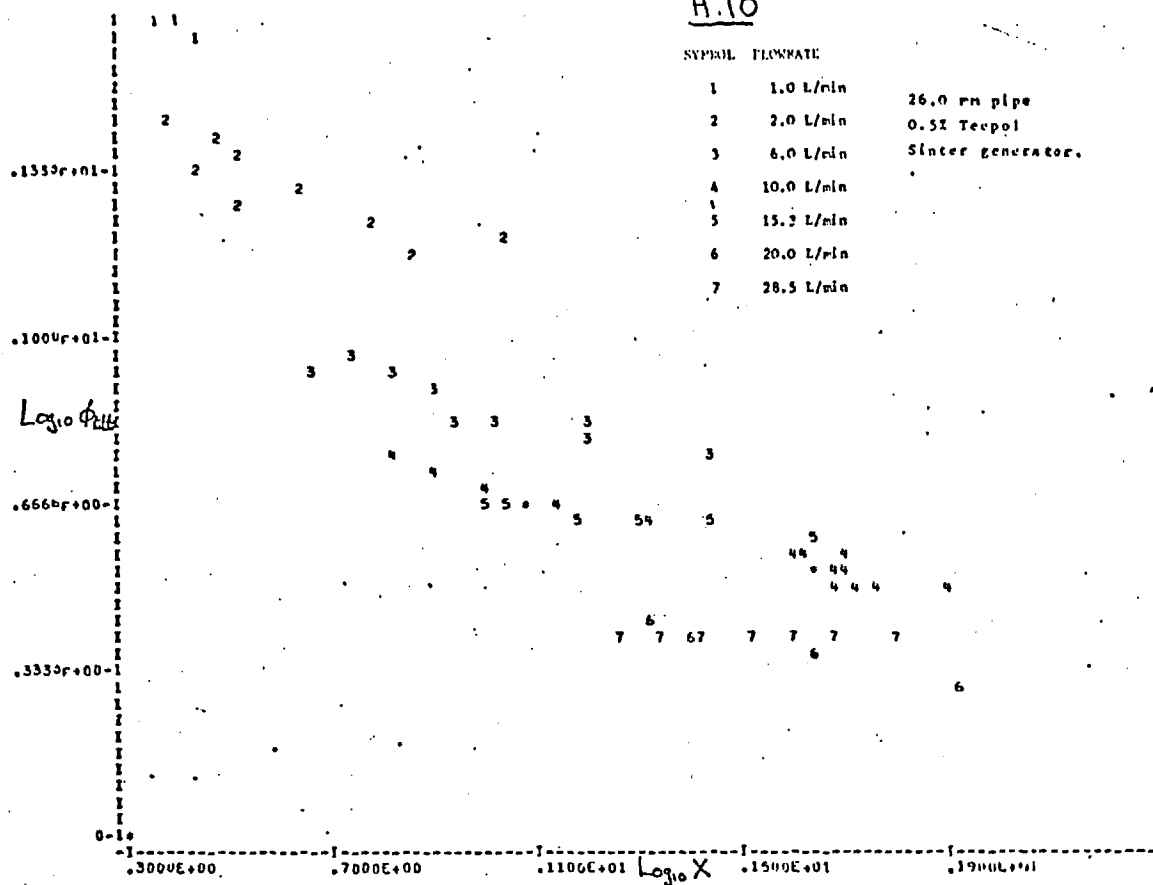
SYMBOL	FLOWRATE
1	2.0 L/min
2	3.8 L/min
3	6.0 L/min
4	8.0 L/min
5	11.5 L/min
6 & 7	20.6 L/min
8	38.5 L/min



H.10

SYMBOL FLOWRATE

1	1.0 L/min	26.0 mm pipe
2	2.0 L/min	0.5% Teepol
3	6.0 L/min	Sinter generator.
4	10.0 L/min	
5	15.0 L/min	
6	20.0 L/min	
7	28.5 L/min	

0 0 0
.1667E+01-1

.1333E+01-1

.1000E+01-

.6666E+00-

.3333E+00-

0-1

.30 .70 1.1 1.5 1.9

Log10 X

H.11

39.1mm pipe
0.5% Teepol
Sinter generator.

SYMBOL FLOWRATE

1	2.0 L/min
2	3.8 L/min
3	6.2 L/min
4	8.0 L/min
5	11.5 L/min
6	20.4 L/min
7	38.0 L/min

Log10 Phi

.30 .70 1.1 1.5 1.9

Log10 X

H.12

.1000F+01-

.6666F+00-

 $\log_{10} \Phi_{11}$

.3333F+00-

0-I*

.6000E+00

.1000E+01

.1400E+01

.1800E+01

.2200E+01

 $\log_{10} X$

11.4 mm pipe

0.5% Teepol

Pump generator

SYMBOL FLOWRATE

1	3.0 L/min
2	4.0 L/min
3	5.0 L/min
4	6.0 L/min
5	7.5 L/min
6	10.0 L/min
7	12.0 L/min
8	16.0 L/min

.1000F+01-

.6666F+00-

 $\log_{10} \Phi_{11}$

.3333F+00-

0-I*

.5892E+00

.9912E+01

.1343E+01

.1795E+01

.2197E+01

 $\log_{10} X$ H.13

17.3 mm pipe

0.5% Teepol

Pump generator

SYMBOL FLOWRATE

1	5.0 L/min
2	8.2 L/min
3	10.2 L/min
4	13.1 L/min
5	16.0 L/min
6	20.0 L/min
7	30.2 L/min

TABLE I.1. PIPE DIA=17.3 MM, DETERGENT= 0 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LE}	SVW	SVA	REGIME
6.1	3.0	1.00	2.7	50.0	1.87	50.90	1.20	.433	.213	EB
6.1	8.4	1.01	4.0	139.1	1.87	20.75	1.46	.433	.592	EB SL
6.1	16.1	1.01	5.8	265.2	1.87	11.77	1.76	.433	1.128	SL
6.1	23.4	1.01	7.6	384.9	1.87	8.49	2.02	.433	1.637	SL
6.1	33.1	1.01	10.2	546.9	1.87	6.26	2.34	.433	2.327	SL
6.2	49.7	1.02	14.9	812.7	1.92	4.47	2.79	.440	3.457	SL
10.4	3.0	1.01	4.7	49.8	4.75	81.35	.99	.737	.212	EB
10.4	3.2	1.01	6.9	52.8	4.75	77.17	1.21	.737	.224	EB
10.4	7.8	1.00	8.9	129.9	4.75	35.22	1.37	.737	.552	EB
10.3	8.4	1.01	8.9	138.2	4.67	32.92	1.38	.730	.588	EB SL
10.3	16.3	1.02	11.8	266.6	4.67	18.49	1.59	.730	1.134	SL
10.3	20.5	1.02	13.1	335.1	4.67	15.13	1.68	.730	1.426	SL
10.3	33.1	1.03	17.2	538.7	4.67	9.97	1.92	.730	2.292	SL
10.3	48.0	1.03	26.9	774.0	4.67	7.24	2.40	.730	3.293	SL
16.3	3.1	1.02	13.0	51.1	10.42	117.61	1.12	1.156	.217	EB
16.3	6.4	1.02	16.2	104.7	10.42	62.60	1.25	1.156	.445	EB
16.3	8.3	1.03	17.6	135.1	10.42	49.98	1.30	1.156	.575	EB
16.3	14.8	1.03	21.4	239.4	10.42	30.22	1.43	1.156	1.019	EB SL
16.3	23.5	1.04	25.4	377.1	10.42	20.25	1.56	1.156	1.604	SL
16.2	33.1	1.05	28.1	527.9	10.31	14.97	1.65	1.149	2.246	SL
16.3	46.9	1.07	40.1	736.4	10.42	11.18	1.96	1.156	3.133	SL
22.6	4.9	1.02	23.8	80.3	18.47	105.17	1.14	1.602	.341	EB
23.4	9.9	1.04	30.4	158.7	19.62	59.23	1.24	1.659	.675	EB SL
22.8	16.1	1.00	34.1	267.8	18.75	37.13	1.35	1.617	1.139	EB SL
22.9	16.3	1.00	34.2	271.1	18.90	36.88	1.35	1.624	1.153	EB SL
22.6	24.2	1.00	40.0	402.3	18.47	25.80	1.47	1.602	1.712	SL
22.9	31.0	1.01	46.3	513.5	18.90	21.05	1.57	1.624	2.185	SL
22.4	31.1	1.06	46.3	489.2	18.18	21.13	1.60	1.588	2.081	SL
22.6	32.8	1.08	44.0	508.1	18.47	20.49	1.54	1.602	2.162	SL
22.6	50.5	1.01	52.9	833.7	18.47	13.60	1.69	1.602	3.547	SL
26.9	3.0	1.04	27.8	48.4	25.05	189.59	1.05	1.907	.206	EB
27.0	8.2	1.06	34.9	129.8	25.21	79.66	1.18	1.914	.552	EB
26.9	15.7	1.07	41.5	245.2	25.05	45.28	1.29	1.907	1.043	EB SL
26.7	23.9	1.08	49.1	370.7	24.72	31.25	1.41	1.893	1.577	EB SL
26.9	34.0	1.10	59.5	515.9	25.05	23.36	1.54	1.907	2.195	SL
31.1	2.9	1.05	34.8	46.3	32.29	222.89	1.04	2.205	.197	EB
31.1	6.6	1.06	39.2	103.3	32.29	109.58	1.10	2.205	.440	EB SL
31.1	17.5	1.09	53.5	266.9	32.29	47.30	1.29	2.205	1.136	SL
31.6	27.2	1.12	64.6	406.4	33.20	32.94	1.39	2.241	1.729	SL
31.1	32.8	1.13	72.4	486.5	32.29	27.68	1.50	2.205	2.070	SL
31.1	46.9	1.15	82.7	681.7	32.29	20.44	1.60	2.205	2.900	SL
37.3	3.0	1.06	45.8	47.2	44.38	255.59	1.02	2.645	.201	DS
36.8	23.8	1.14	75.3	349.2	43.34	42.69	1.32	2.609	1.486	DS SL

TABLE I.2. PIPE DIA=11.4 MM, DETERGENT= 0 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{tt}	SVW	SVA	REGIME
2.0	3.4	.99	7.8	56.9	1.92	17.17	2.01	.327	.557	EB
4.0	2.0	1.04	13.7	31.8	6.47	51.37	1.46	.653	.312	EB
4.0	2.0	.99	13.4	33.3	6.47	50.18	1.44	.653	.327	EB
4.0	3.5	1.05	14.9	55.3	6.47	31.57	1.52	.653	.542	EB
4.0	3.5	.99	14.4	58.8	6.47	30.65	1.49	.653	.576	EB
4.0	5.6	1.06	17.8	87.2	6.47	21.08	1.66	.653	.855	SL EB
4.0	5.6	1.00	17.6	93.0	6.47	20.42	1.65	.653	.911	SL EB
6.0	2.0	1.08	20.8	30.8	13.15	74.47	1.26	.980	.302	EB
6.0	2.0	.99	21.0	33.5	13.15	71.41	1.26	.980	.328	EB
6.0	2.7	1.09	23.6	41.1	13.15	57.60	1.34	.980	.402	EB
6.0	2.7	.99	23.4	45.1	13.15	54.96	1.33	.980	.442	EB
6.0	3.8	1.10	25.7	57.3	13.15	42.89	1.40	.980	.562	EB
6.0	3.8	1.00	25.4	63.0	13.15	40.91	1.39	.980	.617	EB
6.0	5.2	1.12	28.4	77.3	13.15	32.84	1.47	.980	.757	EB SL
6.0	5.2	1.00	28.4	86.2	13.15	31.11	1.47	.980	.844	EB SL
6.0	6.9	1.13	31.7	101.1	13.15	25.82	1.55	.980	.991	SL EB
6.0	6.9	1.00	32.6	114.0	13.15	24.32	1.57	.980	1.117	SL EB
6.0	9.7	1.16	36.3	138.8	13.15	19.39	1.66	.980	1.360	SL
6.0	9.7	1.01	38.4	159.9	13.15	18.07	1.71	.980	1.566	SL
6.0	13.2	1.18	40.5	184.9	13.15	14.97	1.76	.980	1.811	SL
6.0	13.2	1.02	43.8	214.0	13.15	13.92	1.83	.980	2.097	SL
7.5	2.1	1.10	26.7	31.6	19.43	87.75	1.17	1.225	.309	EB
7.5	2.1	1.00	28.8	34.8	19.43	83.57	1.22	1.225	.341	EB
7.5	2.6	1.11	30.6	38.8	19.43	73.09	1.25	1.225	.380	EB
7.5	2.6	1.00	31.0	43.0	19.43	69.38	1.26	1.225	.422	EB
7.5	4.3	1.14	35.9	62.6	19.43	47.62	1.36	1.225	.614	EB
7.5	4.3	1.00	36.0	71.1	19.43	44.69	1.36	1.225	.697	EB
7.5	5.9	1.16	39.9	84.2	19.43	36.48	1.43	1.225	.825	EB SL
7.5	5.9	1.01	41.0	97.0	19.43	33.98	1.45	1.225	.951	EB SL
7.5	8.0	1.19	43.9	111.6	19.43	28.26	1.50	1.225	1.094	SL
7.5	8.0	1.02	45.4	129.6	19.43	26.23	1.53	1.225	1.270	SL
7.5	14.5	1.26	55.1	190.4	19.43	17.31	1.68	1.225	1.865	SL
7.5	14.5	1.05	61.0	228.4	19.43	15.81	1.77	1.225	2.238	SL
10.0	2.8	1.21	47.2	38.2	32.14	92.13	1.21	1.633	.374	EB
10.0	2.8	1.02	47.2	45.3	32.14	84.60	1.21	1.633	.444	EB
10.0	4.3	1.22	53.2	58.2	32.14	63.53	1.29	1.633	.570	EB
10.0	4.3	1.05	53.4	67.8	32.14	58.87	1.29	1.633	.664	EB
10.0	5.1	1.26	55.8	67.0	32.14	55.54	1.32	1.633	.656	EB
10.0	5.1	1.03	56.4	81.6	32.14	50.35	1.32	1.633	.799	EB
10.0	7.1	1.04	63.8	112.9	32.14	37.80	1.41	1.633	1.106	EB SL
10.0	7.1	1.27	57.8	92.9	32.14	41.66	1.34	1.633	.911	EB SL
10.0	7.8	1.29	64.2	99.8	32.14	38.80	1.41	1.633	.978	SL EB
10.0	7.8	1.03	68.8	124.7	32.14	34.73	1.46	1.633	1.221	SL EB
10.0	9.0	1.32	66.3	112.4	32.14	34.66	1.44	1.633	1.101	SL
10.0	9.0	1.09	71.6	136.3	32.14	31.48	1.49	1.633	1.335	SL
10.0	13.9	1.06	85.0	217.8	32.14	21.15	1.63	1.633	2.134	SL
10.0	13.9	1.36	75.8	169.7	32.14	23.96	1.54	1.633	1.663	SL
10.0	14.2	1.37	75.1	170.8	32.14	23.69	1.53	1.633	1.674	SL
10.0	14.2	1.05	89.2	223.5	32.14	20.72	1.67	1.633	2.190	SL
12.0	3.2	1.29	62.6	41.0	44.22	99.18	1.19	1.959	.402	EB
12.0	3.2	1.07	62.6	49.6	44.22	90.24	1.19	1.959	.485	EB
12.0	4.2	1.31	66.6	52.9	44.22	78.89	1.23	1.959	.518	EB
12.0	4.2	1.07	67.0	64.9	44.22	71.18	1.23	1.959	.636	EB
12.0	5.1	1.33	70.1	63.4	44.22	66.99	1.26	1.959	.621	EB SL
12.0	5.1	1.08	71.2	78.0	44.22	60.39	1.27	1.959	.764	EB
12.0	6.3	1.36	74.2	76.9	44.22	56.19	1.30	1.959	.753	EB SL
12.0	6.3	1.09	77.4	95.4	44.22	50.46	1.32	1.959	.934	EB SL

TABLE I.2. PIPE DIA=11.4 MM, DETERGENT= 0 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	Φ_{LCC}	SVW	SVA	REGIME
12.0	10.6	1.44	86.6	121.8	44.22	36.72	1.40	1.959	1.193	SL
12.0	10.6	1.15	95.0	152.2	44.22	32.86	1.47	1.959	1.491	SL
12.0	16.4	1.54	100.0	176.6	44.22	25.89	1.50	1.959	1.731	SL
12.0	16.4	1.19	116.0	227.9	44.22	22.80	1.62	1.959	2.233	SL
14.0	2.7	1.35	74.8	33.2	57.92	134.49	1.14	2.286	.325	EB
14.0	2.7	1.11	74.8	40.3	57.92	122.02	1.14	2.286	.395	EB
14.0	3.8	1.39	79.9	45.4	57.92	101.16	1.17	2.286	.445	EB
14.0	3.8	1.12	81.0	56.2	57.92	90.92	1.18	2.286	.550	EB
14.0	5.6	1.43	87.2	64.9	57.92	73.15	1.23	2.286	.636	EB SL
14.0	5.6	1.13	90.6	81.7	57.92	65.21	1.25	2.286	.800	SL
14.0	9.7	1.52	102.0	105.4	57.92	46.69	1.33	2.286	1.033	SL
14.0	9.7	1.17	110.2	137.4	57.92	40.92	1.38	2.286	1.346	SL

TABLE I.3. PIPE DIA=11.4 MM, DETERGENT= 0 GENERATOR, PUMP

QW	QA	PMID	DPTOT	QA	DPW	X	Φ_{LCC}	SVW	SVA	REGIME
10.0	3.0	1.21	52.0	41.2	32.14	86.39	1.27	1.633	.404	DS
10.0	3.0	1.04	52.8	48.2	32.14	79.93	1.28	1.633	.472	DS
10.0	3.1	1.19	46.9	43.2	32.14	83.34	1.21	1.633	.424	DS EB
10.0	3.1	1.04	47.4	49.6	32.14	77.86	1.21	1.633	.485	EB
10.0	4.6	1.22	54.4	62.6	32.14	59.73	1.30	1.633	.613	DS EB
10.0	4.6	1.05	51.8	73.0	32.14	55.33	1.27	1.633	.715	EB
10.0	4.6	1.23	56.4	62.3	32.14	59.89	1.32	1.633	.610	EB
10.0	4.6	1.05	52.0	73.0	32.14	55.31	1.27	1.633	.716	DS EB
10.0	5.8	1.26	63.5	76.9	32.14	49.41	1.41	1.633	.753	DS
10.0	5.8	1.05	64.8	92.1	32.14	45.15	1.42	1.633	.902	DS
10.0	6.3	1.24	59.1	84.7	32.14	45.65	1.36	1.633	.829	EB
10.0	6.3	1.05	54.6	99.5	32.14	42.12	1.30	1.633	.974	EB SL
10.0	8.0	1.31	75.2	101.9	32.14	38.05	1.53	1.633	.998	DS
10.0	8.0	1.06	74.4	125.8	32.14	34.25	1.52	1.633	1.232	SL
10.0	8.3	1.26	62.1	109.6	32.14	36.17	1.39	1.633	1.074	EB SL
10.0	8.3	1.06	61.4	130.1	32.14	33.21	1.38	1.633	1.274	SL
10.0	8.8	1.27	65.5	114.9	32.14	34.57	1.43	1.633	1.125	SL
10.0	8.8	1.06	64.8	137.3	32.14	31.63	1.42	1.633	1.345	SL
10.0	9.7	1.28	65.1	126.3	32.14	31.79	1.42	1.633	1.237	SL
10.0	9.7	1.06	66.4	151.5	32.14	29.03	1.44	1.633	1.484	SL
10.0	10.0	1.29	67.4	129.1	32.14	31.09	1.45	1.633	1.265	SL
10.0	10.0	1.07	68.6	155.2	32.14	28.36	1.46	1.633	1.520	SL
10.0	11.2	1.37	90.6	136.7	32.14	28.95	1.68	1.633	1.339	SL DS
10.0	11.2	1.07	87.2	174.0	32.14	25.66	1.65	1.633	1.705	SL
10.0	11.6	1.30	70.2	147.9	32.14	27.47	1.48	1.633	1.449	SL
10.0	11.6	1.08	71.0	179.3	32.14	24.95	1.49	1.633	1.757	SL
10.0	13.4	1.37	99.6	162.7	32.14	24.81	1.76	1.633	1.594	SL EB
10.0	13.4	1.08	85.8	206.2	32.14	22.04	1.63	1.633	2.020	SL
10.0	16.8	1.40	105.5	199.5	32.14	20.58	1.81	1.633	1.955	SL
10.0	16.8	1.09	93.8	257.2	32.14	18.13	1.71	1.633	2.520	SL

TABLE I.4. PIPE DIA=11.4 MM, DETERGENT= .50 GENERATOR, T-PIECE

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{Lt}	SVW	SVA	REGIME
2.0	2.4	1.02	5.8	39.0	1.92	23.61	1.74	.327	.383	EB
2.0	3.7	1.02	7.2	60.1	1.92	16.18	1.94	.327	.589	EB
2.0	4.9	1.02	7.8	79.6	1.92	12.65	2.01	.327	.780	EB SL
2.0	6.9	1.02	7.6	112.3	1.92	9.37	1.99	.327	1.100	SL
6.0	2.4	1.03	22.0	38.6	13.15	62.10	1.29	.980	.378	EB
6.0	3.8	1.03	26.4	60.9	13.15	41.60	1.42	.980	.597	EB
6.0	5.3	1.04	29.0	84.7	13.15	31.14	1.49	.980	.830	EB SL
6.0	6.9	1.04	32.2	109.7	13.15	24.79	1.56	.980	1.075	SL
6.0	8.8	1.05	35.2	139.7	13.15	20.06	1.64	.980	1.368	SL
6.0	11.8	1.05	39.0	187.3	13.15	15.52	1.72	.980	1.835	SL
10.0	2.5	1.04	46.8	39.8	32.14	94.18	1.21	1.633	.390	EB
10.0	3.7	1.05	53.0	58.5	32.14	67.06	1.28	1.633	.573	EB
10.0	5.2	1.05	59.0	81.9	32.14	49.87	1.35	1.633	.803	SL EB
10.0	6.9	1.06	63.6	108.3	32.14	39.02	1.41	1.633	1.061	SL
10.0	8.7	1.06	69.6	136.1	32.14	31.90	1.47	1.633	1.333	SL
10.0	10.6	1.06	75.2	165.4	32.14	26.87	1.53	1.633	1.620	SL
14.0	2.9	1.05	78.8	45.7	57.92	111.58	1.17	2.286	.448	EB
14.0	4.4	1.06	88.2	68.8	57.92	77.75	1.23	2.286	.674	EB SL
14.0	6.1	1.07	96.6	94.9	57.92	58.58	1.29	2.286	.930	EB SL
14.0	8.3	1.07	106.8	128.6	57.92	44.84	1.36	2.286	1.260	SL
14.0	10.6	1.08	117.4	163.1	57.92	36.32	1.42	2.286	1.598	SL
14.0	12.8	1.08	124.6	196.1	57.92	30.86	1.47	2.286	1.921	SL

TABLE I.5. PIPE DIA=11.4 MM, DETERGENT= .05 GENERATOR, T-PIECE

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{Lt}	SVW	SVA	REGIME
2.0	2.0	1.02	5.0	32.5	1.92	27.72	1.61	.327	.318	EB
2.0	3.2	1.02	5.8	52.0	1.92	18.37	1.74	.327	.509	EB
2.0	4.4	1.02	6.2	71.5	1.92	13.91	1.80	.327	.700	EB SL
2.0	6.5	1.02	7.4	105.5	1.92	9.88	1.96	.327	1.034	SL
6.0	2.3	1.03	22.2	37.1	13.15	64.37	1.30	.980	.363	EB
6.0	3.2	1.03	24.0	51.6	13.15	48.21	1.35	.980	.506	EB
6.0	4.1	1.03	26.0	66.1	13.15	38.82	1.41	.980	.648	EB
6.0	5.4	1.03	28.8	87.0	13.15	30.53	1.48	.980	.852	EB SL
6.0	6.4	1.03	30.0	102.9	13.15	26.33	1.51	.980	1.008	SL
6.0	8.1	1.03	33.0	129.9	13.15	21.45	1.58	.980	1.273	SL
10.0	2.6	1.04	47.4	41.5	32.14	90.93	1.21	1.633	.406	EB
10.0	4.0	1.04	53.6	63.7	32.14	62.42	1.29	1.633	.624	EB
10.0	5.6	1.04	59.4	88.9	32.14	46.56	1.36	1.633	.871	EB SL
10.0	7.0	1.04	63.4	111.0	32.14	38.33	1.40	1.633	1.087	SL
10.0	8.3	1.05	67.6	131.2	32.14	33.07	1.45	1.633	1.285	SL
10.0	9.9	1.05	73.0	156.0	32.14	28.39	1.51	1.633	1.529	SL
14.0	2.5	1.05	76.0	39.5	57.92	126.93	1.15	2.286	.387	EB
14.0	4.5	1.05	89.2	70.7	57.92	76.05	1.24	2.286	.693	EB
14.0	5.2	1.06	93.6	81.5	57.92	67.11	1.27	2.286	.799	EB
14.0	6.0	1.06	96.6	93.9	57.92	59.25	1.29	2.286	.920	EB SL
14.0	8.3	1.07	110.0	129.0	57.92	44.76	1.38	2.286	1.264	SL
14.0	11.4	1.07	122.8	176.4	57.92	33.98	1.46	2.286	1.729	SL

TABLE I.6. PIPE DIA=11.4 MM, DETERGENT= .05 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{Ltt}	SVW	SVA	REGIME
2.0	2.2	1.05	20.2	35.1	1.92	25.74	3.24	.327	.343	ST
2.0	3.8	1.05	28.4	60.1	1.92	16.02	3.84	.327	.589	ST FM
2.0	7.2	1.07	46.4	112.6	1.92	9.21	4.91	.327	1.103	FM
2.0	10.6	1.08	69.8	163.5	1.92	6.61	6.03	.327	1.602	FM
2.0	12.7	1.09	87.4	193.5	1.92	5.68	6.74	.327	1.896	SL FM
2.0	15.4	1.11	108.2	231.9	1.92	4.82	7.50	.327	2.272	SL
2.0	18.7	1.09	94.0	287.0	1.92	4.03	6.99	.327	2.812	SL
4.0	3.7	1.05	31.4	58.5	6.47	30.09	2.20	.653	.573	ST
4.0	6.5	1.06	44.4	101.7	6.47	18.47	2.62	.653	.996	FM
4.0	9.4	1.07	60.0	145.6	6.47	13.44	3.05	.653	1.427	FM
4.0	13.0	1.09	83.0	198.8	6.47	10.18	3.58	.653	1.948	FM
4.0	15.9	1.10	100.6	240.6	6.47	8.58	3.94	.653	2.357	SL
6.0	3.1	1.05	33.8	49.2	13.15	49.98	1.60	.980	.482	ST
6.0	5.3	1.06	45.6	83.2	13.15	31.42	1.86	.980	.816	ST
6.0	8.2	1.07	60.0	127.8	13.15	21.52	2.14	.980	1.252	DS
6.0	13.1	1.09	87.0	200.8	13.15	14.41	2.57	.980	1.968	DS
6.0	15.2	1.09	98.2	232.0	13.15	12.68	2.73	.980	2.273	SL
6.0	18.7	1.09	97.6	287.1	13.15	10.54	2.72	.980	2.813	SL
7.5	4.5	1.06	46.0	70.8	19.43	44.03	1.54	1.225	.694	ST
7.5	7.6	1.07	63.2	118.4	19.43	27.98	1.80	1.225	1.160	DS
7.5	12.2	1.08	88.4	187.1	19.43	18.63	2.13	1.225	1.834	DS
7.5	14.2	1.09	96.4	217.2	19.43	16.34	2.23	1.225	2.128	SL DS
7.5	17.1	1.07	85.4	265.6	19.43	13.78	2.10	1.225	2.602	SL
10.0	2.7	1.05	48.8	42.8	32.14	88.20	1.23	1.633	.420	ST
10.0	5.4	1.06	62.8	84.7	32.14	48.36	1.40	1.633	.830	ST
10.0	8.0	1.07	78.2	124.4	32.14	34.44	1.56	1.633	1.218	DS ST
10.0	10.9	1.08	93.6	168.0	32.14	26.38	1.71	1.633	1.646	DS
10.0	13.3	1.08	101.2	204.5	32.14	22.20	1.77	1.633	2.003	SL
10.0	17.1	1.08	102.2	264.7	32.14	17.75	1.78	1.633	2.593	SL
10.0	19.6	1.08	106.6	302.2	32.14	15.79	1.82	1.633	2.961	SL
12.0	3.3	1.06	65.0	51.9	44.22	87.15	1.21	1.959	.509	ST
12.0	5.6	1.07	79.4	87.2	44.22	55.13	1.34	1.959	.855	ST DS
12.0	9.9	1.08	101.6	152.3	44.22	33.70	1.52	1.959	1.492	DS SL
12.0	11.9	1.08	106.0	183.4	44.22	28.66	1.55	1.959	1.796	SL
12.0	14.6	1.08	114.6	224.6	44.22	23.98	1.61	1.959	2.201	SL
14.0	2.9	1.06	77.0	45.5	57.92	111.81	1.15	2.286	.446	ST
14.0	6.0	1.07	97.0	93.1	57.92	59.52	1.29	2.286	.912	DS ST
14.0	10.2	1.08	118.0	156.7	57.92	37.59	1.43	2.286	1.535	DS
14.0	11.8	1.08	124.6	181.3	57.92	33.09	1.47	2.286	1.776	SL
14.0	13.8	1.09	135.2	211.2	57.92	28.91	1.53	2.286	2.069	SL
16.0	4.8	1.07	106.8	74.8	73.17	81.14	1.21	2.613	.733	DS
16.0	11.4	1.10	146.8	173.2	73.17	38.54	1.42	2.613	1.697	DS
16.0	13.7	1.10	154.6	207.9	73.17	32.83	1.45	2.613	2.037	DS SL
16.0	15.5	1.11	166.4	233.6	73.17	29.57	1.51	2.613	2.289	SL
16.0	20.6	1.12	192.4	308.0	73.17	23.14	1.62	2.613	3.018	SL

TABLE I.7. PIPE DIA=11.4 MM, DETERGENT= .50 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LH}	SVW	SVA	REGIME
1.0	1.1	1.08	16.8	16.9	.57	26.23	5.42	.163	.165	ST
1.0	1.7	1.11	21.5	25.6	.57	18.11	6.13	.163	.250	ST
1.0	1.7	1.03	22.6	27.3	.57	17.53	6.29	.163	.267	ST
1.0	3.1	1.15	30.0	44.9	.57	10.91	7.24	.163	.440	ST DS
1.0	3.1	1.05	32.6	49.2	.57	10.42	7.55	.163	.482	ST DS
1.0	4.1	1.19	38.4	57.1	.57	8.71	8.20	.163	.560	DS
1.0	4.1	1.06	43.8	64.1	.57	8.22	8.75	.163	.628	DS
1.0	5.8	1.09	59.8	88.6	.57	6.14	10.23	.163	.868	DS
1.0	5.8	1.26	51.2	76.4	.57	6.61	9.46	.163	.749	DS
1.0	7.7	1.34	66.0	95.8	.57	5.31	10.75	.163	.938	DS
1.0	7.7	1.11	78.0	114.9	.57	4.85	11.68	.163	1.126	DS
1.0	9.6	1.39	75.3	114.7	.57	4.46	11.48	.163	1.124	DS
1.0	9.6	1.13	87.4	141.0	.57	4.03	12.37	.163	1.382	DS
1.0	14.1	1.12	127.4	209.4	.57	2.86	14.93	.163	2.052	DS
1.0	15.7	1.13	140.6	230.3	.57	2.62	15.68	.163	2.256	DS
1.0	17.5	1.14	147.0	255.0	.57	2.39	16.04	.163	2.498	SL DS
1.0	19.5	1.15	154.2	281.9	.57	2.18	16.42	.163	2.762	SL
2.0	1.0	1.02	17.4	16.2	1.92	50.85	3.01	.327	.159	ST
2.0	1.0	1.08	16.5	15.4	1.92	52.16	2.93	.327	.151	ST
2.0	1.9	1.10	22.2	28.6	1.92	30.12	3.40	.327	.280	ST
2.0	1.9	1.03	23.0	30.6	1.92	29.11	3.46	.327	.300	ST
2.0	2.6	1.12	25.4	38.6	1.92	23.05	3.63	.327	.378	ST
2.0	2.6	1.04	26.4	41.6	1.92	22.20	3.71	.327	.408	ST
2.0	4.9	1.17	35.7	69.4	1.92	13.55	4.31	.327	.680	ST
2.0	4.9	1.05	38.4	77.2	1.92	12.85	4.47	.327	.756	ST
2.0	8.0	1.27	53.0	104.5	1.92	9.19	5.25	.327	1.023	DS
2.0	8.0	1.09	62.6	121.9	1.92	8.51	5.71	.327	1.194	DS
2.0	11.5	1.37	71.1	139.1	1.92	6.95	6.08	.327	1.363	DS
2.0	11.5	1.13	86.0	169.6	1.92	6.30	6.69	.327	1.661	DS
2.0	14.9	1.46	84.5	169.6	1.92	5.71	6.63	.327	1.661	DS
2.0	14.9	1.16	107.8	213.4	1.92	5.09	7.49	.327	2.091	DS
2.0	17.0	1.11	128.8	255.0	1.92	4.43	8.18	.327	2.498	DS SL
2.0	17.8	1.54	95.5	191.7	1.92	5.03	7.05	.327	1.878	DS
2.0	17.8	1.19	130.0	248.1	1.92	4.42	8.22	.327	2.431	DS
2.0	21.8	1.66	110.6	218.4	1.92	4.36	7.58	.327	2.140	DS
2.0	21.8	1.24	158.8	292.0	1.92	3.77	9.09	.327	2.861	DS
2.0	22.4	1.15	171.8	323.6	1.92	3.55	9.45	.327	3.170	SL DS
2.0	25.0	1.17	182.6	355.6	1.92	3.25	9.75	.327	3.484	SL DS
2.0	33.5	1.20	208.4	465.7	1.92	2.54	10.41	.327	4.562	SL
4.0	1.4	1.09	20.2	21.3	6.47	71.76	1.77	.653	.209	ST
4.0	1.4	1.03	20.8	22.7	6.47	69.55	1.79	.653	.222	ST
4.0	2.7	1.13	27.0	39.8	6.47	41.05	2.04	.653	.390	ST
4.0	2.7	1.04	28.4	43.2	6.47	39.40	2.10	.653	.423	ST
4.0	4.8	1.17	36.7	67.9	6.47	25.32	2.38	.653	.665	ST
4.0	4.8	1.06	39.0	75.5	6.47	24.01	2.46	.653	.740	ST
4.0	7.8	1.24	48.4	104.8	6.47	16.99	2.74	.653	1.026	DS ST
4.0	7.8	1.08	54.0	120.2	6.47	15.86	2.89	.653	1.178	DS ST
4.0	11.1	1.32	61.2	140.2	6.47	12.87	3.08	.653	1.373	DS
4.0	11.1	1.10	72.2	166.9	6.47	11.79	3.34	.653	1.636	DS
4.0	14.0	1.38	71.2	168.4	6.47	10.76	3.32	.653	1.650	DS
4.0	14.0	1.13	88.6	205.6	6.47	9.74	3.70	.653	2.015	DS
4.0	14.2	1.09	91.2	217.2	6.47	9.43	3.76	.653	2.128	DS
4.0	16.4	1.44	81.1	189.3	6.47	9.56	3.54	.653	1.854	DS
4.0	16.4	1.15	104.4	237.5	6.47	8.54	4.02	.653	2.327	DS
4.0	20.2	1.53	93.5	218.9	6.47	8.22	3.80	.653	2.145	DS
4.0	20.2	1.19	128.8	282.7	6.47	7.24	4.46	.653	2.770	DS SL
4.0	19.6	1.11	129.4	293.9	6.47	7.18	4.47	.653	2.880	DS

TABLE 1.7. PIPE DIA=11.4 MM, DETERGENT= .50 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LH}	SVW	SWA	REGIME
4.0	24.4	1.13	162.2	357.7	6.47	5.99	5.01	.653	3.504	SL DS
4.0	29.8	1.17	193.4	425.5	6.47	5.10	5.47	.653	4.169	SL
4.0	37.4	1.17	212.4	530.6	6.47	4.19	5.73	.653	5.198	SL
4.0	47.4	1.16	219.2	679.7	6.47	3.39	5.82	.653	6.659	SL
4.0	47.9	1.16	219.0	688.6	6.47	3.35	5.82	.653	6.746	SL
4.0	54.2	1.17	246.6	772.3	6.47	3.02	6.18	.653	7.566	SL
4.0	54.5	1.17	245.8	772.8	6.47	3.02	6.17	.653	7.572	SL
6.0	2.9	1.16	33.5	41.4	13.15	55.87	1.60	.980	.405	ST
6.0	2.9	1.05	35.4	45.8	13.15	53.11	1.64	.980	.449	ST
6.0	5.3	1.21	42.9	72.2	13.15	33.74	1.81	.930	.707	ST
6.0	5.3	1.07	45.8	81.6	13.15	31.73	1.87	.980	.800	ST
6.0	5.5	1.07	47.4	85.2	13.15	30.62	1.90	.980	.835	ST
6.0	5.5	1.21	44.1	75.0	13.15	32.64	1.83	.980	.735	ST
6.0	9.5	1.29	58.1	121.8	13.15	20.86	2.10	.980	1.193	DS
6.0	9.5	1.10	64.6	143.0	13.15	19.26	2.22	.980	1.401	DS
6.0	9.8	1.31	60.0	123.9	13.15	20.45	2.14	.980	1.214	ST DS
6.0	9.8	1.11	66.2	146.3	13.15	18.82	2.24	.980	1.434	ST DS
6.0	11.5	1.06	75.0	178.9	13.15	16.03	2.39	.980	1.753	DS
6.0	12.7	1.35	68.8	155.2	13.15	16.58	2.29	.980	1.520	DS
6.0	12.7	1.12	79.4	187.8	13.15	15.07	2.46	.980	1.840	DS
6.0	14.4	1.14	90.0	208.2	13.15	13.65	2.62	.980	2.040	DS
6.0	14.4	1.41	75.1	169.4	13.15	15.14	2.39	.980	1.659	DS
6.0	15.8	1.42	79.4	184.2	13.15	14.02	2.46	.980	1.805	DS
6.0	15.8	1.14	96.2	228.7	13.15	12.58	2.70	.980	2.241	DS
6.0	18.0	1.08	115.0	274.9	13.15	10.93	2.96	.980	2.693	DS
6.0	19.8	1.52	91.9	215.9	13.15	11.90	2.64	.980	2.115	DS
6.0	19.8	1.18	119.2	276.4	13.15	10.52	3.01	.980	2.708	DS
6.0	21.0	1.55	96.6	224.3	13.15	11.42	2.71	.980	2.198	DS
6.0	21.0	1.20	127.2	289.7	13.15	10.05	3.11	.980	2.839	DS
6.0	24.2	1.58	103.8	252.8	13.15	10.20	2.81	.980	2.477	DS
6.0	24.2	1.21	138.0	330.6	13.15	8.92	3.24	.980	3.239	DS
6.0	25.2	1.11	159.0	375.8	13.15	8.24	3.48	.980	3.682	DS SL
6.0	29.7	1.75	121.3	281.4	13.15	8.95	3.04	.980	2.757	DS
6.0	29.7	1.28	179.2	382.6	13.15	7.68	3.69	.980	3.749	DS
6.0	30.5	1.14	190.6	443.7	13.15	7.06	3.81	.980	4.347	SL
6.0	38.2	1.16	217.4	544.4	13.15	5.85	4.07	.980	5.334	SL
6.0	48.0	1.16	224.4	687.7	13.15	4.78	4.13	.980	6.737	SL
7.5	5.2	1.24	47.3	69.4	19.43	42.14	1.56	1.225	.680	ST
7.5	5.2	1.08	50.8	79.5	19.43	39.36	1.62	1.225	.779	ST
7.5	11.9	1.34	70.9	146.9	19.43	21.23	1.91	1.225	1.439	DS
7.5	11.9	1.13	80.6	173.4	19.43	19.54	2.04	1.225	1.699	DS
7.5	14.0	1.40	77.1	165.3	19.43	18.82	1.99	1.225	1.619	DS
7.5	14.0	1.14	90.0	203.4	19.43	16.97	2.15	1.225	1.993	DS
7.5	17.5	1.47	87.9	196.5	19.43	15.88	2.13	1.225	1.925	DS
7.5	17.5	1.16	107.0	248.6	19.43	14.12	2.35	1.225	2.435	DS
7.5	21.8	1.57	101.2	230.2	19.43	13.51	2.28	1.225	2.255	DS
7.5	21.8	1.21	130.6	299.0	19.43	11.86	2.59	1.225	2.929	DS
10.0	5.2	1.29	56.7	66.7	32.14	55.25	1.33	1.633	.654	ST DS
10.0	5.2	1.10	60.6	78.0	32.14	51.12	1.37	1.633	.764	ST DS
10.0	13.1	1.44	82.9	150.2	32.14	26.04	1.61	1.633	1.472	DS
10.0	13.1	1.16	94.0	186.3	32.14	23.38	1.71	1.633	1.826	DS
10.0	13.7	1.07	101.0	212.2	32.14	21.55	1.77	1.633	2.079	DS
10.0	21.7	1.09	138.2	330.5	32.14	14.53	2.07	1.633	3.238	DS
10.0	23.4	1.62	110.1	239.1	32.14	16.60	1.85	1.633	2.343	DS
10.0	23.4	1.23	137.2	315.0	32.14	14.47	2.07	1.633	3.087	DS
10.0	26.6	1.10	160.8	400.7	32.14	12.22	2.24	1.633	3.925	SL
10.0	32.8	1.12	182.0	486.3	32.14	10.25	2.38	1.633	4.765	SL

TABLE I.7. PIPE DIA=11.4 MM, DETERGENT= .50 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{let}	SVW	SVL	REGIME
10.0	43.1	1.15	213.0	623.1	32.14	8.18	2.57	1.633	6.104	SL
10.0	52.5	1.17	256.0	743.5	32.14	6.95	2.82	1.633	7.284	SL
12.0	2.7	1.29	60.0	34.7	44.22	114.91	1.16	1.959	.340	ST
12.0	2.7	1.11	61.8	40.4	44.22	106.54	1.18	1.959	.396	ST
12.0	5.1	1.34	67.9	63.3	44.22	67.04	1.24	1.959	.620	SI
12.0	5.1	1.11	71.8	76.1	44.22	61.16	1.27	1.959	.745	ST
12.0	6.0	1.34	68.1	74.4	44.22	58.18	1.24	1.959	.729	ST
12.0	6.0	1.11	72.0	89.3	44.22	53.12	1.28	1.959	.874	ST
12.0	7.9	1.39	76.9	94.3	44.22	46.60	1.32	1.959	.924	ST
12.0	7.9	1.13	83.2	115.6	44.22	42.10	1.37	1.959	1.132	ST
12.0	11.0	1.44	85.4	126.4	44.22	35.55	1.39	1.959	1.238	ST DS
12.0	11.0	1.16	94.2	157.9	44.22	31.81	1.46	1.959	1.547	ST DS
12.0	15.1	1.50	94.3	166.9	44.22	27.47	1.46	1.959	1.635	DS ST
12.0	15.1	1.18	108.2	212.9	44.22	24.33	1.56	1.959	2.086	DS ST
12.0	22.4	1.62	112.5	229.8	44.22	20.19	1.59	1.959	2.251	DS
12.0	22.4	1.22	132.8	304.1	44.22	17.56	1.73	1.959	2.980	DS
12.0	29.7	1.28	162.8	386.6	44.22	14.01	1.92	1.959	3.787	SL
12.0	33.4	1.12	190.2	495.2	44.22	11.84	2.07	1.959	4.852	SL
12.0	44.2	1.15	233.4	635.1	44.22	9.41	2.30	1.959	6.222	SL
16.0	3.2	1.05	94.4	50.5	73.17	114.98	1.14	2.613	.495	ST DS
16.0	9.0	1.07	124.4	139.3	73.17	46.96	1.30	2.613	1.365	DS
16.0	13.1	1.08	141.6	201.1	73.17	33.95	1.39	2.613	1.970	DS
16.0	17.7	1.09	157.6	269.9	73.17	26.18	1.47	2.613	2.644	DS SL
16.0	23.0	1.10	178.2	348.2	73.17	20.89	1.56	2.613	3.411	SL
16.0	29.1	1.12	206.4	433.7	73.17	17.13	1.68	2.613	4.249	SL
16.0	33.4	1.13	227.4	490.8	73.17	15.29	1.76	2.613	4.808	SL
16.0	38.9	1.15	253.8	564.1	73.17	13.47	1.86	2.613	5.527	SL

TABLE I.8. PIPE DIA=17.3 MM, DETERGENT= .50 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{let}	SVW	SVL	REGIME
1.9	29.5	1.11	52.4	444.1	.24	2.61	14.70	.135	1.889	
2.0	6.0	1.03	18.7	97.4	.27	10.60	8.40	.142	.414	
2.3	11.2	1.04	24.3	179.5	.34	6.98	8.47	.163	.764	
2.0	11.8	1.05	26.0	187.8	.27	5.92	9.90	.142	.799	
2.0	23.8	1.10	45.3	362.2	.27	3.28	13.07	.142	1.541	
3.8	6.0	1.01	17.9	99.1	.82	18.43	4.69	.269	.422	
6.0	6.0	1.00	18.2	99.9	1.81	27.37	3.17	.425	.425	
6.1	11.8	1.02	23.6	191.9	1.87	15.55	3.56	.433	.817	
6.2	24.0	1.07	35.1	375.6	1.92	8.64	4.28	.440	1.598	
8.0	30.0	1.08	44.5	464.5	3.00	8.93	3.85	.567	1.976	
11.5	6.0	1.00	19.9	99.8	5.66	48.40	1.87	.815	.424	
11.7	11.9	1.02	26.8	195.0	5.83	27.19	2.14	.830	.829	
11.5	24.0	1.06	41.0	376.9	5.66	14.81	2.69	.815	1.603	
20.4	5.8	.99	25.7	97.3	15.44	81.99	1.29	1.446	.414	
20.0	12.1	1.02	34.2	196.8	14.91	42.99	1.51	1.418	.837	
20.6	25.1	1.08	50.2	388.3	15.70	23.89	1.79	1.461	1.652	
19.9	30.7	1.09	57.4	468.5	14.78	19.56	1.97	1.411	1.993	
20.6	4.9	.99	23.4	82.9	15.70	95.40	1.22	1.461	.353	ST DS
20.6	2.3	.97	18.7	39.5	15.70	183.51	1.09	1.461	.168	ST
20.6	9.6	1.02	29.9	157.3	15.70	53.83	1.38	1.461	.669	ST DS
20.6	11.6	1.03	33.1	187.6	15.70	45.91	1.45	1.461	.798	ST DS
20.6	13.6	1.04	35.2	217.9	15.70	40.13	1.50	1.461	.927	ST DS
20.6	14.7	1.04	36.3	235.2	15.70	37.51	1.52	1.461	1.001	ST DS
20.6	18.6	1.06	40.5	292.1	15.70	30.82	1.61	1.461	1.243	ST DS
20.6	23.3	1.10	45.7	352.8	15.70	25.77	1.71	1.461	1.501	DS SL
20.6	27.2	1.09	49.0	415.9	15.70	22.40	1.77	1.461	1.769	DS ST
20.4	33.0	1.11	52.5	494.8	15.44	18.94	1.84	1.446	2.105	SL
20.6	37.6	1.12	54.6	560.5	15.70	17.09	1.86	1.461	2.385	SL
20.6	42.3	1.14	57.4	616.7	15.70	15.59	1.91	1.461	2.623	DS SL
20.7	48.6	1.16	59.3	700.7	15.84	13.94	1.94	1.468	2.981	SL
38.5	5.3	1.04	51.4	85.1	46.91	158.04	1.05	2.730	.362	
38.5	11.4	1.07	60.4	177.5	46.91	82.09	1.13	2.730	.755	
38.0	26.4	1.16	83.3	379.9	45.85	40.48	1.35	2.694	1.616	

Table I-8: Pipe Dia= 17.3 mm. Detergent= 0.5% Generator Sinter
Flow rate and regime data, no pressure drop results.

Q_W	Q_A	V_W	V_A	Regime	Q_W	Q_A	V_W	V_A	Regime
3.8	5.9	0.43	0.28	ST-S	31.1	5.1	2.29	0.37	ST
6.0	5.2	0.44	0.38	ST		12.9		0.95	ST DS
	13.6		1.00	ST		15.4		1.13	DS ST
	14.6		1.07	ST FM		22.2		1.63	SL
	23.0		1.69	FM		36.3		2.69	SL
	24.5		1.80	FM		52.6		3.87	SL
	26.5		1.95	SL FM	36.9	5.5	2.69	0.40	DS ST
	30.3		2.23	SL FM		14.9		1.09	DS ST
	33.2		2.44	FM SL		15.2		1.12	DS
	48.1		3.54	SL		17.7		1.30	DS
	63.3		4.65	SL		22.3		1.65	SL
10.4	5.2	0.76	0.32	ST		23.9		1.76	SL
	13.5		0.99	ST		25.0		1.84	SL
	14.8		1.09	ST		35.7		2.62	SL
	19.8		1.45	ST		51.6		3.79	SL
	21.5		1.57	ST FM					
	25.3		1.85	SL FM	39.4	25.7	2.89	1.88	SL DS
	26.0		1.91	FM	41.3	6.1	3.04	0.45	DS
	33.4		2.45	SL FM		13.7		1.01	DS
	35.0		2.58	SL		25.4		1.87	SL DS
	44.7		3.29	SL		36.0		2.65	SL
	64.7		4.76	SL					
16.3	5.0	1.20	0.36	ST	44.8	20.4	3.3	1.5	DS
	12.0		0.88	ST	47.2	43.5	3.47	3.20	DS SL
	14.5		1.07	ST		46.9		3.43	SL
	18.3		1.34	ST					
	20.3		1.49	ST SL					
	21.9		1.61	ST SL					
	22.7		1.66	SL					
	26.7		1.96	SL					
	33.4		2.45	SL					
	46.4		3.41	SL					
	65.9		4.84	SL					
22.5	5.1	1.65	0.37	ST					
	12.7		0.93	ST					
	14.4		1.06	ST					
	23.7		1.74	SL					
	34.0		2.5	SL					
	51.1		3.76	SL					
	61.5		4.52	SL					
27.0	5.0	1.98	0.37	ST					
	13.3		0.98	ST					
	15.0		1.1	ST					
	15.2		1.12	ST DS					
	15.9		1.18	DS SL					
	18.8		1.38	SL					
	22.7		1.67	SL					
	34.9		2.59	SL					
	48.9		3.59	SL					

TABLE I.9. PIPE DIA=17.3 MM, DETERGENT= .05 GENERATOR, SINTER

QW	QA	PMID	OPTOT	QA	DPW	X	ϕ_{ut}	SVW	SVA	REGIME
4.0	8.5	1.02	18.3	138.3	.89	14.32	4.53	.284	.588	FM-ST
4.1	8.5	1.03	18.3	138.1	.93	14.64	4.43	.291	.588	ST
6.0	17.0	1.04	27.4	272.9	1.81	11.21	3.89	.425	1.161	FM ST
6.1	17.0	1.08	26.4	263.3	1.87	11.58	3.76	.433	1.120	FM ST
6.1	23.9	1.05	36.0	378.9	1.87	8.49	4.39	.433	1.612	FM
6.1	33.0	1.03	21.1	534.4	1.87	6.34	3.36	.433	2.274	SL
8.4	8.4	1.02	19.4	136.8	3.27	27.68	2.44	.596	.582	ST
10.4	16.9	1.04	29.7	269.9	4.75	18.28	2.50	.737	1.148	ST
10.4	24.0	1.06	38.5	377.8	4.75	13.54	2.85	.737	1.607	DS
10.4	33.7	1.04	25.0	542.4	4.75	9.95	2.29	.737	2.308	SL
14.7	8.3	1.03	23.5	134.4	8.70	45.78	1.64	1.042	.572	ST
16.3	17.0	1.06	34.7	268.5	10.42	27.09	1.82	1.156	1.142	ST
16.3	24.2	1.07	42.9	378.3	10.42	19.99	2.03	1.156	1.610	SL DS
16.3	33.6	1.05	32.0	531.9	10.42	14.91	1.75	1.156	2.263	SL
20.6	8.3	1.04	29.1	132.9	15.70	61.85	1.36	1.461	.565	ST
22.4	16.7	1.07	41.4	259.7	18.18	36.63	1.51	1.588	1.105	ST
22.6	23.9	1.07	42.7	372.3	18.47	26.95	1.52	1.602	1.584	SL
22.6	33.2	1.08	46.5	511.6	18.47	20.32	1.59	1.602	2.177	SL
24.6	7.2	1.05	32.1	114.8	21.42	81.98	1.22	1.744	.488	ST
24.6	8.8	1.06	35.8	138.9	21.42	69.13	1.29	1.744	.591	ST
27.0	17.0	1.08	46.2	261.8	25.21	42.67	1.35	1.914	1.114	SL ST
26.9	23.9	1.09	50.2	366.6	25.05	31.63	1.42	1.907	1.559	SL
27.0	32.4	1.10	58.6	489.3	25.21	24.50	1.52	1.914	2.082	SL
29.1	7.5	1.06	37.9	118.4	28.74	92.08	1.15	2.063	.504	ST
29.1	8.8	1.07	41.5	136.6	28.74	80.72	1.20	2.063	.581	ST
31.1	17.3	1.10	52.3	263.5	32.29	47.81	1.27	2.205	1.121	SL DS
31.1	23.9	1.11	61.0	359.0	32.29	36.28	1.37	2.205	1.527	SL
30.9	33.0	1.11	100.4	494.9	31.92	27.21	1.77	2.191	2.105	SL
33.1	8.0	1.08	45.4	123.6	36.01	98.43	1.12	2.347	.526	ST
37.0	17.2	1.12	64.6	255.4	43.75	56.66	1.22	2.623	1.086	SL
37.0	24.0	1.14	75.4	350.8	43.75	42.66	1.31	2.623	1.493	SL
37.0	33.0	1.17	87.4	472.2	43.75	32.62	1.41	2.623	2.009	SL
40.2	7.5	1.10	57.6	113.4	50.59	124.80	1.07	2.850	.482	
40.2	8.9	1.08	68.6	137.8	50.59	106.18	1.16	2.850	.586	DS ST
41.1	17.3	1.15	75.2	251.3	52.59	62.47	1.20	2.914	1.069	DS SL
41.1	23.9	1.17	85.3	341.7	52.59	47.46	1.27	2.914	1.454	SL DS
41.1	33.7	1.19	88.2	474.4	52.59	35.41	1.30	2.914	2.018	SL DS
47.6	6.8	1.13	87.2	100.1	68.00	159.70	1.13	3.375	.426	DS

TABLE I.10. PIPE DIA=26.0 MM, DETERGENT= .50 GENERATOR, SIMIER

QW	QA	PMIO	DPTOT	QA	DPW	X	ϕ_{ut}	SVW	SVA	REGIME
1.0	14.1	1.04	17.7	226.6	.01	2.75	39.43	.031	.427	
1.0	15.7	1.04	20.3	250.9	.01	2.51	42.23	.031	.473	
1.0	17.5	1.05	22.2	278.8	.01	2.29	44.16	.031	.525	
1.0	19.5	1.05	24.0	309.4	.01	2.08	45.92	.031	.583	
2.0	6.0	1.06	9.9	94.2	.04	10.78	16.08	.063	.177	ST-S
2.3	11.2	1.08	11.2	172.3	.05	7.13	15.13	.072	.324	ST-S
2.0	11.8	1.09	11.4	180.0	.04	6.05	17.26	.063	.339	ST-S
2.0	17.0	1.03	15.3	274.5	.04	4.27	19.99	.063	.517	
2.0	22.4	1.05	21.8	356.9	.04	3.38	23.86	.063	.672	FM
2.0	23.8	1.17	13.4	340.2	.04	3.38	18.71	.063	.641	
2.0	25.0	1.06	25.1	394.5	.04	3.08	25.60	.063	.743	
1.9	29.5	1.19	17.0	412.1	.04	2.71	22.04	.060	.776	
2.0	33.5	1.07	29.1	524.0	.04	2.40	27.57	.063	.987	
3.8	6.0	1.04	9.7	95.8	.12	18.74	9.08	.119	.181	ST-S
4.0	14.2	1.03	12.3	230.6	.13	9.15	9.77	.126	.434	ST FM
4.0	19.6	1.03	14.6	316.8	.13	6.92	10.65	.126	.597	FM
4.0	24.4	1.04	17.7	391.6	.13	5.73	11.72	.126	.738	
4.0	29.8	1.05	23.5	472.8	.13	4.84	13.51	.126	.891	
4.0	37.4	1.05	22.5	593.4	.13	3.97	13.22	.126	1.118	
4.0	47.4	1.04	18.7	759.7	.13	3.21	12.05	.126	1.431	SL FM
4.0	47.9	1.04	18.2	766.9	.13	3.18	11.89	.126	1.444	
4.0	54.2	1.04	17.9	869.9	.13	2.85	11.79	.126	1.638	
4.0	54.4	1.04	16.8	873.1	.13	2.84	11.42	.126	1.644	SL
6.0	6.0	1.04	9.5	96.5	.26	27.86	6.02	.188	.182	ST-W
6.0	11.5	1.02	11.3	187.2	.26	15.67	6.57	.188	.353	ST-S
6.1	11.8	1.07	11.0	183.9	.27	15.88	6.39	.191	.346	
6.0	18.0	1.03	12.9	291.8	.26	10.61	7.02	.188	.550	
6.2	24.0	1.13	12.9	353.9	.28	8.90	6.82	.195	.667	ST-W
6.0	25.2	1.03	15.5	406.1	.26	7.93	7.69	.188	.765	
6.0	30.5	1.04	19.7	486.7	.26	6.74	8.67	.188	.917	
6.0	38.2	1.05	22.5	606.0	.26	5.55	9.27	.188	1.141	
6.0	48.0	1.04	18.8	768.1	.26	4.53	8.47	.188	1.447	
8.2	9.3	1.00	13.9	155.3	.45	24.49	5.54	.257	.292	ST-W
8.2	13.4	1.04	11.4	213.7	.45	18.20	5.02	.257	.403	ST-S
8.2	18.4	1.07	13.8	287.2	.45	13.94	5.52	.257	.541	ST-S
8.2	26.9	1.09	13.3	409.5	.45	10.13	5.42	.257	.771	ST-S
8.0	30.0	1.16	14.6	431.6	.43	9.26	5.81	.251	.813	
8.2	33.3	1.11	14.5	500.0	.45	8.46	5.66	.257	.942	FM ST
8.2	40.2	1.13	16.1	593.4	.45	7.23	5.97	.257	1.118	FM
10.0	3.2	1.17	6.0	45.7	.64	80.14	3.06	.314	.086	
10.0	5.2	1.42	6.2	60.9	.64	57.84	3.11	.314	.115	
10.0	6.0	1.52	6.8	65.9	.64	52.70	3.26	.314	.124	
10.0	6.1	1.47	7.6	68.9	.64	51.21	3.45	.314	.130	
10.0	6.4	1.50	8.1	70.9	.64	49.58	3.56	.314	.134	
10.0	6.6	1.53	8.0	71.8	.64	48.71	3.54	.314	.135	
10.0	6.8	1.57	6.9	72.1	.64	48.09	3.28	.314	.136	
10.0	7.7	1.64	7.8	78.1	.64	44.10	3.49	.314	.147	
10.0	7.9	1.65	8.0	79.9	.64	43.17	3.54	.314	.151	
10.0	8.5	1.71	8.2	83.1	.64	41.20	3.58	.314	.156	
10.0	8.9	1.72	8.8	86.4	.64	39.71	3.71	.314	.163	
10.0	13.7	1.02	11.7	223.0	.64	21.02	4.28	.314	.420	
10.0	21.7	1.03	13.0	351.8	.64	14.09	4.51	.314	.663	
10.0	26.6	1.03	14.1	430.6	.64	11.80	4.69	.314	.811	
10.0	32.8	1.04	16.1	527.9	.64	9.85	5.02	.314	.994	
10.0	43.1	1.04	18.6	690.6	.64	7.77	5.39	.314	1.301	
10.0	52.5	1.05	20.7	836.5	.64	6.56	5.69	.314	1.576	

TABLE I-10. PIPE DIA=26.0 MM, DETERGENT= .50 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LH}	SVW	SVA	REGIME
11.5	6.0	1.04	7.9	95.9	.82	49.37	3.11	.361	.181	ST-W
11.7	11.9	1.07	11.3	185.2	.84	27.90	3.66	.367	.349	
11.5	24.0	1.14	15.0	351.6	.82	15.33	4.28	.361	.662	ST-W
12.0	33.4	1.03	15.7	538.3	.88	11.36	4.22	.377	1.014	
12.0	44.2	1.04	19.1	707.3	.88	8.92	4.66	.377	1.332	
15.5	9.3	1.05	22.5	147.8	1.35	43.32	4.09	.480	.279	ST-W
15.3	16.1	1.06	25.8	252.8	1.35	26.96	4.38	.480	.476	ST-W
15.3	23.3	1.12	26.2	347.0	1.35	20.03	4.41	.480	.654	ST-S
15.3	33.7	1.18	26.6	475.7	1.35	14.90	4.44	.480	.896	ST FM
15.3	44.1	1.22	28.7	600.9	1.35	11.98	4.62	.480	1.132	FM ST
15.3	50.4	1.24	27.6	677.1	1.35	10.74	4.53	.480	1.275	SL FM
15.3	56.3	1.27	29.4	740.4	1.35	9.85	4.67	.480	1.394	SL
16.0	3.2	1.01	7.0	52.6	1.46	112.62	2.19	.502	.099	ST-W
16.0	9.0	1.02	11.4	146.8	1.46	45.77	2.80	.502	.276	ST-W
16.0	13.1	1.02	11.7	213.3	1.46	32.98	2.83	.502	.402	ST-W
16.0	17.7	1.03	12.1	287.7	1.46	25.36	2.88	.502	.542	ST-W
16.0	23.0	1.03	12.7	373.3	1.46	20.18	2.95	.502	.703	
16.0	29.1	1.03	13.9	471.0	1.46	16.45	3.09	.502	.887	
16.0	33.4	1.03	15.1	539.7	1.46	14.60	3.22	.502	1.017	FM ST
16.0	38.9	1.04	16.7	625.8	1.46	12.80	3.39	.502	1.179	
20.4	5.8	1.05	8.7	92.0	2.23	84.29	1.98	.640	.173	ST-W
20.0	12.1	1.10	11.5	183.5	2.15	44.53	2.31	.628	.346	ST-W
20.6	25.1	1.18	14.3	353.8	2.27	25.03	2.51	.647	.666	ST-W
19.9	30.7	1.21	15.2	424.4	2.13	20.56	2.67	.625	.799	
28.5	16.5	1.17	24.9	234.7	4.00	47.78	2.49	.895	.442	ST-S
28.5	21.2	1.27	24.6	279.1	4.00	39.89	2.48	.895	.526	ST-W
28.5	27.6	1.34	24.3	343.7	4.00	32.56	2.46	.895	.647	ST-W
28.5	11.1	1.06	25.7	173.8	4.00	64.42	2.53	.895	.327	ST-W
28.5	36.9	1.42	24.1	432.5	4.00	26.03	2.45	.895	.815	ST SL
28.5	46.6	1.50	26.9	516.7	4.00	21.82	2.59	.895	.973	SL ST
28.5	57.6	1.57	26.3	610.4	4.00	18.54	2.56	.895	1.150	SL ST
38.5	5.3	1.16	9.4	76.1	6.77	167.09	1.18	1.209	.143	ST-W
38.5	11.4	1.22	12.8	156.3	6.77	87.51	1.37	1.209	.294	ST-W
38.0	26.4	1.35	17.5	327.0	6.62	43.66	1.63	1.193	.616	ST-W

TABLE I-11. PIPE DIA=39.1 MM, DETERGENT= .50 GENERATOR, SINTER

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LH}	SVW	SVA	REGIME
2.0	6.0	1.08	6.6	92.5	.01	10.88	34.60	.028	.077	ST-S
2.0	11.8	1.11	7.0	176.4	.01	6.11	35.63	.028	.147	ST-S
2.3	11.2	1.11	6.9	168.8	.01	7.20	31.31	.032	.141	ST-s
2.0	23.8	1.19	7.3	333.0	.01	3.42	36.39	.028	.277	FM
1.9	29.5	1.22	7.7	402.8	.01	2.74	39.09	.026	.335	
3.8	6.0	1.06	6.3	94.1	.02	18.91	19.28	.053	.078	ST-S
6.0	6.0	1.05	5.9	94.8	.04	28.10	12.51	.083	.079	ST-S
6.1	11.8	1.09	6.8	180.3	.04	16.05	13.24	.085	.150	ST-s
6.2	24.0	1.16	7.2	346.3	.04	9.00	13.43	.086	.288	ST-S
8.0	30.0	1.19	7.1	421.6	.06	9.37	10.67	.111	.351	ST-S
11.5	6.0	1.06	5.4	94.3	.12	49.80	6.77	.160	.079	ST-W
11.7	11.9	1.09	6.1	181.5	.12	28.19	7.09	.162	.151	ST-W
11.5	24.0	1.16	6.7	344.0	.12	15.50	7.54	.160	.287	ST-S
20.4	5.8	1.07	4.9	90.4	.32	85.03	3.91	.283	.075	ST-w
20.0	12.1	1.12	6.0	179.7	.31	45.00	4.40	.278	.150	ST-w
20.6	25.1	1.21	6.0	345.5	.33	25.33	4.29	.286	.288	ST-W
19.9	30.7	1.24	6.7	413.9	.31	20.82	4.67	.276	.345	ST-W
38.5	5.3	1.18	4.2	74.6	.98	168.77	2.08	.534	.062	ST-W
38.5	11.4	1.25	5.4	152.6	.98	88.56	2.35	.534	.127	ST-W
38.0	26.4	1.38	6.0	318.2	.95	44.26	2.51	.527	.265	ST-w

TABLE I-12. PIPE DIA=11.4 MM, DETERGENT= .50 GENERATOR, PUMP

QW	QA	PMID	OPTOT	QA	DPW	X	ϕ_{uc}	SVW	SA	REGIME
3.0	2.6	1.11	21.7	38.5	3.91	32.90	2.36	.490	.377	
3.0	2.6	1.04	24.0	41.1	3.91	31.84	2.48	.490	.403	
3.0	5.2	1.18	37.2	72.4	3.91	18.50	3.08	.490	.709	
3.0	5.2	1.06	41.2	80.7	3.91	17.52	3.25	.490	.791	
4.0	1.8	1.05	11.3	28.5	6.47	56.48	1.32	.653	.279	
4.0	1.8	1.01	11.6	29.5	6.47	55.47	1.34	.653	.289	
4.0	2.4	1.08	19.7	36.9	6.47	44.52	1.75	.653	.362	
4.0	2.4	1.02	20.2	39.0	6.47	43.31	1.77	.653	.382	
4.0	3.6	1.13	28.2	53.1	6.47	31.91	2.09	.653	.520	
4.0	3.6	1.03	29.8	57.9	6.47	30.56	2.15	.653	.567	
4.0	4.1	1.13	27.6	60.4	6.47	28.48	2.07	.653	.592	
4.0	4.1	1.03	29.4	66.1	6.47	27.24	2.13	.653	.647	
4.0	5.5	1.17	37.1	78.3	6.47	22.41	2.40	.653	.767	
4.0	5.5	1.04	39.8	87.6	6.47	21.18	2.48	.653	.858	
4.0	5.7	1.17	38.2	81.0	6.47	21.74	2.43	.653	.793	
4.0	5.7	1.05	41.6	90.6	6.47	20.56	2.54	.653	.887	
4.0	6.0	1.17	38.5	84.8	6.47	20.83	2.44	.653	.831	
4.0	6.0	1.04	42.2	95.3	6.47	19.65	2.55	.653	.934	
4.0	7.1	1.21	45.8	97.7	6.47	18.23	2.66	.653	.957	
4.0	7.1	1.05	49.4	111.8	6.47	17.04	2.76	.653	1.095	
4.0	8.6	1.24	52.1	115.0	6.47	15.63	2.84	.653	1.127	
4.0	8.6	1.06	60.6	134.6	6.47	14.45	3.06	.653	1.319	
4.0	9.0	1.24	50.1	120.8	6.47	15.00	2.78	.653	1.183	
4.0	9.0	1.06	58.6	140.7	6.47	13.90	3.01	.653	1.378	
4.0	12.1	1.28	55.9	157.6	6.47	11.75	2.94	.653	1.544	
4.0	12.1	1.08	68.8	186.6	6.47	10.80	3.26	.653	1.828	
5.0	1.7	1.09	17.3	25.6	9.56	73.96	1.35	.816	.251	
5.0	1.7	1.03	18.4	27.0	9.56	72.03	1.39	.816	.265	
5.0	3.2	1.12	24.9	46.8	9.56	43.17	1.61	.816	.458	
5.0	3.2	1.04	27.2	50.5	9.56	41.56	1.69	.816	.495	
5.0	6.1	1.20	40.5	83.6	9.56	25.35	2.06	.816	.819	
5.0	6.1	1.06	46.0	94.2	9.56	23.89	2.19	.816	.923	
5.0	9.5	1.27	55.6	122.4	9.56	17.74	2.41	.816	1.199	
5.0	9.5	1.09	65.6	143.6	9.56	16.39	2.62	.816	1.407	
6.0	2.9	1.11	26.0	43.2	13.15	54.69	1.41	.980	.423	
6.0	2.9	1.02	28.0	46.7	13.15	52.60	1.46	.980	.457	
6.0	4.5	1.14	33.2	65.0	13.15	37.80	1.59	.980	.637	
6.0	4.5	1.03	36.2	71.9	13.15	35.94	1.66	.980	.704	
6.0	6.1	1.20	41.4	83.4	13.15	29.78	1.77	.980	.817	
6.0	6.1	1.06	46.4	94.2	13.15	28.02	1.88	.980	.923	
6.0	7.6	1.21	48.7	103.3	13.15	24.64	1.92	.980	1.012	
6.0	7.6	1.05	55.8	119.3	13.15	22.93	2.06	.980	1.169	
6.0	9.7	1.29	58.0	123.8	13.15	20.53	2.10	.980	1.213	
6.0	9.7	1.09	67.8	146.3	13.15	18.89	2.27	.980	1.433	
6.0	11.1	1.28	62.2	142.4	13.15	18.20	2.18	.980	1.395	
6.0	11.1	1.07	73.8	170.7	13.15	16.63	2.37	.980	1.673	
6.0	13.8	1.37	74.8	166.5	13.15	15.51	2.39	.980	1.631	
6.0	13.8	1.10	93.4	206.9	13.15	13.92	2.67	.980	2.027	
6.0	20.0	1.47	91.2	223.7	13.15	11.65	2.63	.980	2.191	
6.0	20.0	1.14	119.2	290.1	13.15	10.23	3.01	.980	2.843	
6.0	20.0	1.47	91.7	223.8	13.15	11.64	2.64	.980	2.193	
6.0	20.0	1.13	117.4	290.3	13.15	10.22	2.99	.980	2.844	
6.0	26.4	1.60	99.7	271.5	13.15	9.53	2.75	.980	2.660	
6.0	26.4	1.19	167.4	364.9	13.15	8.21	3.57	.980	3.575	
6.0	32.0	1.23	200.2	429.3	13.15	7.05	3.90	.980	4.206	
7.5	1.7	1.13	28.2	24.6	19.43	107.63	1.20	1.225	.241	
7.5	1.7	1.04	28.8	26.8	19.43	103.20	1.22	1.225	.262	

Flow dispersed at all times

TABLE 1-2. PIPE DIA=11.4 MM, DETERGENT= .50 GENERATOR, PUMP

QW	QA	PMID	DPTOT	QA	DPW	X	Φ_{ut}	SVW	SA	REGIME
7.5	3.9	1.17	35.9	54.7	19.43	52.87	1.36	1.225	.536	
7.5	3.9	1.05	38.2	60.7	19.43	50.18	1.40	1.225	.595	
7.5	6.9	1.23	47.7	92.0	19.43	32.90	1.57	1.225	.902	
7.5	6.9	1.07	55.2	105.8	19.43	30.69	1.69	1.225	1.036	
7.5	10.7	1.31	63.2	133.9	19.43	23.14	1.80	1.225	1.312	
7.5	10.7	1.10	74.4	160.3	19.43	21.14	1.96	1.225	1.571	
10.0	2.0	1.18	40.6	28.1	32.14	121.87	1.12	1.633	.275	
10.0	2.0	1.04	42.2	31.8	32.14	114.53	1.15	1.633	.312	
10.0	3.7	1.21	49.2	50.4	32.14	72.23	1.24	1.633	.494	
10.0	3.7	1.05	51.8	58.3	32.14	67.19	1.27	1.633	.571	
10.0	3.8	1.05	53.2	60.6	32.14	65.21	1.29	1.633	.594	
10.0	3.8	1.21	50.2	52.4	32.14	70.12	1.25	1.633	.514	
10.0	4.2	1.22	51.3	56.8	32.14	64.90	1.26	1.633	.556	
10.0	4.2	1.05	53.4	65.9	32.14	60.22	1.29	1.633	.646	
10.2	6.0	1.06	65.8	94.5	33.28	44.77	1.41	1.666	.926	
10.0	6.5	1.26	60.1	85.1	32.14	45.01	1.37	1.633	.833	
10.0	6.5	1.07	65.0	100.7	32.14	41.38	1.42	1.633	.986	
10.0	7.4	1.06	69.6	116.5	32.14	36.64	1.47	1.633	1.142	
10.0	7.4	1.27	62.9	97.0	32.14	40.14	1.40	1.633	.951	
10.0	9.9	1.32	71.0	123.6	32.14	31.89	1.49	1.633	1.211	
10.0	9.9	1.08	79.6	151.2	32.14	28.83	1.57	1.633	1.481	
10.0	10.3	1.33	72.1	127.9	32.14	30.88	1.50	1.633	1.253	
10.0	10.3	1.08	82.0	156.9	32.14	27.88	1.60	1.633	1.538	
10.3	13.0	1.07	97.2	202.5	33.85	23.08	1.69	1.682	1.984	
10.0	13.8	1.38	82.6	164.2	32.14	24.42	1.60	1.633	1.609	
10.0	13.8	1.10	99.2	206.0	32.14	21.81	1.76	1.633	2.018	
10.0	14.4	1.40	84.6	169.9	32.14	23.63	1.62	1.633	1.665	
10.0	14.4	1.11	100.6	215.3	32.14	21.00	1.77	1.633	2.109	
10.0	14.8	1.11	102.2	223.7	32.14	20.39	1.78	1.633	2.192	
10.0	14.8	1.40	85.8	176.2	32.14	22.96	1.63	1.633	1.727	
10.2	18.8	1.08	126.2	289.3	33.28	16.67	1.95	1.666	2.835	
10.0	18.9	1.49	98.1	209.0	32.14	19.24	1.75	1.633	2.048	
10.0	18.9	1.14	126.4	273.6	32.14	16.82	1.98	1.633	2.681	
10.0	23.3	1.58	109.6	243.6	32.14	16.48	1.85	1.633	2.386	
10.0	23.3	1.17	148.4	327.8	32.14	14.20	2.15	1.633	3.212	
10.0	28.0	1.21	182.6	381.7	32.14	12.28	2.38	1.633	3.740	
10.0	30.2	1.23	197.4	403.8	32.14	11.61	2.48	1.633	3.956	
12.0	5.4	1.33	70.2	66.8	44.22	63.85	1.26	1.959	.655	
12.0	5.4	1.09	75.0	81.0	44.22	58.01	1.30	1.959	.793	
12.0	9.7	1.40	83.1	113.7	44.22	39.29	1.37	1.959	1.114	
12.0	9.7	1.12	94.2	142.5	44.22	35.11	1.46	1.959	1.396	
12.0	11.4	1.43	88.9	130.7	44.22	34.49	1.42	1.959	1.281	
12.0	11.4	1.12	102.8	166.8	44.22	30.54	1.52	1.959	1.634	
12.0	18.9	1.55	109.0	200.7	44.22	23.03	1.57	1.959	1.966	
12.0	18.9	1.15	130.4	268.7	44.22	19.91	1.72	1.959	2.632	
16.0	3.8	1.11	99.8	57.3	73.17	101.22	1.17	2.613	.561	
16.0	6.9	1.11	111.2	103.5	73.17	60.19	1.23	2.613	1.014	
16.0	11.0	1.14	132.6	160.6	73.17	40.57	1.35	2.613	1.573	
16.0	17.4	1.19	153.2	243.4	73.17	27.75	1.45	2.613	2.384	

Flow dispersed at all times

TABLE 13. PIPE DIA=17.3 MM, DETERGENT= .50 GENERATOR, PUMP.

QW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LE}	SVW	SVL	REGIME
5.0	1.8	1.00	5.8	29.8	1.32	67.11	2.10	.355	.127	
5.0	3.5	1.01	11.5	57.5	1.32	37.67	2.95	.355	.244	
5.0	4.8	1.02	15.3	78.3	1.32	28.66	3.41	.355	.333	
5.0	5.2	1.01	15.9	85.6	1.32	26.60	3.47	.355	.364	
4.9	6.9	1.02	19.9	112.0	1.27	20.55	3.96	.347	.477	
5.0	8.4	1.02	20.2	136.5	1.32	17.60	3.92	.355	.581	
4.9	9.7	1.03	21.4	157.0	1.27	15.28	4.10	.347	.668	
5.0	11.6	1.03	22.4	188.0	1.32	13.29	4.12	.355	.800	
5.0	15.7	1.02	23.0	256.1	1.32	10.16	4.18	.355	1.090	
5.0	15.7	1.03	23.2	253.8	1.32	10.21	4.20	.355	1.080	
5.0	16.0	1.03	23.3	258.7	1.32	10.04	4.20	.355	1.100	
5.0	19.2	1.03	21.8	311.1	1.32	6.55	4.07	.355	1.323	
5.0	23.4	1.02	17.5	380.0	1.32	7.18	3.64	.355	1.617	
5.0	26.4	1.02	15.7	430.5	1.32	6.45	3.45	.355	1.831	
5.0	47.3	1.02	19.5	767.0	1.32	3.88	3.85	.355	3.263	
8.2	10.8	1.02	21.5	177.1	3.13	21.68	2.62	.581	.753	
8.2	17.5	1.04	30.3	281.1	3.13	14.36	3.11	.581	1.196	
8.2	22.5	1.05	35.1	359.1	3.13	11.56	3.35	.581	1.528	
10.3	3.3	1.01	8.0	54.5	4.67	74.40	1.31	.730	.232	
10.4	5.0	1.01	11.4	82.3	4.75	52.26	1.55	.737	.350	
10.2	6.0	1.18	11.3	84.7	4.59	47.31	1.57	.723	.360	
10.3	7.6	1.00	14.0	126.0	4.67	35.79	1.73	.730	.536	
10.4	10.8	1.01	19.0	177.1	4.75	26.69	2.00	.737	.753	
10.3	13.0	1.28	20.5	169.0	4.67	25.27	2.10	.730	.719	
10.5	19.4	1.04	31.4	311.4	4.83	16.29	2.55	.744	1.325	
10.2	18.8	1.38	26.0	226.7	4.59	18.84	2.38	.723	.964	
10.3	29.2	1.05	43.0	461.4	4.67	11.29	3.04	.730	1.963	
10.3	31.5	1.06	44.5	495.4	4.67	10.59	3.09	.730	2.108	
10.3	34.5	1.06	46.5	539.8	4.67	9.80	3.16	.730	2.297	
10.3	39.4	1.06	48.0	616.4	4.67	8.73	3.21	.730	2.622	
10.2	44.9	1.07	50.8	700.0	4.59	7.73	3.33	.723	2.978	
10.3	53.8	1.06	42.6	845.2	4.67	6.63	3.02	.730	3.596	
13.1	10.1	1.01	21.5	166.5	7.11	34.54	1.74	.929	.708	
13.1	17.6	1.03	30.9	284.7	7.11	21.45	2.08	.929	1.211	
13.1	23.9	1.05	39.5	380.7	7.11	16.54	2.36	.929	1.620	
13.1	29.9	1.06	47.0	470.6	7.11	13.68	2.57	.929	2.002	
13.1	36.9	1.07	52.5	576.7	7.11	11.42	2.72	.929	2.454	
16.0	3.8	1.29	12.8	49.2	10.09	109.21	1.13	1.134	.209	
16.0	6.9	1.31	17.0	87.5	10.09	65.47	1.30	1.134	.372	
16.0	11.0	1.42	20.5	129.4	10.09	45.20	1.43	1.134	.551	
16.0	17.4	1.49	26.0	194.1	10.09	31.08	1.61	1.134	.826	
19.9	3.9	1.00	17.5	64.9	14.78	113.95	1.09	1.411	.276	
20.1	4.3	1.00	19.1	71.3	15.04	105.72	1.13	1.425	.304	
20.1	9.3	1.01	24.0	152.8	15.04	54.08	1.26	1.425	.650	
19.9	11.3	1.02	26.3	184.9	14.78	45.31	1.33	1.411	.787	
20.1	13.4	1.02	30.0	217.8	15.04	39.51	1.41	1.425	.926	
19.9	20.4	1.04	38.2	326.9	14.78	27.30	1.61	1.411	1.391	
19.9	33.2	1.07	52.5	516.6	14.78	18.09	1.88	1.411	2.198	
20.1	41.7	1.09	62.0	637.7	15.04	15.06	2.03	1.425	2.713	
30.2	4.7	1.02	34.9	76.9	30.67	140.66	1.07	2.141	.327	
30.4	9.3	1.03	41.5	149.8	31.02	78.45	1.16	2.155	.637	
30.0	13.1	1.04	45.1	209.8	30.31	57.62	1.22	2.127	.893	
30.1	19.3	1.06	53.3	304.2	30.49	41.50	1.32	2.134	1.294	
30.0	19.8	1.06	52.7	312.1	30.31	40.47	1.32	2.127	1.328	
30.2	33.0	1.08	67.4	508.1	30.67	26.34	1.48	2.141	2.162	
30.2	43.1	1.11	76.5	648.3	30.67	21.09	1.58	2.141	2.758	

Flow dispersed at all times

TABLE I-14. PIPE DIA=11.4 MM, DETERGENT= .05 GENERATOR, PUMP

GW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LH}	SVW	SVA	REGIME
7.5	3.1	1.15	33.7	44.9	19.43	63.58	1.32	1.225	.440	Flow dispersed at all times
7.5	3.1	1.03	35.0	49.9	19.43	60.34	1.34	1.225	.489	
7.5	4.7	1.18	42.2	66.0	19.43	44.85	1.47	1.225	.647	
7.5	4.7	1.04	44.6	74.9	19.43	42.11	1.52	1.225	.734	
7.5	6.7	1.23	52.1	90.3	19.43	33.57	1.64	1.225	.885	
7.5	6.7	1.06	56.6	105.4	19.43	31.08	1.71	1.225	1.033	
7.5	8.3	1.27	59.1	108.5	19.43	28.27	1.74	1.225	1.063	
7.5	8.3	1.07	65.8	129.5	19.43	25.88	1.84	1.225	1.268	
7.5	10.3	1.29	64.6	132.2	19.43	23.61	1.82	1.225	1.296	
7.5	10.3	1.07	72.6	159.4	19.43	21.52	1.93	1.225	1.561	
7.5	13.9	1.32	71.8	175.0	19.43	18.35	1.92	1.225	1.714	
7.5	13.9	1.08	78.0	213.8	19.43	16.60	2.00	1.225	2.095	

TABLE I-15. PIPE DIA=17.3 MM, DETERGENT= .50 GENERATOR, VALVE

GW	QA	PMID	DPTOT	QA	DPW	X	ϕ_{LH}	SVW	SVA	REGIME
10.3	4.6	.99	9.6	77.1	4.67	55.23	1.43	.730	.328	Flow dispersed at all times
10.4	5.6	1.02	14.1	91.9	4.75	47.41	1.72	.737	.391	
10.4	11.1	1.01	20.9	182.5	4.75	26.02	2.10	.737	.777	
10.2	12.6	1.02	23.2	205.8	4.59	22.97	2.25	.723	.876	
10.4	13.6	1.02	24.2	222.2	4.75	21.85	2.26	.737	.945	
10.5	15.9	1.03	28.0	257.7	4.83	19.30	2.41	.744	1.096	
10.4	19.4	1.04	32.1	312.3	4.75	16.14	2.60	.737	1.329	
10.5	23.8	1.05	37.8	379.1	4.83	13.68	2.80	.744	1.613	
12.9	30.2	1.06	46.8	473.6	6.92	13.40	2.60	.915	2.015	
14.5	19.2	1.04	33.9	309.1	8.49	21.78	2.00	1.028	1.315	
14.7	19.7	1.04	34.6	316.7	8.70	21.56	1.99	1.042	1.347	
14.3	24.9	1.13	40.5	367.2	8.29	17.90	2.21	1.014	1.562	
15.2	37.9	1.07	52.2	588.3	9.22	12.75	2.38	1.078	2.503	
15.9	17.9	1.12	32.3	267.0	9.98	26.08	1.80	1.127	1.136	
18.1	49.4	1.12	70.2	735.6	12.52	12.02	2.37	1.283	3.129	
18.9	58.3	1.13	83.4	864.8	13.50	10.82	2.49	1.340	3.679	
20.6	4.3	.98	18.8	73.2	15.70	106.64	1.09	1.461	.311	
20.6	8.3	1.00	23.5	138.9	15.70	60.50	1.22	1.461	.591	
20.6	15.2	1.03	32.5	245.5	15.70	36.27	1.44	1.461	1.044	
20.6	21.3	1.06	40.1	336.1	15.70	27.31	1.60	1.461	1.430	
20.6	28.9	1.09	48.2	443.4	15.70	21.20	1.75	1.461	1.886	
20.6	34.3	1.11	53.7	513.8	15.70	18.47	1.85	1.461	2.186	
20.6	45.6	1.16	64.2	658.1	15.70	14.67	2.02	1.461	2.800	
27.2	40.3	1.10	67.3	611.0	25.54	20.33	1.62	1.929	2.599	
26.9	67.1	1.17	96.7	961.1	25.05	13.26	1.96	1.907	4.089	
31.1	7.7	1.03	39.5	124.8	32.29	94.13	1.11	2.205	.531	
30.9	19.6	1.06	53.6	309.1	31.92	41.89	1.30	2.191	1.315	
31.1	32.7	1.09	67.5	500.5	32.29	27.32	1.45	2.205	2.129	
31.1	52.4	1.13	84.8	773.8	32.29	18.41	1.62	2.205	3.292	

Appendix J:

FLOW REGIMES FOR FOAM IN HORIZONTAL PIPES

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Previous flow pattern maps that have been used for correlating two-phase flows have been devised from data with a wide range of variables, pipe size and phase properties, and have resulted in the type of flow pattern map as shown in Fig. 1 {1}. These maps are for liquid/gas systems which are non-foaming. Here the effect of foaming agent on the flow regimes has been investigated.

SYSTEM INVESTIGATED

The foaming agent used was "Teepol L", a secondary alkyl sulphate, CMC of 0.1% volume and a surface tension of 31 dynes/cm. The changes in flow regime that occurred were due to the ability to form and maintain a system of small bubbles in the liquid, rather than the effect of the change of liquid properties which previous work has shown to be small.

The flow regimes which were obtained depended on the mode of gas/liquid mixing although in the length of piping used in the test (≈ 6 metres) each injection method established the regimes after a short entrance length. The methods of injection gave three ranges of gas/liquid interaction (i.e. bubble size):

- a) very low mixing; a simple T-piece in which the gas and liquid are introduced as an upper and lower layer, with no bubbles being formed.
- b) good mixing; the gas is blown into the liquid via a sinter disc, bubble size of .2 mm to 1 mm.
- c) intense mixing; the gas/liquid stream is passed through a mixing valve and a centrifugal pump, bubble size of 20-100 microns.

FLOW REGIMES

The flow regimes present in the experiments (established by visual means) with the foaming agent were:

- 1) dispersed flow; bubbles of gas in liquid such that they are concentrated in upper part of pipe but present throughout the liquid.
- 2) stratified foam/liquid flow; bubbles fill the upper part of the pipe (i.e. foam) and little coalescence occurs. The lower part of the pipe is filled with liquid, clear of bubbles. The liquid layer moves faster than the foam layer. The interface can be flat or wavy. This region would be elongated bubble in pure liquid/gas flow.
- 3) foam flow; the foam fills the pipe completely.
- 4) foam/air slug flow; the foam flows in the pipe with slugs of air flowing faster than the foam. The slugs are formed by breakdown of the foam and are not formed in the gas/liquid mixing process.

(Air flows such as to be within the annular flow region were not reached with this equipment.)

EXPERIMENTS & RESULTS

Four levels of foaming agent were investigated:

- i) Zero - pure water.
- ii) Trace - 0.002% by volume; surface tension = 59 dynes/cm. A foam of only a few seconds persistence could form.
- iii) Below CMC - 0.05% by volume; surface tension = 36 dynes/cm. A stable foam formed.
- iv) Above CMC - 0.5% by volume; surface tension = 31 dynes/cm. A stable foam formed.

Flow pattern maps for all injection methods with pure water were carried out as a check on the equipment and were found to be as reported in the literature, within experimental error.

The flow pattern map for the trace of foaming agent gave equivalent results to that of pure water, any bubbles produced breaking down to form elongated bubbles or slugs.

The flow patterns for 0.05% and 0.5% foaming agent under low mixing conditions were again similar to that of water. However, at high water rates (greater than 1 m/s) in the slug regime it was obvious that the slug was slowly breaking down as small bubbles were formed, with a sufficient length of pipe the pattern could change to dispersed flow.

The flow patterns for the 0.05% and 0.5% foaming agents under good mixing conditions were as shown in Figs. 2 and 3, for a 17 mm pipe. The plots can be seen to be substantially similar in form, with a gross shift to the right at higher foaming agent (i.e. beyond CMC) concentrations. This is what might be expected from the fact that a level beyond the CMC would make the bubbles less liable to breakage and hence slug formation.

The effect on flow pattern of a smaller pipe diameter, 11 mm, is as shown in Fig. 4. This shows a large shift of the slugging region to higher air rates, the other regions being about equivalent. This major shift shows that in foam flow there is a dependence on pipe size, maybe a reflection on the bubble to pipe diameter ratio. However, pressure drops in the 11 mm pipe are such that accelerating forces are important and could start to affect the flow pattern.

Under conditions of intense mixing the flow pattern for 0.05% and 0.5% solution was at all times dispersed and no slugging occurred, even at maximum air rates of ≈ 5 m/s in 17 mm pipe and ≈ 10 m/s in 11 mm pipe.

REFERENCE

1. Mandhane, J.M., Gregory G.A. & Aziz, K., Int.Jl.of Multiphase Flow (1974) 1, 537-553. "A flow pattern map for gas-liquid flow in horizontal pipes".

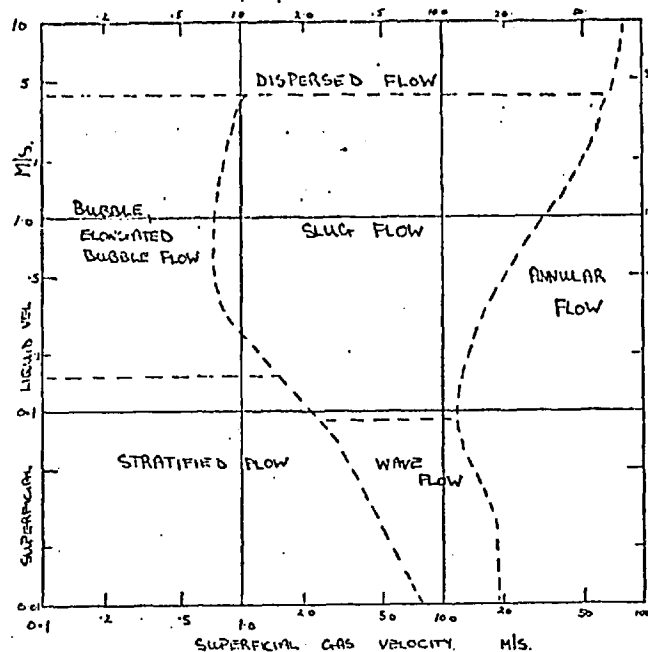


Fig 1 Flow Pattern map for Air/water. (1).

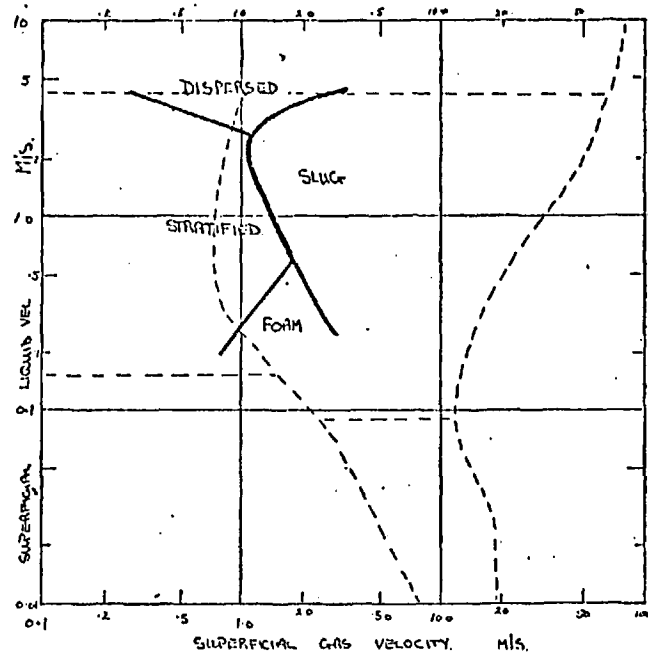


Fig 2. Flow Pattern map for Air/(0.052 cSt) in 17 mm pipe.

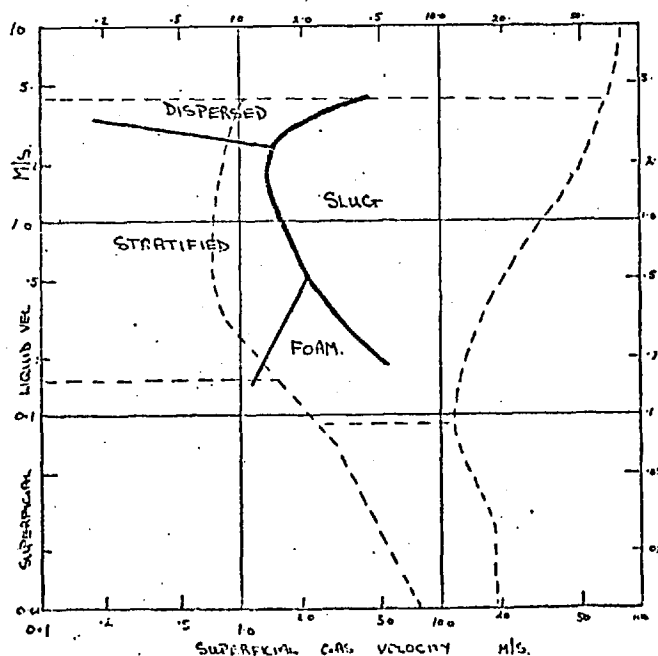


Fig 3. Flow Pattern map for Air/(0.52 cSt) in 17 mm pipe.

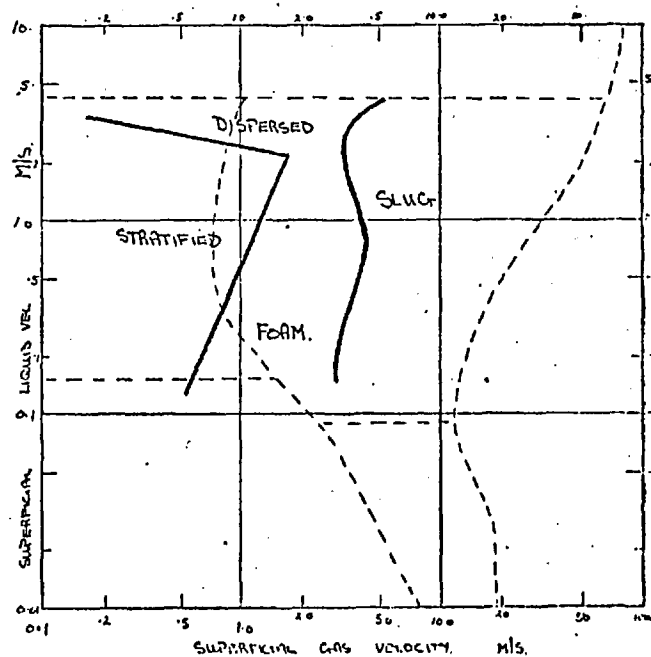


Fig 4. Flow Pattern map for Air/(0.52 cSt) in 11 mm pipe.