AN EXPERIMENTAL STUDY OF CROSSFLOW FILTRATION AND THE DESIGN OF A PROTOTYPE CROSSFLOW FILTER

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by

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SUMMARY

There were three objectives for the research reported in this thesis. Firstly to design a prototype crossflow filter for the filtration of injection water used in oil extraction. Secondly to perform some experiments into crossflow filtration to gain data for the first objective, and also knowledge about the process of crossflow filtration. Finally, it was intended to gain a theorectical understanding of the process.

Crossflow filtration differs from cake filtration in the direction of flow of the filtrate and feed. In cake filtration, the feed and filtrate both flow in the same direction, whereas in crossflow filtration, the feed flows parallel to the filter medium, and the filtrate flows through the medium. This leads to a number of advantages which are described in the Introduction.

Frequently in the extraction of oil, water is injected into the oil reservoir, to increase the production of oil. This water, known as injection water, has to be treated very carefully to avoid damage to the formation. It is thought that crossflow filtration might provide a viable method of filtering injection water. This is because the present methods use large and bulky filters, and space is very expensive on an offshore oil production platform. Crossflow filters promise to be more compact, and this is the reason for the interest in this method of filtration for this application.

Several sets of experiments are reported in this thesis. The effect of backflushing variables on the effectiveness of a backflush was studied. It was found that this effectiveness was dependent on the backflush pressure, and almost independent of the backflush duration. It was also found that the optimal operating cycle is one with frequent, short, high pressure backflushes. With this strategy, a constant average filtrate flux was achieved.

Experiments are reported that tested whether the crossflow channel geometry has any influence on the filtrate flux. No evidence was found to suggest that the channel geometry had any effect.

Experiments to ascertain the effect of pressure and crossflow velocity on the particle concentration in the filtrate are reported. There is again no evidence to suggest that these variables do influence the filtrate quality, but the data obtained are not completely reliable.

These experiments were used to design a prototype crossflow filter. This prototype is designed to filter 6 litres per second and consists of 8 separate plates. Triangular crossflow channels and square filtrate channels are developed in the design. The design procedure included a computer analysis to optimize this channel geometry.

The major omission in the design is that the membrane is not specified. A survey of available membranes is included in the thesis, as is a suggested procedure for testing these membranes for suitability. Part of the procedure has already been fulfilled for one of the more suitable membranes.

Several suggestions as to how a theorectical understanding of the process might be achieved are given in the Discussion. Only a cursory attempt at modelling the process has been reported in this thesis, but a number of approaches that might be used are described.

INTRODUCTION

When the work that is reported in this thesis began, there were three objectives. The first objective was to design a prototype crossflow filtration system for the filtration of injection water used in oil extraction. The second objective was to perform sufficient experiments to enable the process variables to be optimised in that design. The third objective was to gain a theoretical understanding of crossflow filtration. Thus, there are three aspects to this project, at three different levels of generality. Firstly, there is the design of a prototype for a particular application. Secondly there is the experimentation with crossflow filtration. Finally there is the theoretical aspect of a particular membrane process. This section will introduce these various aspects of the thesis.

CROSSFLOW FILTRATION.

In most filtration processes the direction of the feed and filtrate flows is directly through the filter medium. The feed divides into two different streams, the filtered fluid, and the cake which is collected on the filter medium. This is shown diagrammatically in Fig 2.1. This flow arrangement causes the filtration to be essentially a batch operation. The cake becomes thicker as more fluid is filtered, until the thickness makes the filtration rates too slow. Then the cake is removed and the process recommenced.

There are ways in which the operation can be made continuous, by removing the cake continously. These adaptions usually involve large capital costs, and the feed concentration must be relatively high for the cake to become thick enough to be removed continuously. In the case where the feed suspension is very fine, the particles are not collected on the surface of the medium, but tend to block the pores. In this case, the volume filtered before the medium is completely blocked is low, and it is very difficult to recover the particulate material. Cake filtration is therefore not suitable for fine or dilute suspensions.

The usual way to filter these fine suspensions is some form of depth filtration. These filters use a bed of fine material (for example sand, gravel, or diatomaceous earth) to filter and collect the particulate material in the feed. The particles that are filtered become attached to the bed material and are hard to remove. For this reason the particles in the feed suspension are not usually recovered. After a lengthy period of operation, the bed has to be reactivated. These filters are very large bulky pieces of plant, and although the running costs are low, they are unsuitable for some applications.

For these reasons crossflow filtration was introduced. The principle of operation is more similar to other membrane processes like reverse osmosis and ultrafiltration, than cake or depth filtration. Here the flow of the feed is parallel to the filter medium. A pressure difference causes some of the fluid to pass through the filter medium, and thus clarified. There is also some fluid that does not pass through the medium and this fluid is called the reject. Thus there are three flows involved, the feed and the reject which travel parallel to the medium surface, and the filtrate flow which travels perpendicular to the medium. This situation is shown diagrammatically in Fig 2.2.

The major difference between crossflow and cake filtration, is that the growth of the cake on the medium surface is limited by the shear stress of the feed flowing parallel to the surface. Thus the thickened product is not collected as a cake, but as a concentrated (relative to the feed) suspension. When a very fine suspension is filtered in this way, the membrane does not become blocked as quickly as for cake filtration. Thus crossflow filtration can be used to filter dilute suspensions with fine particulate material. The thickened product can not be made as concentrated as in cake filtration, because it has to

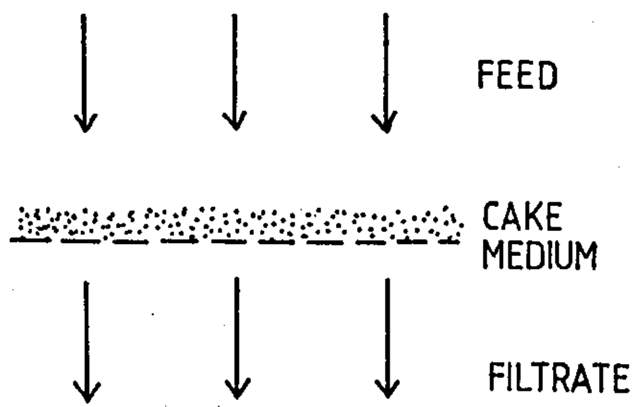


Fig. 2.1 Diagrammatic representation of Cake Filtration.

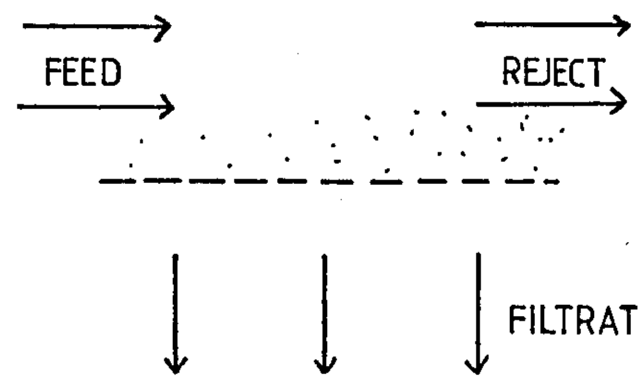


Fig. 2.2 Diagrammatic representation of Crossflowfiltration

flow along as a liquid.

The original concept of crossflow filtration was that the fluid flowing across the filter medium would prevent the formation of any cake, and a constant flux would be achieved. In practice this does not occur. A cake is built up on the medium surface, and the filtrate flux declines with time. Some researchers have reported that the filtrate flux reaches an equilibrium value, and only minor flux decline is observed after this time. This is the case when the particles in the feed are much larger than the pores in the filter medium. Other researchers have reported that the filtrate flux declines eventually to zero, and there is no evidence of an equilibrium value. This is the case where the particles in the feed suspension are similar in size to the pores in the filter membrane. This latter behaviour is a result of the particles blocking the filter medium.

In the latter example, in most applications the flux decline is too rapid to permit economic operation. A procedure to restore the membrane has to employed to achieve reasonably constant average fluxes. Many procedures have been proposed and these are described later in this thesis. Principal among these is flow reversal, or backflushing as it is more commonly known. A high proportion of the experimental work reported in this thesis is concerned with this technique.

The principal advantages of crossflow filtration over other filtration methods are as follows.

1. Crossflow filtration offers the possiblity of continuous operation, with low labour costs, for a relatively inexpensive capital investment.

2. Crossflow filters are compact in size as opposed to depth filters.

3. Crossflow filtration is suitable for both thickening and clarification, although in any thickening operation the achievable product concentration is low.

4. Crossflow filtration is suitable for the filtration of both dilute and fine suspensions.

5. For the thickening of fine suspensions, the particulate material is easily recoverable, although at a low concentration. Thus crossflow filtration can be used as an initial operation prior to evaporation or other separation methods.

6. Crossflow filters are completely enclosed. They can therefore be used for material that is easily contaminated. This applies especially to the food industry.

Crossflow filtration also has its disadvantages. Pumping costs are high, especially as not all the fluid is filtered. The major disadvantage however is the embryonic nature of the technology, and the lack of suitable filter media.

INJECTION WATER FILTRATION.

The process that is used for the extraction of oil is described in the Literature Survey. Frequently water is forced down the well to increase the pressure in the reservoir, and thus increase the amount of oil recovered. This water is known as injection water.

This injection water has to be treated, and one of the principle forms of treatment is filtration. On land based rigs this filtration does not present a major difficulty. Depth filtration is quite adequate, and the cost of this treatment is relatively low. The quality of filtrate produced by depth filtration is good, and is quite sufficient for the application.

The development of oil fields offshore created new economic conditions. The costs structure was radically altered and previously economic processes became quite uneconomic under this new structure. A particular example was injection water filtration. The major cost of a item of equipment on an oil rig, is not the capital cost of the equipment alone, but is related to the mass of the equipment, and the space it occupies. If a piece of equipment is large, it means that the offshore platform must be that much larger to accomodate. If a piece of equipment is heavy, the platform must be that much more sturdy to withstand the increased load. On a platform, space and mass are at a premium.

Depth filters are large and heavy. Thus any method of filtration that is lighter and more compact is more suitable for the filtration of injection water than depth filtration. As a result, the filtration systems used for injection water changed as a result of the development of offshore technology. Depth filters were made smaller, by using more suitable media (for example, diatomaceous earth, as opposed to sand), and cartridge filters were introduced. These filters are more compact than the earlier filtration systems, but there are still tremendous savings to be made if still more compact filters can be developed.

There are other requirements of the application. Principle amongst these is reliability. If the filtration system becomes inoperative, the cost in lost production is extremely high. The filters have to withstand severe operating conditions, and the feed conditions will vary with time and geographic location. Crossflow filtration seems to offer the possibilty of meeting these requirements. A high flux crossflow filter will be compact and light, and should be able to compete favourably with the filtration presently used for this application.

For the reasons detailed above, it was decided that it was worth investigating the possibilty of using crossflow filtration, for the treatment of injection water, particularly for offshore oil fields. This thesis examines crossflow filtration, and the design of a prototype crossflow filter that can be used for field trials is presented. Although the research was particularly dedicated to this one application, the results reported could be applied to other applications. Previous research in crossflow filtration, and information that might be useful in gaining a theoreticall understanding of the process are given in the Literature Survey.

LITERATURE SURVEY

INTRODUCTION

Crossflow filtration is a recent innovation in separation technology. For this reason, there is little literature devoted to the subject. It is desirable therefore, to consult literature peripheral to this topic to achieve an understanding of the intricacies of the process.

This survey has been divided into four sections. The sections deal with different aspects of this thesis, and treating the literature in this way will help to make the survey more coherent.

OIL PRODUCTION LITERATURE.

The chief objective of this thesis is the design of a crossflow filter for the filtration of injection water as used in oil production. An understanding of oil production processes will give an indication of the importance of injection water filtration.

Petroleum production literature is voluminous, and it is outside the scope of this thesis to provide an exhaustive survey of even a very particular aspect of this literature. For example, the 1980 Engineering Index (1) abstracted over 70 papers under the keywords "Oil Well Production - Water Flooding". The major part of this particular section is simply a precis of a chapter from "Our Industry" by British Petroleum (2).

Contrary to popular belief, oil is not found in large underground "lakes". (The technical term reservoir is probably the origin of this misconception.) Oil is found within the pore structure of porous rock. This oil-bearing rock is then "capped" by a layer of impermeable rock.

Recovery of the oil is concerned with extracting as much of this oil as is possible, from the pores of the reservoir. This recovery has been divided into three classes, depending on the mechanism that is employed to extract the oil.

Primary recovery is extraction by the naturally occurring forces within the reservoir. There are three principle mechanisms that may occur. Firstly, water drive, which is found when water is sealed in with the oil, under the cap. The water is under very high pressure and thus compressed. When the oil is tapped the compression of the water forces the oil out of the rock and into the production well. Secondly, there is solution gas drive which generally is of lesser importance. This mechanism is caused by gas (which is often found with oil in reservoirs) coming out of solution when the pressure in the reservoir drops. The gas expands, and the oil is forced from the porous rock. The third mechanism is called gas cap drive. This is similar to solution gas drive, but the gas, prior to the commencement of extraction, is not in solution but exists as a discrete fluid above the oil. Where it exists, water drive is usually the most powerful production mechanism.

When the pressure within the reservoir becomes insufficient to force the oil out of the rock, it is necessary, if further oil is to be extracted, to resort to "Enhanced Oil Recovery " (EOR). Enhanced oil recovery has been classified into two types; primary and secondary recovery. The distinction between the two is not very clear: Collins (3) suggests that secondary recovery is the first non-primary recovery that is employed in a field, and the tertiary recovery includes all other non-primary extraction processes.

The most common secondary recovery that is used is water flooding. Water flooding, as the name implies, is the process of injecting water into the reservoir at various circumferential sites. The injected water raises the pressure within the reservoir and enables further oil to be recovered. Collins (3) lists nine other possible secondary recovery schemes, and eight of these involve the injection of water, usually with some additives.

Although it is difficult to give an accurate quantitative estimate of the extra oil recovered (the actual values vary widely from field to field), "Our Industry" (2) suggests that primary recovery can be as low as 20%. Water flooding can increase this to 40%. Although primary and secondary recovery are normally considered as sequential processes, it is now usual to operate these recovery mechanisms simultaneously. Primary and secondary recovery are then combined to produce a steady production of oil.

Tertiary recovery does not rely on raising the pressure in the reservoir, but either on raising the viscosity of the injection water, or on lowering the viscosity (or surface tension) of the oil. There are three main classes of tertiary recovery: thermal, chemical, and micro-biological. Thermal recovery is the use of either hot water or steam as the injection fluid, which assists in lowering the viscosity of the oil. Chemical recovery is the use of various chemical additives in the injection water which assist in oil recovery (for example, surfactants and viscosifiers). Micro-biological recovery is a highly speculative recovery method in which bacteria are grown in the oil well. The bacteria produce various chemicals which assist in oil extraction. Tertiary recovery is not frequently used because the cost of it tends to be greater than the value of extra oil extracted. A great deal of research effort is being devoted to try and make tertiary recovery more cost effective.

It is essential that the water used for injection into the oil reservoir is properly treated. Collins (3) describes oil reservoirs as "depth filters, the best and the most expensive." If poor quality water is used, the rock formation can be irreparably damaged, causing a major reduction in total oil recovery.

Injection water can be obtained from a number of sources: municipal water supply, rivers, lakes, artesian, and in particular from the ocean. Each source has different treatment priorities; indeed municipal water might not require any treatment. Several aspects of injection water quality must be considered. These include suspended solids, salt concentrations, soluble gases, amd bacterial activity. For example, if large quantities of a corrosive gas are dissolved in the water, an increase in temperature downstream may force the gas out of solution and cause severe corrosion. Deposits of corroded material could then find their way into the injection water after the filtration process, and cause irreparable formation damage to the reservoir.

This thesis is concerned with the filtration of suspended solids, primarily from seawater. Chemical and bacterial treatment is outside the scope of this work. Spencer and Harding (4) mention several methods of filtration employed for injection water used in oil production. They include sand and gravel gravity filters, pressure filters and rapid sand filters. These methods were satisfactory until the development of off-shore fields, for they were economical and gave a good quality filtrate. Their chief disadvantage is size; they tend to be rather bulky and heavy pieces of plant.

The development of offshore technology had made these bulky filters very expensive; the chief cost of filtration is the platform cost (the cost of the platform space required to site the filters). Cartridge type membrane filters and high flux and high dirt-holding capacity deep bed filters have become the more economic type of filtration.

MEMBRANE PROCESSES LITERATURE.

Crossflow filtration is primarily a membrane process. The literature devoted to membrane processes generally has information relevant to crossflow filtration. A review of the membrane processes literature is given here, and features salient to crossflow filtration are emphasised.

There are three basic membrane processes of interest: Reverse Osmosis, Ultrafiltration, and crossflow or microfiltration. Dialysis, a fourth membrane process is not relevant here, because the driving force is the osmotic pressure rather than an applied pressure. Crossflow filtration literature will be treated separately, so this section is concerned with reverse osmosis and ultrafiltration literature.

The distinction between reverse osmosis, ultrafiltration, and filtration is not clear. Kirk-Othmer (5) suggests the following classification according to the size of the solute being separated.

PROCESS	SOLUTE SIZE
Reverse Osmosis	< 1nm
Ultrafiltration	1nm -100nm
Filtration	> 100nm

Scott (6) suggests the following classification.

PROCESS	SOLUTE SIZE
Reverse Osmosis	< 2nm
Ultrafiltration	2nm - 1 micron
Filtration	> 1 micron

Perry (7) suggests reverse osmosis is applicable when the solute and solvent sizes are similar, and that ultrafiltration applies when the solute size is ten times greater than the solvent.

In reverse osmosis the osmotic pressures are high (typically 3.4 MPa or higher - (8)). Hydraulic pressure must exceed this to enable separation to occur, and applied pressures of 6.8 - 10 MPa are usual. In ultrafiltration the osmotic pressures are much lower, and hydraulic pressures of 10 to 100 kPa are employed commercially (6). In filtration, the osmotic pressures are negligible, and applied pressures are usually lower than for ultrafiltration. Scott (6) suggests a pressure of 100 kPa as being normal.

Both reverse osmosis and ultrafiltration have been used commercially for several applications. Harrison (9) suggests that the most likely applications for these membrane processes are in the food industry, primarily because of the sterile and delicate nature of the required product. Ten years ago membrane processes were regarded as a promising form of new technology that was about to "take-off". Today the situation is similar; they are still promising, but their commercial employment is low. The major reason why these membrane processes are not used more widely is that the normal filtration rates (fluxes) obtained are very low. If the filtration fluxes could be markedly improved, the process economics would become a lot more favourable.

One reason why the fluxes are so low is a phenomenon known as Concentration Polarization. This is a well observed and well documented phenomenon and many authors have described it (10 - 15). It arises because of mass transfer considerations across the membrane. At the membrane surface, the solvent is passed through the membrane and the solute is not. This forces an increase in the concentration of solute at the surface. Thus a concentration gradient is developed near the membrane surface, with the solute concentration at the surface significantly higher than that in the mainstream. This is known as concentration polarization.

In reverse osmosis, the rate of mass transfer is proportional to the effective pressure difference across the membrane. The effective pressure difference is the difference between the applied hydraulic pressure, and the osmotic pressure of the solution. As the concentration of solute rises, so does its osmotic pressure, and consequently the rate of mass transfer decreases with concentration polarization. In ultrafiltration, this effect is further compounded by a feature of the common solutes used in this separation method. The viscosity of the fluid often increases dramatically with increasing solute concentration (sugar is a typical example which is often separated by ultrafiltration). This viscosity effect further reduces the filtrate flux.

It will be shown in the next section that a similar effect to concentration polarization may occur in crossflow filtration. Concentration polarization is a deleterious feature, and many ways of alleviating or minimising it have been proposed. The most common is the use of a high axial velocity to create a high shear stress at the membrane surface.

Sheppard and Thomas (12,13) have described the use of a high axial velocity to maximise flux. Flux decline was a lot less rapid with a high velocity than without, and filtration fluxes generally were maintained at a higher level. They determined that the concentration of solute at the membrane surface declined with increasing velocity.

Other means of reducing concentration polarization have been proposed. Lowe and Durkee (15) achieved a lesser degree of concentration polarization by using numerous Latex spheres in the feed. As the feed flowed around these small free-moving spheres, turbulence was promoted close to the membrane, and the concentration of solute at the surface decreased. Ultrasonics have also been used (16,17), and increases in filtrate fluxes were achieved. The use of reverse osmosis in wastewater renovation is well documented (18-23). Various applied pressure and fluxes have been reported. All reports recommend pretreatment of the feed, and especially filtration. This is to prevent a rapid flux decline during operation. There is surprising agreement as to the fluxes that can be achieved in this process. Loeb and Manjikan (29) report fluxes 0.8 - 2.4 m/d, at 4 MPa, while Kuiper et al (23) achieved fluxes of 1 - 2 m/d also at 4 MPa. Sheppard and Thomas (12) record a flux of 1.2 m/d at a pressure of 5.3 MPa.

The major difficulty reported in these studies was not the low initial fluxes, but the rapid decline in flux as the membrane aged. Originally this was attributed to compaction of the membrane under the high pressure to which it was subjected. It is now believed that the major cause of this flux decline is not compaction, but is due to fouling of the membrane. In wastewater renovation literature, the principle cause of fouling is precipitation of dissolved salts in the membrane surface.

Carter and Hoyland (24) have reported that the rate of deposition of a fouling layer is independent of the concentration of the foulant, and also independent of the Reynolds number of the tangential flow. However they found that the equilibrium thickness of the fouling layer is dependent on the Reynolds number; at a high crossflow velocity the layer is thinner. Sheppard and Thomas (12) reported that the rate of flux decline with time was very dependent on the axial velocity. They found that at a velocity of 8 m/s there was only a 10% decrease in flux over a 10 day period. There was an 80% reduction in flux over the following 2 days, when the velocity was reduced to 0.5 m/s.

Sheppard and Thomas also reported experiments with rough and smooth membrane supports. The rough support gave no evidence of a greater degree of turbulence promotion (as measured by the effect of concentration polarization) but was noticeably more susceptible to fouling.

The cleaning of membranes has been investigated by Belfort (26). A number of methods of in-situ cleaning have been suggested. These include reversal of the direction of flow, air flushing, the use of detergents (both anionic and enzymatic), and foam ball swabbing. One method which Belfort and Marx (25) found to be especially promising is the use of a sacrificial layer. Dynamically formed membranes use a similar principle. Csurny et al (27) and Awokoya and De Cicco (28) have investigated the advantages of dynamically formed liquid membranes. They believe such a method to be promising, because the membrane can be regenerated frequently, and thus flux decline is not a major difficulty.

There is a considerable volume of literature devoted to membrane processes. This is not meant to be an exhaustive review of that literature, but an indication of some of the points raised that might be pertinent to crossflow filtration.

CROSSFLOW FILTRATION LITERATURE.

Crossflow filtration is a new innovation in separation technology. It is not well documented, and the literature that is devoted to it, is scattered through many different publications, making it very difficult to collate it all.

Crossflow filtration has been proposed for many different applications. These include both thickening and clarifying operations. Since the application of the process varies so markedly, data from one particular process need not necessarily be directly applicable to a different application.

Tiller et al (29) mentions crossflow filtration in a paper entitled "Delayed Cake Filtration", even though they dismiss it: "In general, the degree of thickening is limited..." Csurny et al (27), and Awokoya and De Cicco (28) use crossflow filtration for the processing of wastewater in the pulp and paper industry. Knibbs (32-24) wrote three papers dedicated to the use of crossflow filtration of injection water used in oil production. Obviously, these latter papers represent the literature that is most directly applicable to this thesis. Klein (35) and Rushton et al (36) analyse the use of crossflow filtration as a thickening operation. Henry's paper (31) is the most general; his paper is concerned more about the process, as opposed to the application.

Most reports have studied the effect of crossflow velocity. Henry (31) suggested that for laminar flow, the filtrate flux is proportional to the shear rate to the power on n (a normal power-law type relation). Experimentally Henry suggested n was between 0.5 and 1.3, depending on the feed suspension. For turbulent flow Henry reported a similar relation, with an exponent value of between 1 and 1.2 .

Rushton et al (36) suggests an equation for the flux-velocity relation of the form

$$\text{Flux} = a + b*\text{velocity}$$

where a and b are constants of the system and the suspension. Other reports (27,28,33,37) all reported significantly higher fluxes with increased crossflow velocities, and that flux decline was less at these higher velocites. No attempt was made to model these results.

The effect of pressure on filtrate flux is interesting. Henry (31) found that a 100% increase in pressure caused only a 20% increase in filtrate flux. Csurny et al (27) stated "increasing the pressure, increases flux, but frequently follows a decay, sometimes back to or below the flux at the lower pressure". Awokoya and De Cicco also reported a very weak flux-pressure relation. Harrison et al (37) reported that an increase in pressure can even be detrimental to the filtrate flux. Knibbs (34) suggests that the filtrate flux is proportional to the square root of the pressure differential, and presents experimental data that supports this relation.

The effect of feed concentration has not been well reported. Henry (31) suggests that it is not an important variable, but filtrate flux does decline at higher concentrations. Harrison et al (37) suggest that it is the nature of the suspension (for example, whether the particles in the feed are deformable or rigid) rather than the concentration, that has the major influence on filtrate flux.

Filtration fluxes and other process variables reported in the literature vary widely. The following table summarises this information.

AUTHOR	TYPICAL FILTRATE FLUXES (m/d)	TYPICAL PRESSURE DIFFERENCES (kPa)	TYPICAL CROSSFLOW VELOCITIES (m/s)
Csurny et al (27)	6.3-24.5	34-272	3-7
Awokoya & De Cicco (28)	13.8-20.2	340	3-8
Henry (31)	0.5-2.5	68-272	n/a
Knibbs (33)	432-840	340-680	1
Knibbs (34)	216	310-650	0.11
Rushton et al (36)	n/a	34	0.03-0.08
Harrison et al (37)	173-518	68-272	0-8

Table of operating conditions of crossflow filters as described in the literature

Apart from the crossflow, the most common method detailed in the literature for preventing or minimising flux decline is the reversal of flow, or more commomly, backflushing or backwashing. Crossflow filtration, by definition, requires that the membrane be unsupported on the feed side during normal filtration mode. Were the membrane to be supported on the feed side, the crossflow would not help in preventing flux decline. Thus, unless the feed side of the membrane can be supported during the backflush mode only, backflushing requires that the membrane be strong and robust.

Various operating cycles and backflush parameters are reported in the literature. The following table summarises this information.

AUTHOR	BACKFLUSHING FLUID	DURATION OF BACKFLUSH	BACKFLUSH PRESSURE DIFFERENCE	TIME BETWEEN BACKFLUSHES
Rushton et al (36)	Filtrate	60 secs	n/a	10 mins
Knibbs (33)	Feed	60 secs	408 kPa	6 mins
Knibbs (34)	Feed	60 secs	n/a	5.5 mins
Knibbs (34)	Feed	60 secs	210 kPa	8 mins
Harrison et al (37)	Filtrate	10-20 secs	41-136 kPa	n/a
Klein (36)	Filtrate	1-2 secs	50-100 kPa	*

* Klein stresses that the frequency of backflushing is a decision of the operator and will vary markedly from one application to another.

Table of backflushing cycles as detailed in the literature

Knibbs (34) recommends that the membrane be cleaned periodically by circulating and aerating a warm solution of an enzymatic detergent around the filter. Klein (35) recommends washing the filter with solvent that will not damage the membrane. Obviously the membrane and suspension of a particular application will be important factors in the choice of a solvent.

There is no literature that indicates the use of other methods of minimising flux decline for crossflow filtration. In particular, there is no evidence to suggest that air scour, air backflushing, turbulence promotion, or ultrasonics have been used for crossflow filtration systems.

LITERATURE RELATED TO THEORETICAL ASPECTS OF CROSSFLOW FILTRATION.

This section provides a cursory examination of the literature that might be useful in producing a mathematical model of crossflow filtration. In particular it examines some of the literature devoted to the fluid mechanics of flow over a permeable surface.

Hermia (38) gives a resume of blocking filtration. He modifies the theory of cake filtration and presents three models and derives the equations that describe these models. The models are called; a) Complete Blocking Filtration, b) Intermediate Blocking Filtration, and c) Standard Blocking Filtration. One of these models might be applicable to crossflow filtration, or perhaps some combination of these models. A further analysis of this subject is given in Appendix 2.

It has been suggested that a curved membrane might lead to increased filtrate fluxes. There is some literature which examines the nature of the boundary layer along curved surfaces (39-2). Both Meroney and Bradshaw (39) and Shivaprasad (42) found that a convex wall decreased the turbulent intensity of the boundary layer, and this effect was reversed for a concave surface. Meroney and Bradshaw (38) also reported that the shear stress decreased steeply outside the near wall region over a convex surface, whereas for a concave surface the shear stress remains high well beyond the point where it normally diminishes.

Boundary layer flow over permeable surfaces has been studied by several groups and is well reported (43-48). Simpson (43) in his paper, extends Coles "law of the wake" formulation and applies it to flow with transpiration. Ariye (45) reported that "... suction has a substantial effect on the boundary layer markedly increasing the velocity gradient near the wall." He reported that as a result of this, frictional resistance coefficients are greater in the case with suction than without, usually by a factor of approximately 2. He reported this frictional coefficient ratio to be a function of the ratio of mass flowrates (mass permeate flowrate/mass crossflow flowrate).

Rekin (46) presents velocity profiles for fluid flow over surfaces with suction and injection. It is apparent that suction in the region very close to the wall causes a much higher velocity gradient. Injection causes a nearly linear velocity profile (constant velocity gradient). He hypothesises a simplified equation to represent the velocity profile within the boundary layer as follows

$$\ln (1 + B*u)/\ln(1 + B) = y^{-1/n}$$

where B = Injection or Suction Parameter
u = Dimensionless local velocity.
y = Dimensionless local co-ordinate
n = Exponent for power law dependence

Rekin (46) also reports the change in the boundary layer thickness due to suction or injection. Suction decreases the boundary layer thickness, while injection increases it. He derives the following formula to describe the boundary layer thickness change due to suction or injection.

$$\frac{d}{d_o} = \frac{T*(1 + B)}{T_o} \quad \frac{(n+1)(n+2)/n}{(n_o+1)(n_o+2)/n}$$

where d = Boundary layer thickness
T = Shear stress at wall
n = Power law dependance exponent
B = Suction or injection parameter
o = Subscript denoting unblown condition.

Bushmarin (47) gives some parametric equations of boundary layer flow over porous surfaces. The velocity of suction or injection, or the velocity at the edge of the boundary layer does not appear explicitly in the equations. This approach is specifically aimed at the situation where the suction, or injection, varies with time.

Aleksin et al (48) present a numerical solution to boundary layer flow over a porous plate. This method is suitable for when the boundary conditions vary rapidly with time.

Trajectory calculations for particles in a moving fluid may well be very valuable in any modelling of crossflow filtration. The literature for this approach is both voluminous and well known. It is not presented here, but many standard textbooks will be able to assist with information concerning these calculations.

EXPERIMENTAL EQUIPMENT

Three different experimental rigs were used to obtain the results reported in this thesis. Each of the rigs was designed and constructed principally for a particular set of measurements, and when new experiments were needed, the rig often had to be modified, or even superseded by a new rig. Each of the three rigs is described below, and for convenience they will be referred to as Rigs A,B, and c. This nomenclature is used throughout the thesis.

RIG A.

This rig was used for the laser anemometer measurements. It was constructed entirely of Perspex, a transparent polymeric material. Although reasonably suitable for laser anemometry, the rig was found to be unsuitable for all the other experiments.This was for two reasons. For the photographic experiments, the optical quality of the perspex was insufficient, and the heat generated by the intense light source could also cause the rig to distort. It was very difficult indeed to change the membrane in Rig A, and this made the rig unsuitable for any of the flux-time measurements. For these reasons Rig B was constructed, and Rig A was only used for the laser anamometer measurements. A flow diagram for the operation of Rig A is given in Fig 4.1 and drawings of the rig, including the critical dimensions are given in Fig 4.2. The pump was a rotary screw type "MONO" pump, and the pressure gauge was a 50 mm dial type.

RIG B.

This rig was constructed primarily for the high speed photography meaurements. For this reason, glass sides were used on the channel to improve the optical quality. The rig was also used extensively for the backflushing experiments, and for these tests, the glass was replaced by Perspex. It

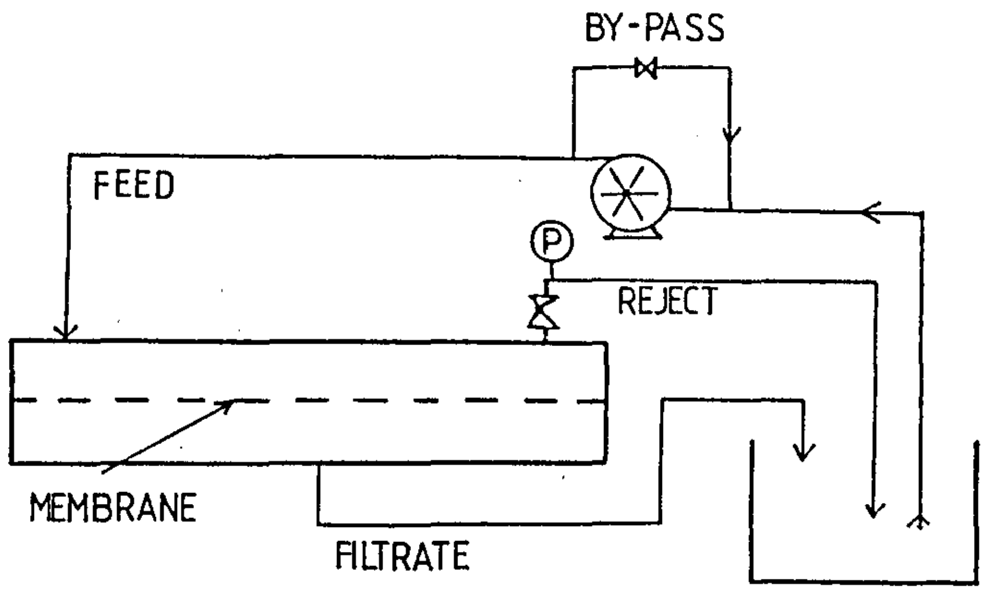


Fig. 4.1 Flow Diagram of Rig A.

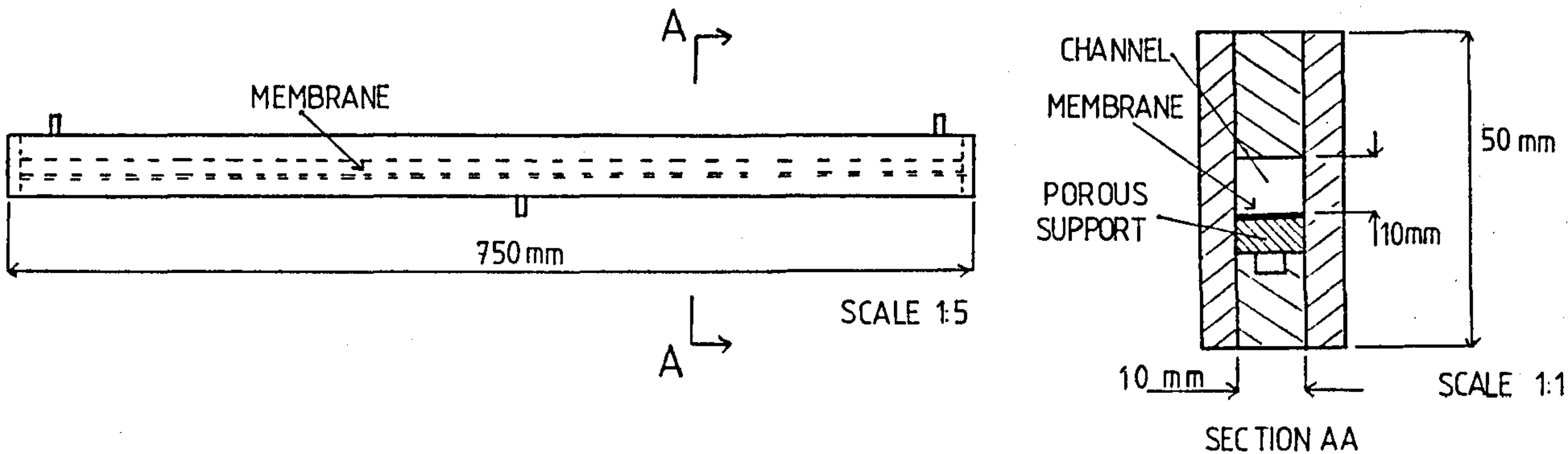


Fig.4.2 Diagram of Rig A with
Cross-section

was relatively simple, if tedious, to change the membrane in this rig, and this made it suitable for the flux-time measurements.

Three different flow arrangements were used for the experiments using this rig. For the photography experiments, and the flux-time measurements without backflushing, the circuit was as in Fig 4.3A. For the experiments which involved filtrate backflushing, a second pump was introduced and the applicable flow diagram is as in Fig 4.3B. When backflushing with compressed air was used, the circuit was modified again, and is represented in Fig 4.3C. Detailed drawings of Rig B are given in Figs 4.4A, 4.4B, and 4.4C.

For all these circuits, the principal pump was a 2.5 hp (1.88 kW) "MONO" pump. The additional pump which was required for filtrate backflushing, was a smaller "MONO" pump. The feed pump was mounted on a platform approximately 2 metres above the rig. It is advantageous to keep the pressure on the filtrate side of the membrane as close to atmospheric pressure as possible, but it is also desirable to keep the feed concentration constant. To minimise the filtrate pressure, the filtrate was collected in a small container at same level as the rig. When a small volume of filtrate had been collected, this was directed back into the main feed tank. There is a minimal change in concentration because of this, because the filtrate to feed volume ratio is kept small. This technique was used for all but the laser anamometer measurements.

The pressure gauges used are either 100 mm or 50 mm dial type gauges. All valves were gate valves, except the valve which initiated the actual backflush, which was a quick action cock type valve. For the air backflushes, compressed air from the University's main compressor was used, with a reducing valve to give the required pressure. The maximum pressure available was 544 kPa.

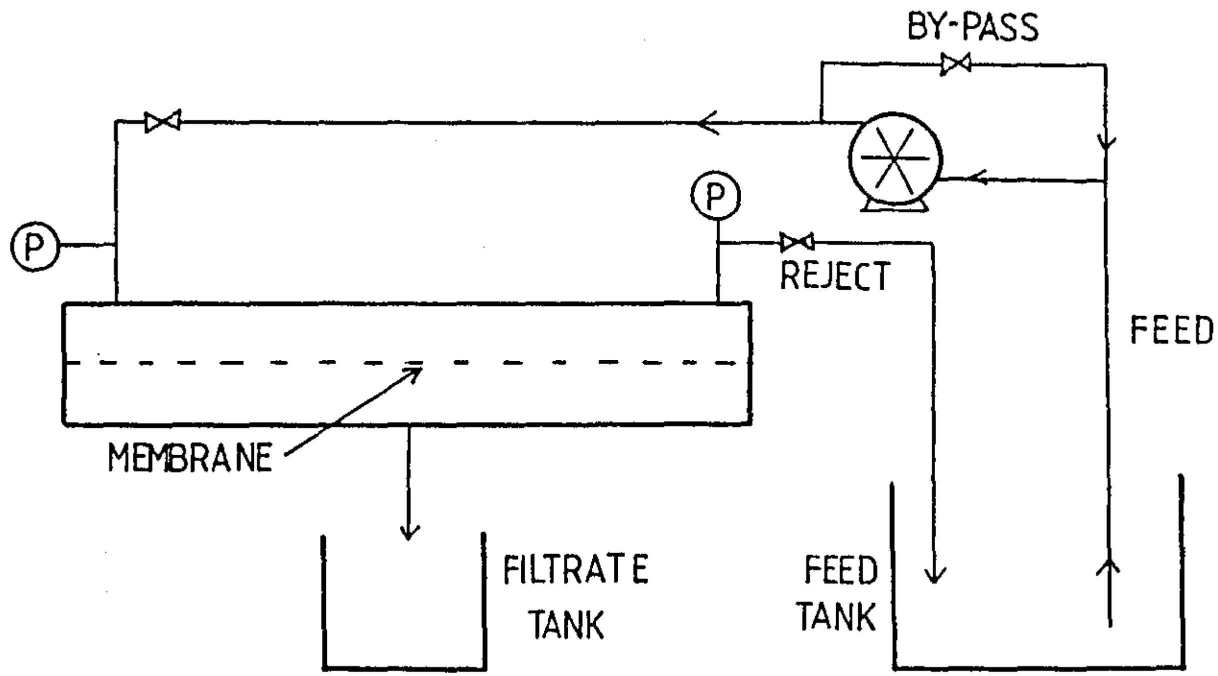


Fig. 4.3A Flow Diagram for Rig B. (No backflushing)

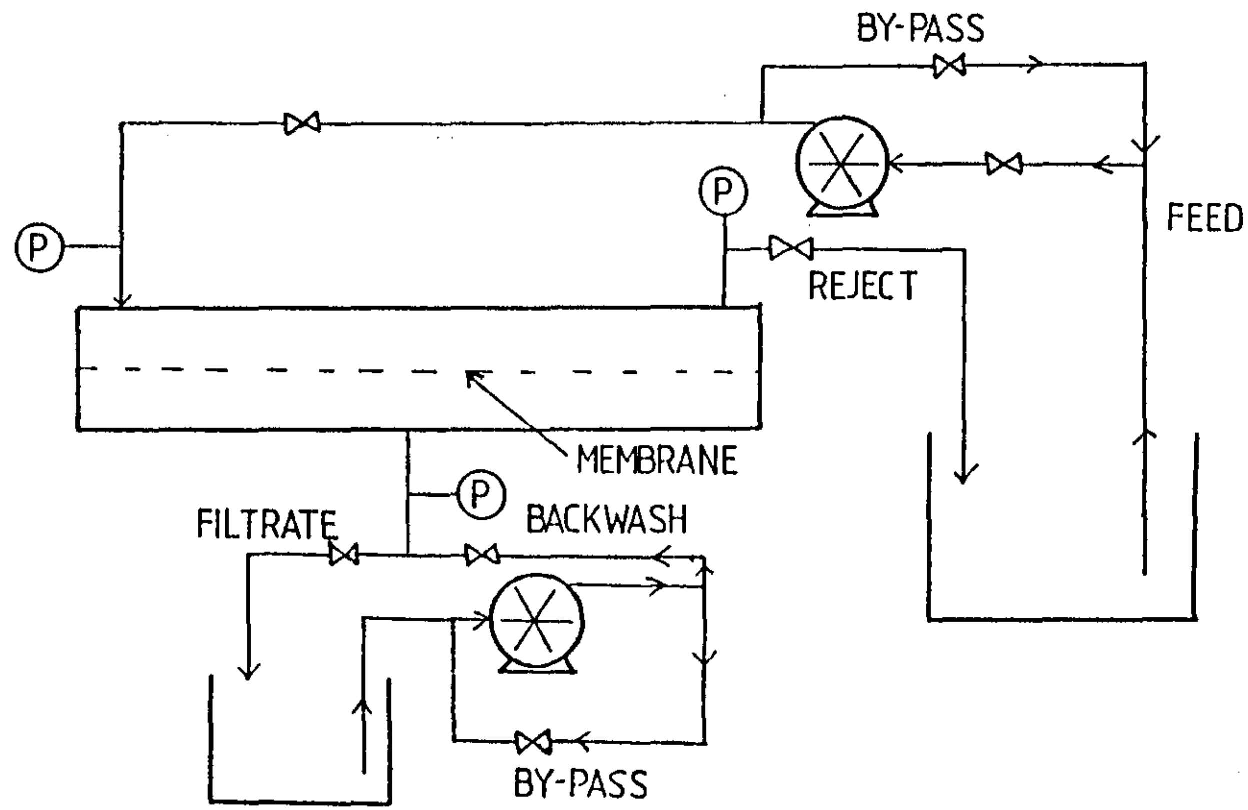


Fig. 4.3B Flow diagram for Rig B. (Filtrate backflushing)

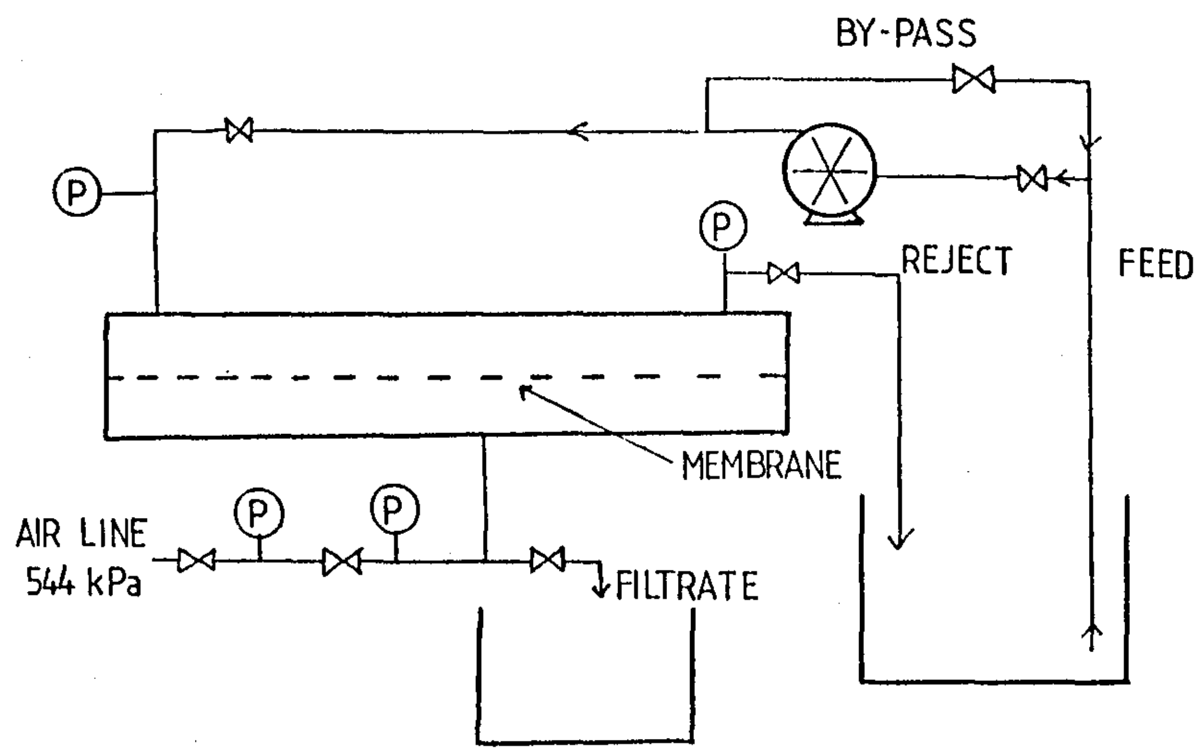


Fig.4.3C. Flow diagram for Rig C. (Air backflushing)

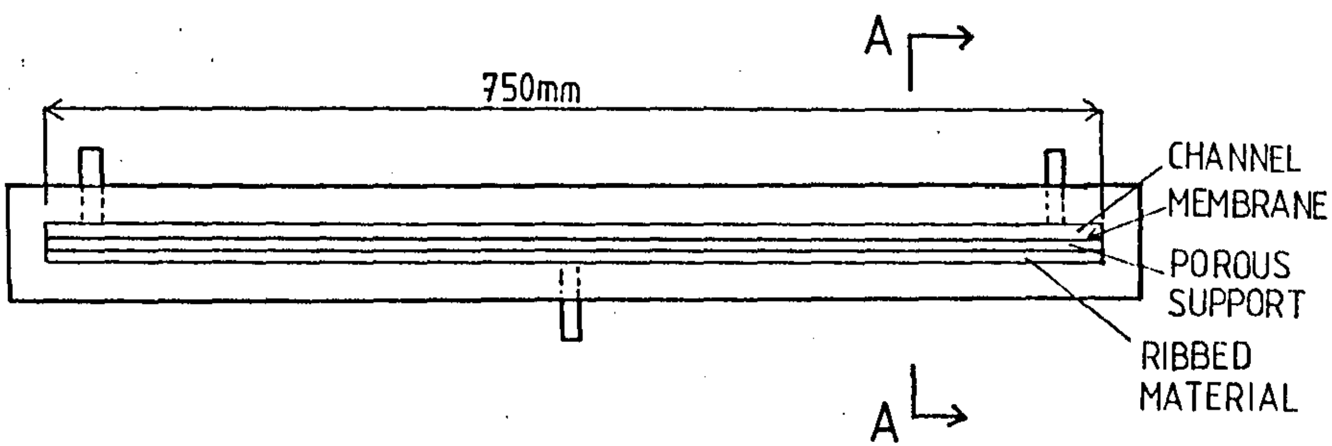


Fig.4.4A Diagram of Rig B.

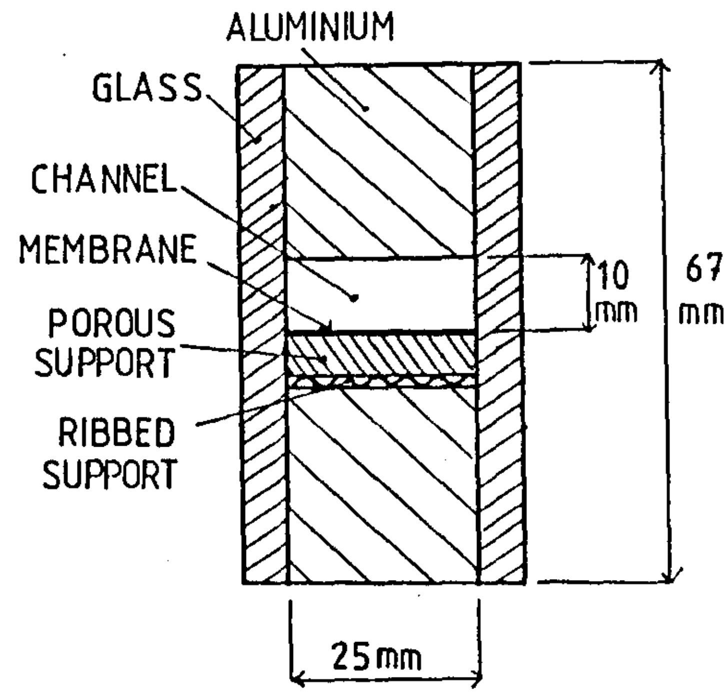


Fig.4.4B Cross-section of
Rig B.

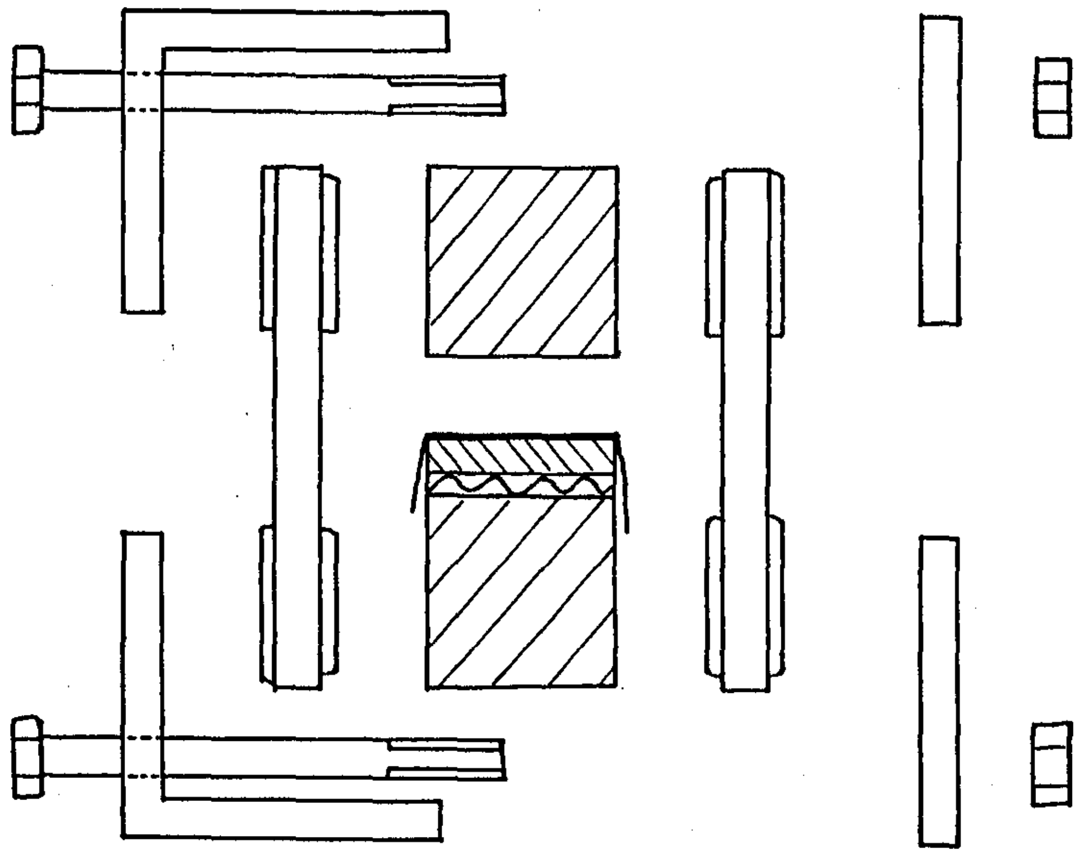


Fig 4.4C Assembly diagram of Rig B.

RIG C.

Rig C was constructed for two reasons. Firstly, it was decided to test the effect of different channel geometries. This required a multi-channel rig with channels of different sizes and shapes. Secondly there was some difficulty in keeping Rig B as clean as is required for filtrate quality experiments.

Rig C was constructed entirly of aluminium, and had five channels of varying shapes and sizes. These channels were kept entirely separate, and only one channel was used at a time. The operating circuit was similar to that used for Rig B, and a flow diagram is shown in Fig 4.5. Drawings of the rig are given in Fig 4.6A, and 4.6B. The channel geometries referred to are given in the table below.

CHANNEL No.	SHAPE	HEIGHT	MEMBRANE WIDTH.
1.	Triangular (60° * 75° * 45°)	5 mm	6.3 mm
2.	Triangular (90° * 45° * 45°)	5 mm	10 mm
3.	Triangular (90° * 45° * 45°)	7.5 mm	15 mm
4	Square	5 mm	5 mm
5	Rectangular	5 mm	10 mm

MEASUREMENTS.

All pressures were determined using pressure gauges of a suitable range. Times were measured either by a stopwatch, or a digital wrist watch. Fluxes were measured indirectly by determination of the flowrate, followed by division by the applicable area. The flowrates were measured by either rotameters, or the "bucket and stopwatch technique". The latter involves collecting a predetermined amount of fluid in a graduated cylinder, and recording the

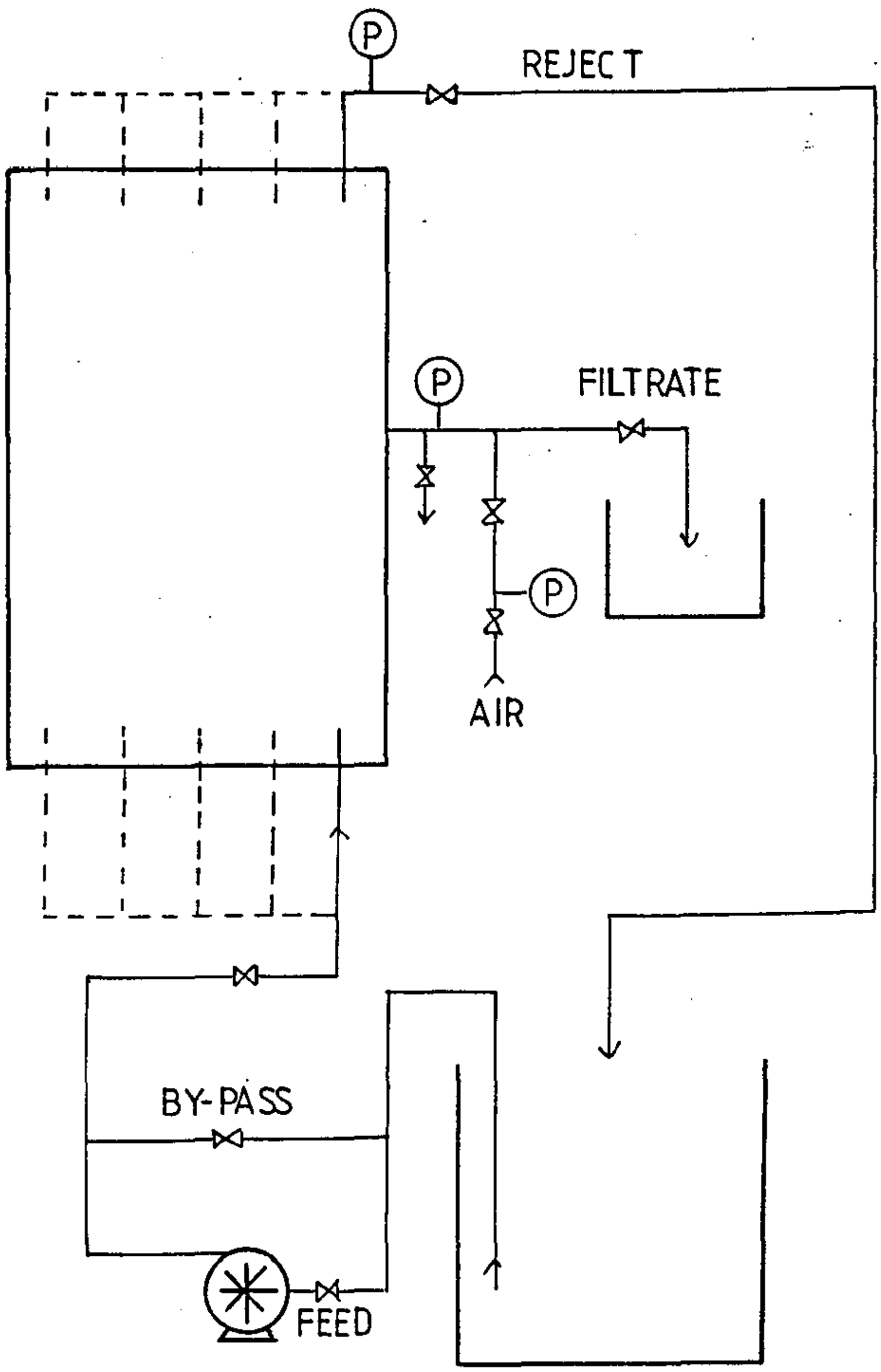


Fig. 4.5 Flow diagram for Rig C.

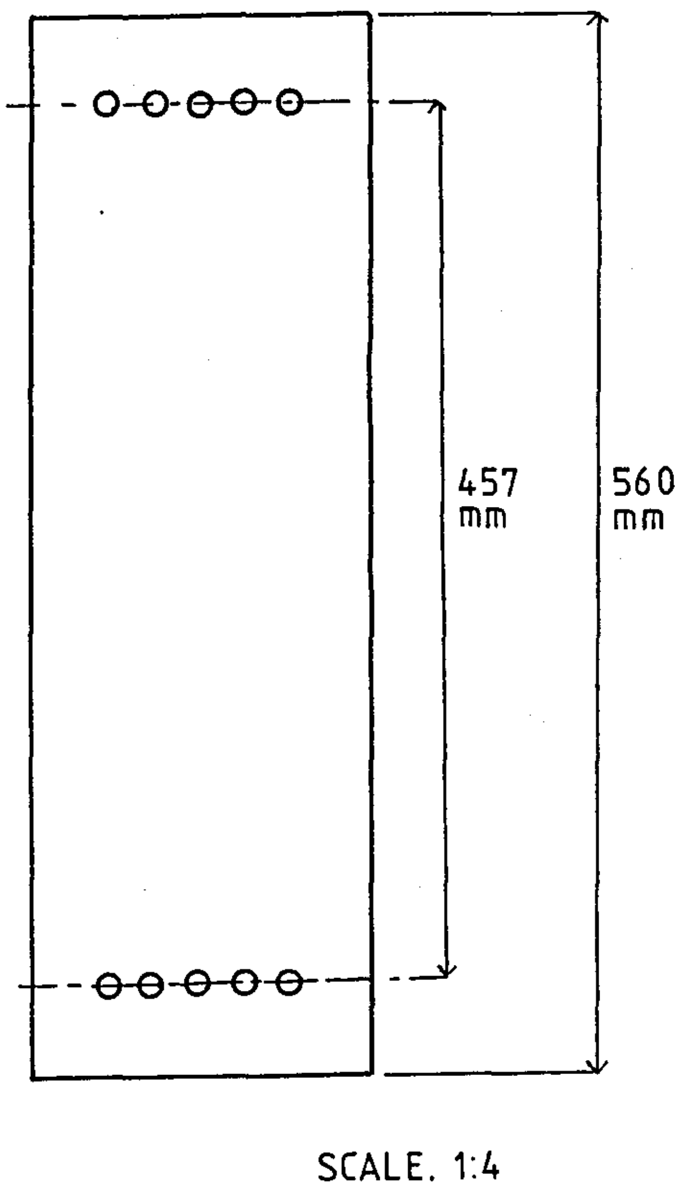


Fig. 4.6A Diagram of Rig C.

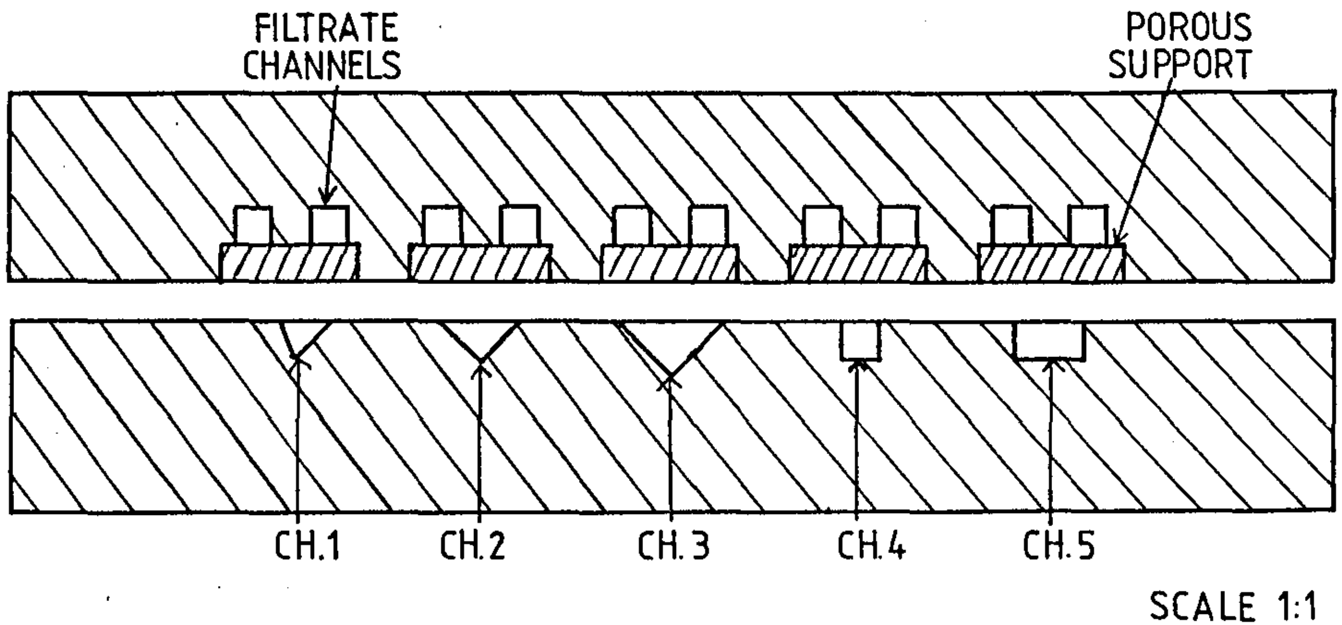


Fig. 4.6B Cross-section of Rig C.

time taken. This method is very accurate, but time-averaged, rather than instantaneous, values are measured. It is also a rather tedious method of flow measurement. Rotameters are less accurate, but are easy to read, and measure a reasonably instantaneous value. Crossflow velocities were also measured in the same way.

BACKFLUSHING EXPERIMENTS

INTRODUCTION

When a membrane filter is operated at a constant pressure, the filtrate flux declines with time. A typical example is given in Graph 5.1. After a relatively short period of time , the flux is so low as to be impractical. It is then necessary to revive the membrane, so as to restore the filtrate flux to a practical level.

There are several ways of reviving a membrane. Replacement of the membrane is possible, though it is seldom likely to be practical. The membrane can be cleaned, either mechanically or chemically. A particular type of mechanical cleaning is Flow Reversal, more commonly known as backwashing or backflushing.

Backflushing is the process where a fluid is forced through the membrane in the opposite direction to the filtration. The driving force for this process is a pressure difference which is usually created by increasing the pressure on the filtrate side of the membrane. The velocity of the fluid dislodges some of the material that is blocking the membrane, and the filtrate flux is increased. Obviously backflushing is essentially a batch operation: it is not possible to filter and backflush over the same section of membrane simultaneously.

There are three main variables associated with flow reversal; a) the backflushing fluid, b) the pressure differential across the membrane, and c) the duration of the backflush. The experiments described in this chapter were performed to determine the effect of these variables on membrane restoration.

There is an additional method of cleaning which is similar to backflushing involving the use of purged air bubbles. The air forms slugs that scour the membrane. At the end of each slug is a region of low pressure, and it is this vacuum that creates the cleaning effect. While this method is not strictly backflushing, it is very closely aligned to it, and for this reason will be considered in this chapter.

EXPERIMENTAL PROCEDURE

Two fluids were used for backflushing, filtered water and compressed air. Various combinations of pressure and backflush duration were used in the experiments.

The procedure for initiating a filtrate backwash is as follows. First, the restrictor at the end of the crossflow channel is opened. This minimises the pressure differential across the membrane which must be overcome to cause a backflush, while maintaining the same crossflow velocity. The valves around the backwash pump are then set in the appropriate positions. To initiate the actual backwash, the backwash pump is switched on, the pump's by-pass valve having been preset to deliver the required pressure. The pump is turned off after the backflush duration has elapsed and the previous mentioned steps reversed to return to the normal filtration mode.

The procedure to initiate an air backflush is slightly different. First the restrictor at the end of the crossflow channel is opened. Then the valve on the line to the filtrate tank is closed. The airline is throttled so as to deliver the required pressure. The actual backflush is initiated by opening a quick acting cock on the airline. This cock is closed to end the backflush, and the previous steps are reversed to return to normal filtration mode.

The procedure for the purged air can be used both during backflushing and also during normal filtration. The air was introduced into the feed line before the filtration cell, and a quick operating cock allowed the air to be introduced in spurts. The rig was mounted vertically for these experiments, so that the air could travel up the channel in continuous slugs. Two alternative strategies were used. Firstly, the air was introduced in very short bursts every 10 seconds in filtration mode. An alternative strategy was to introduce the air for a short period during backwashes. Data for both these strategies are given in

Graphs 5.3 & 5.4.

All these experiments were performed on Rig B. The fluxes were determined by measuring the filtrate flowrate, and then dividing by the membrane area. The filtrate flowrate was measured either by a measuring cylinder and stopwatch technique, or by a rotameter. The backflush pressure is the pressure differential across the membrane during the backflush. This was measured by the difference between two pressure gauges, one in the filtrate line and the other in the crossflow channel. The duration of the backflush refers to the time that the above pressure was applied.

RESULTS

All the results are presented in filtrate flux v time graphs. A description of each graph is given below. For the graphs the following convention is used. A vertical line on the graph represents a discontinuity of the previous filtration mode. This can be a backflush (either filtrate or compressed air), the commencement or termination of air purging, or the stopping and restarting of the pump for any reason. Usually the details of the discontinuity are given to the right of this line. Where only one type of discontinuity is shown on the graph the details of it are given in the graph's caption. "BW" on a graph designates a backflush (either filtrate or air), "AIR" designates a region where air purging is in operation and "NO AIR" a region where the air purging is inoperative.

GRAPH 5.1 This graph shows a typical example of how the filtrate flux decays with time. The membrane was Versapor 0.8 microns, with a pressure difference of 136 kPa and a crossflow velocity of 0.76 m/s.

GRAPH 5.2 This graph includes data for several filtrate backflushes of differing pressures and durations. The run lasted for over six hours and during that time the filtrate flux declined steadily. The membrane was Versapor

0.8 microns with a pressure difference during filtration of 136 kPa and a crossflow velocity of 0.79 m/s.

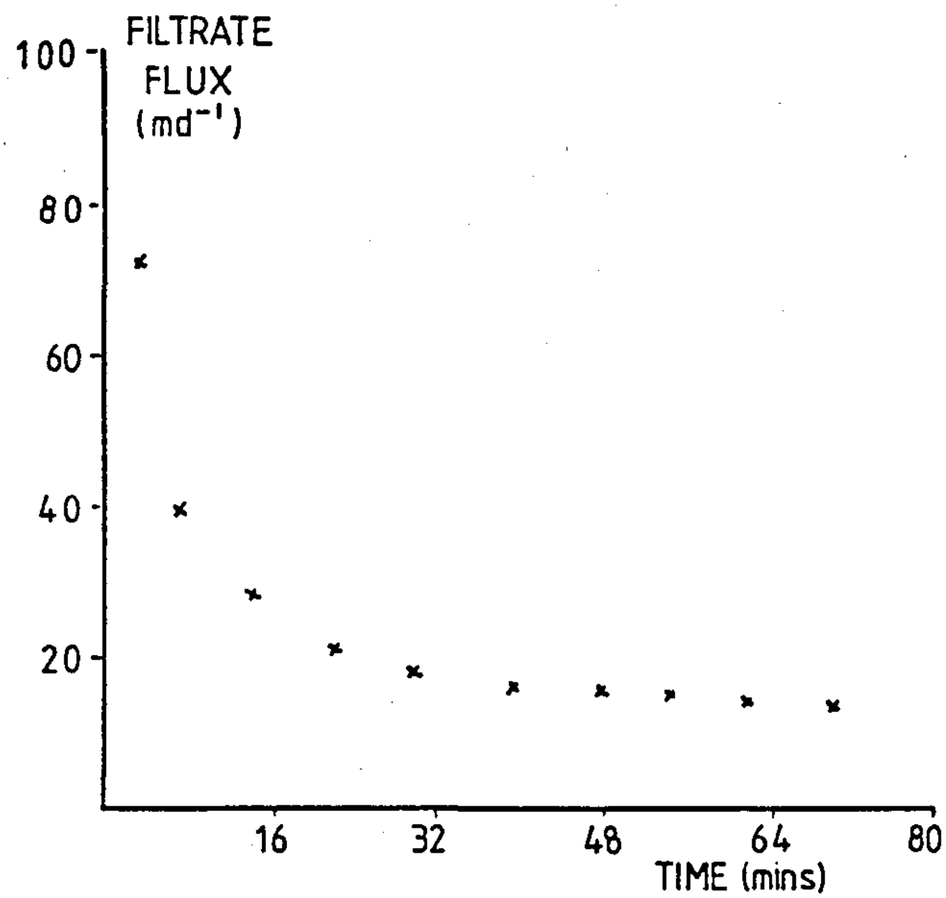
GRAPH 5.3 This graph demonstrates the influence that the introduction of air slugs during filtration has on the filtrate flux. The graph has five regions: two periods where air purging takes place, in between three periods where normal filtration occurs. No backflushes are included on this graph. The air slugs were introduced into the feed line at a pressure of 544 kPa every 30 seconds for a period of approximately 1 second. An increase in the pressure in the channel, of approximately 34 kPa, was noted when the air was introduced. This increase took approximately 10 seconds to decay. The membrane in this experiment was Versapor 0.8 microns, with a filtration pressure of 136 kPa and a crossflow velocity of 1.3 m/s.

GRAPH 5.4 This graph examines the effect of an air slug during a filtrate backflush. Eight backflushes are shown with every alternate backflush having a 1 second pulse of air introduced during the backflush. The air pulse was introduced into the feed line prior to the filtrate channel at a pressure of 544 kPa. All the backflushes gave an effective pressure difference across the membrane of 41 kPa and had a duration of 10 seconds. The pressure difference during filtration was 136 kPa. The membrane was Versapor 0.8 microns.

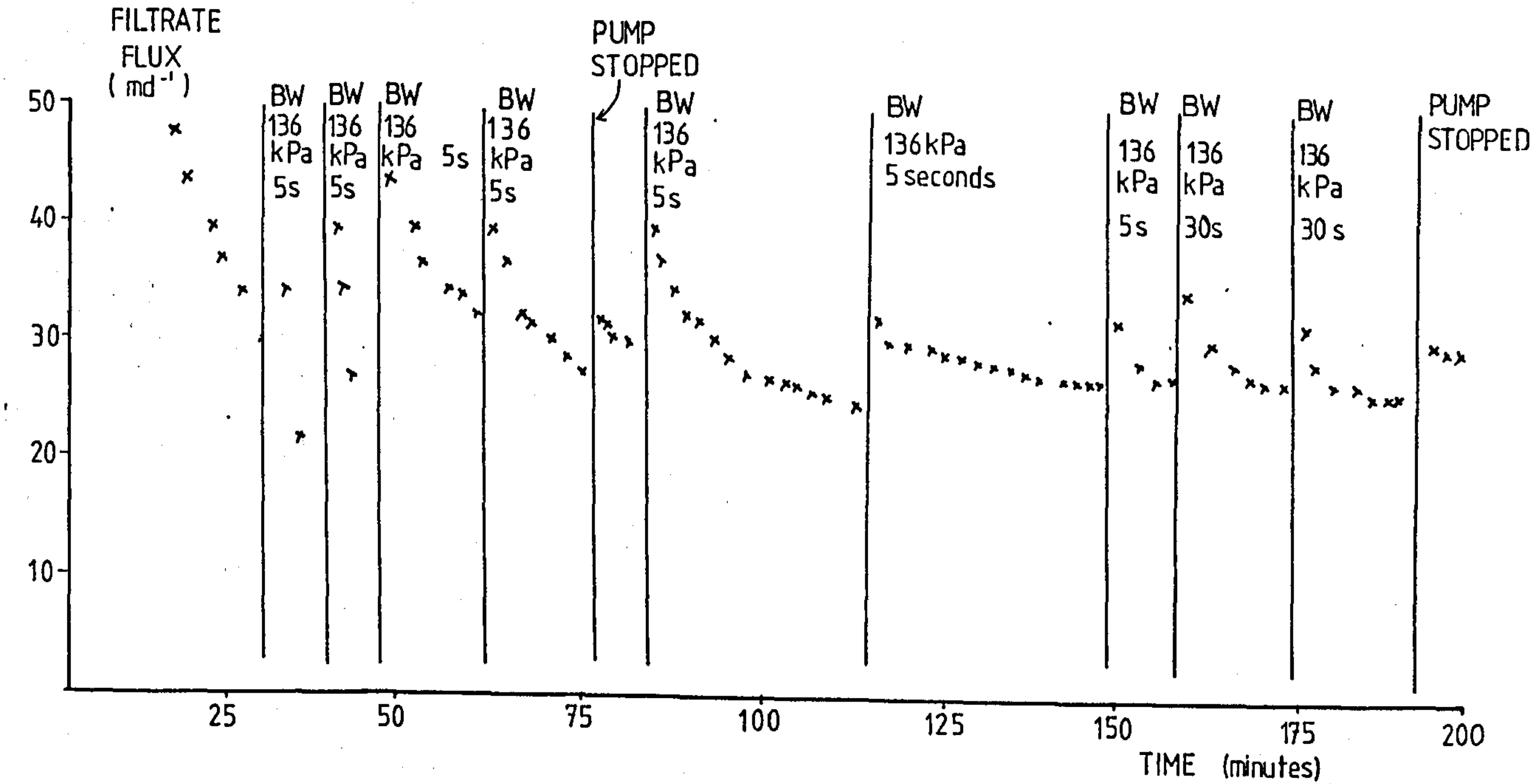
GRAPH 5.5 This graph examines backflushing with compressed air. Various backflush pressures and durations were used and the run lasted for 75 minutes. For the final 45 minutes of the run backflushes of a 1 second duration and 272 kPa pressure difference were found to give a flux that did not decline over this time. The filtration pressure was 136 kPa and crossflow velocity was 0.88 m/s and the membrane was Versapor 0.8 microns..

GRAPH 5.6 This run is similar to the previous one. Backflushes of 1 second duration and 272 kPa effective pressure difference gave a constant flux for the entire run of 75 minutes. The membrane was Versapor 1.2 microns with a filtration pressure of 136 kPa and a crossflow velocity of 1.2 m/s.

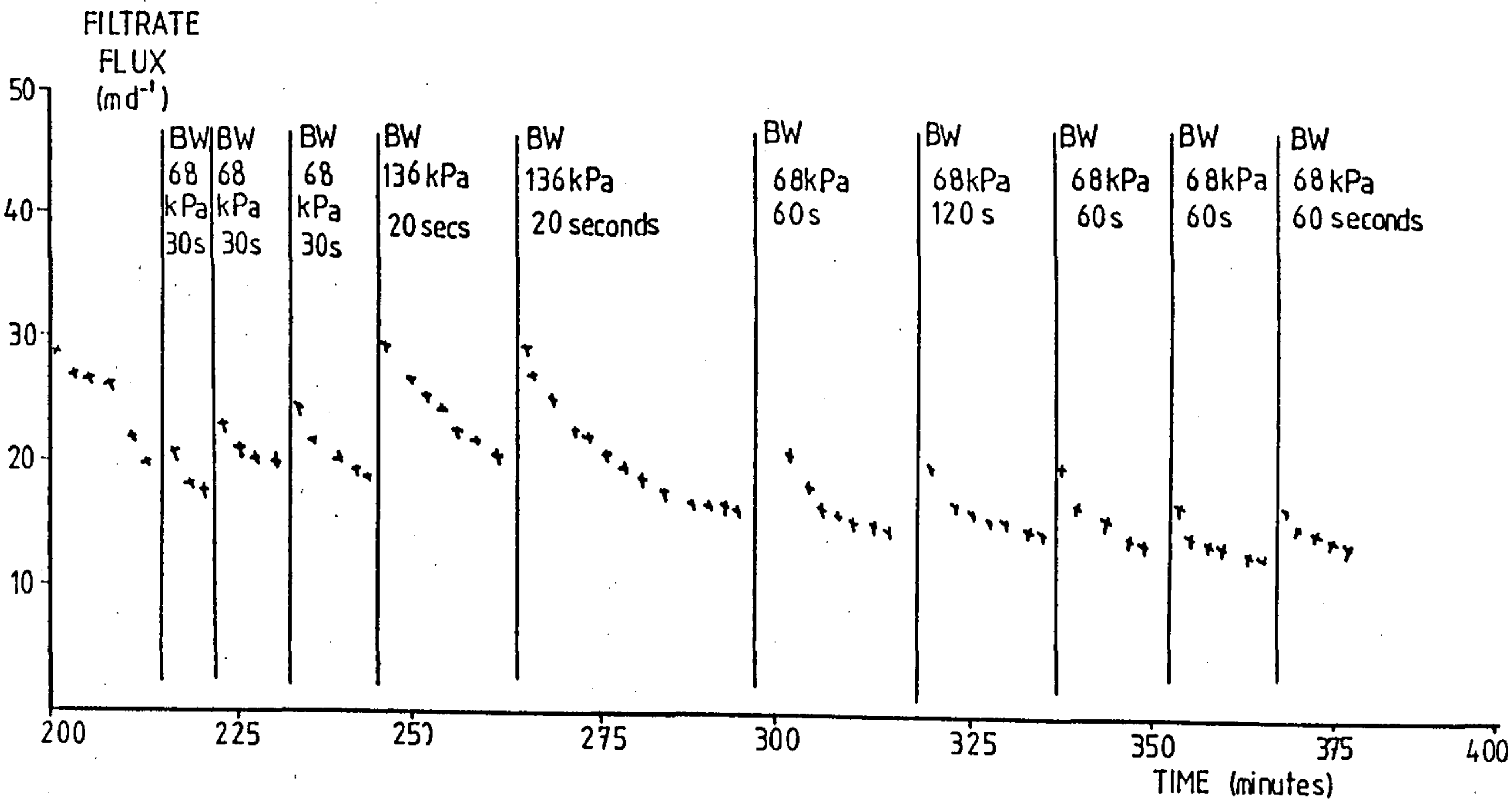
GRAPH 5.7 This run is similar to that in Graph 5.6, except that a higher flux was achieved. The backflushes were all 272 kPa and 1 second in duration. The membrane was Versapor 1.2 microns with a filtration pressure of 136 kPa and a crossflow velocity of 1.2 m/s.



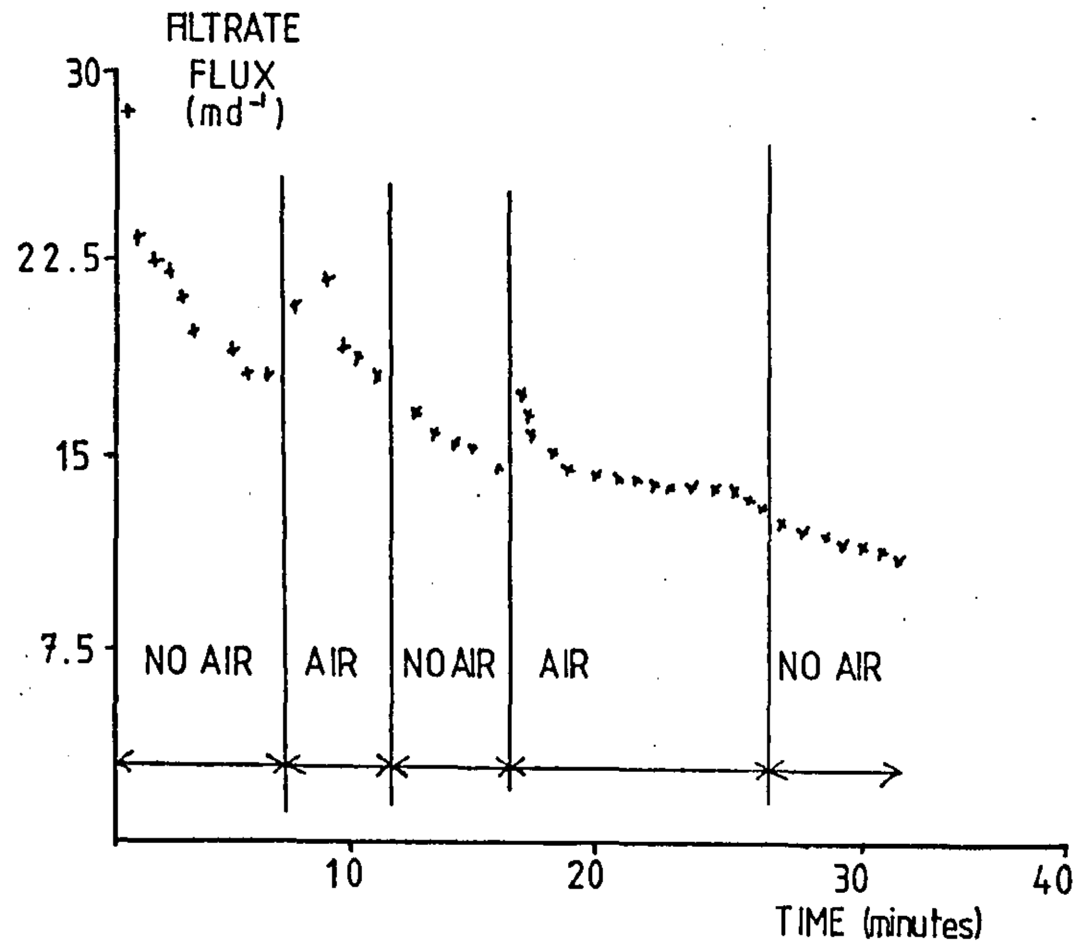
GRAPH 5.1 Flux vs time curve showing a typical example of how flux decreases with time.



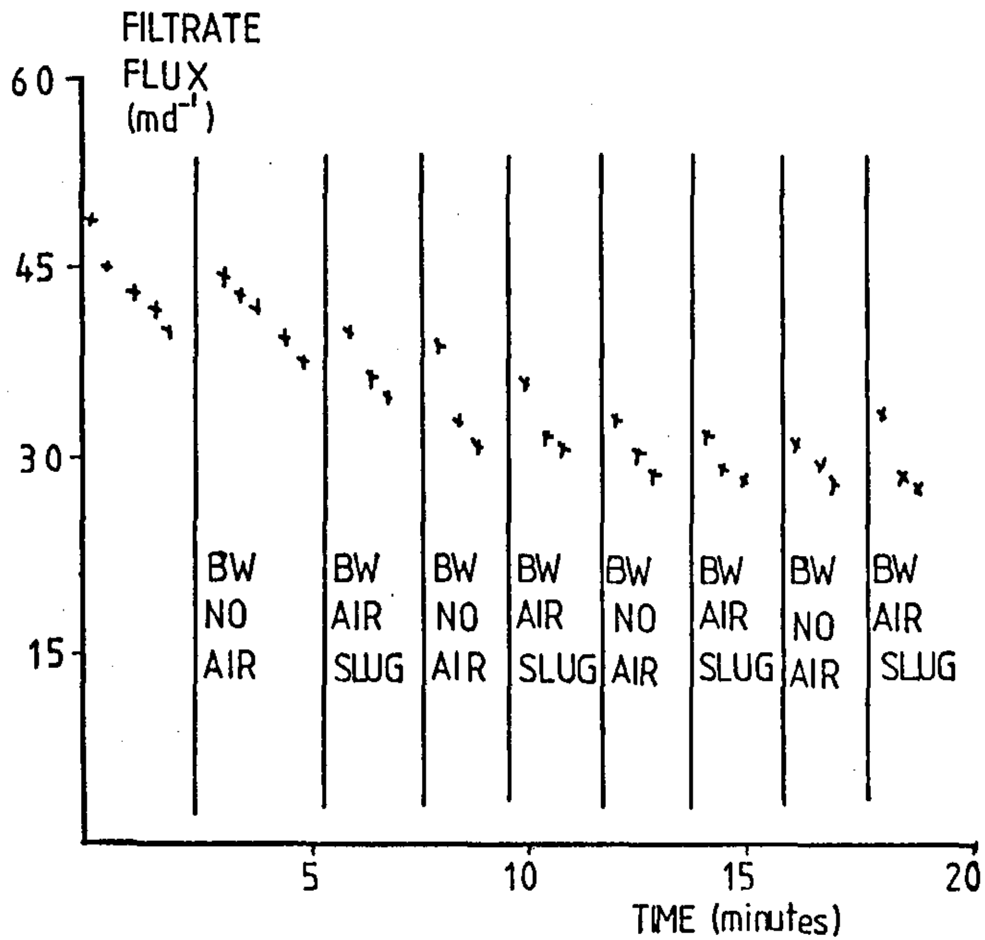
GRAPH 5.2(a) Flux vs time curve with several filtrate backflushes of different pressures and durations.
Continued overleaf.



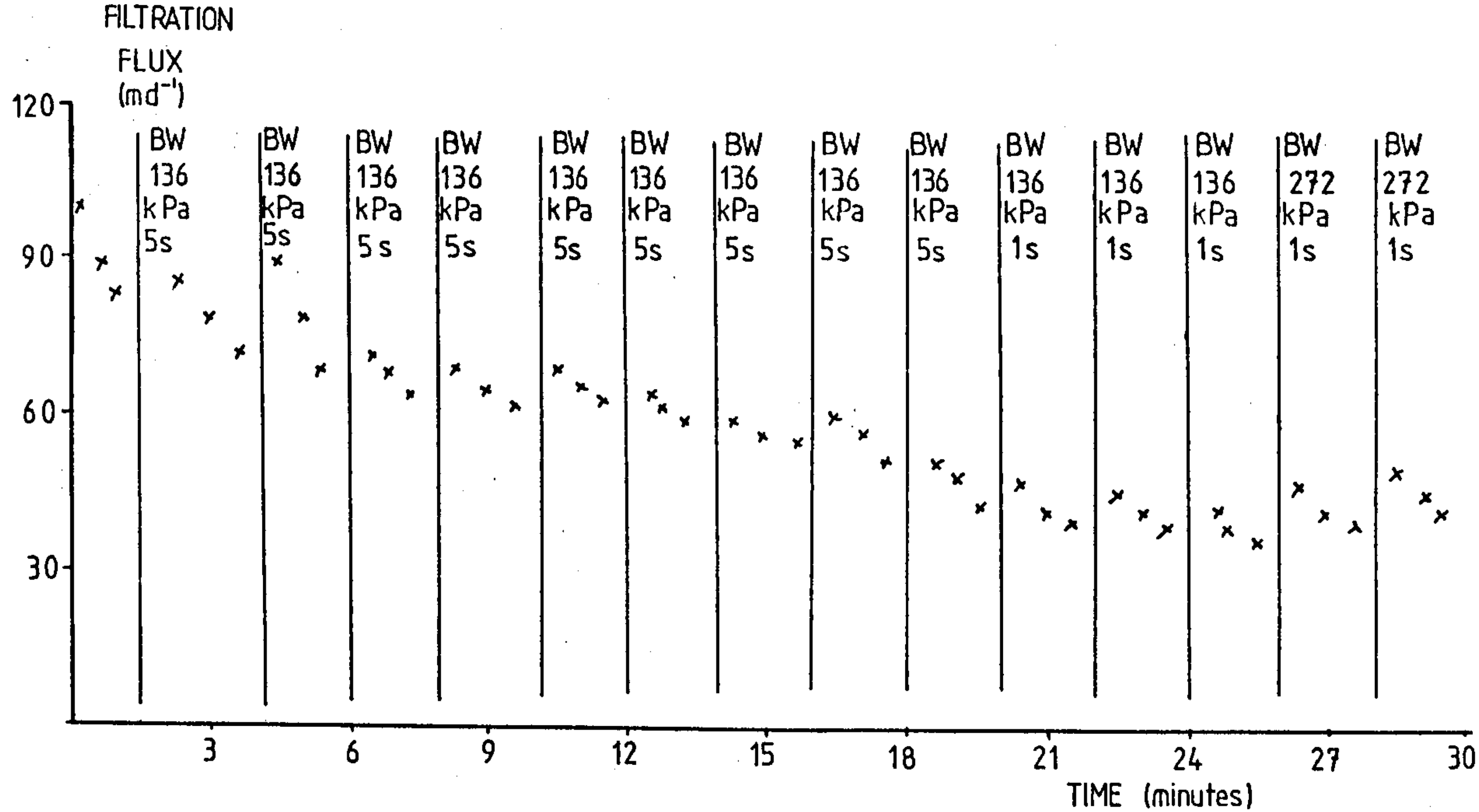
GRAPH 5.2(b) Continuation of graph 2(a).



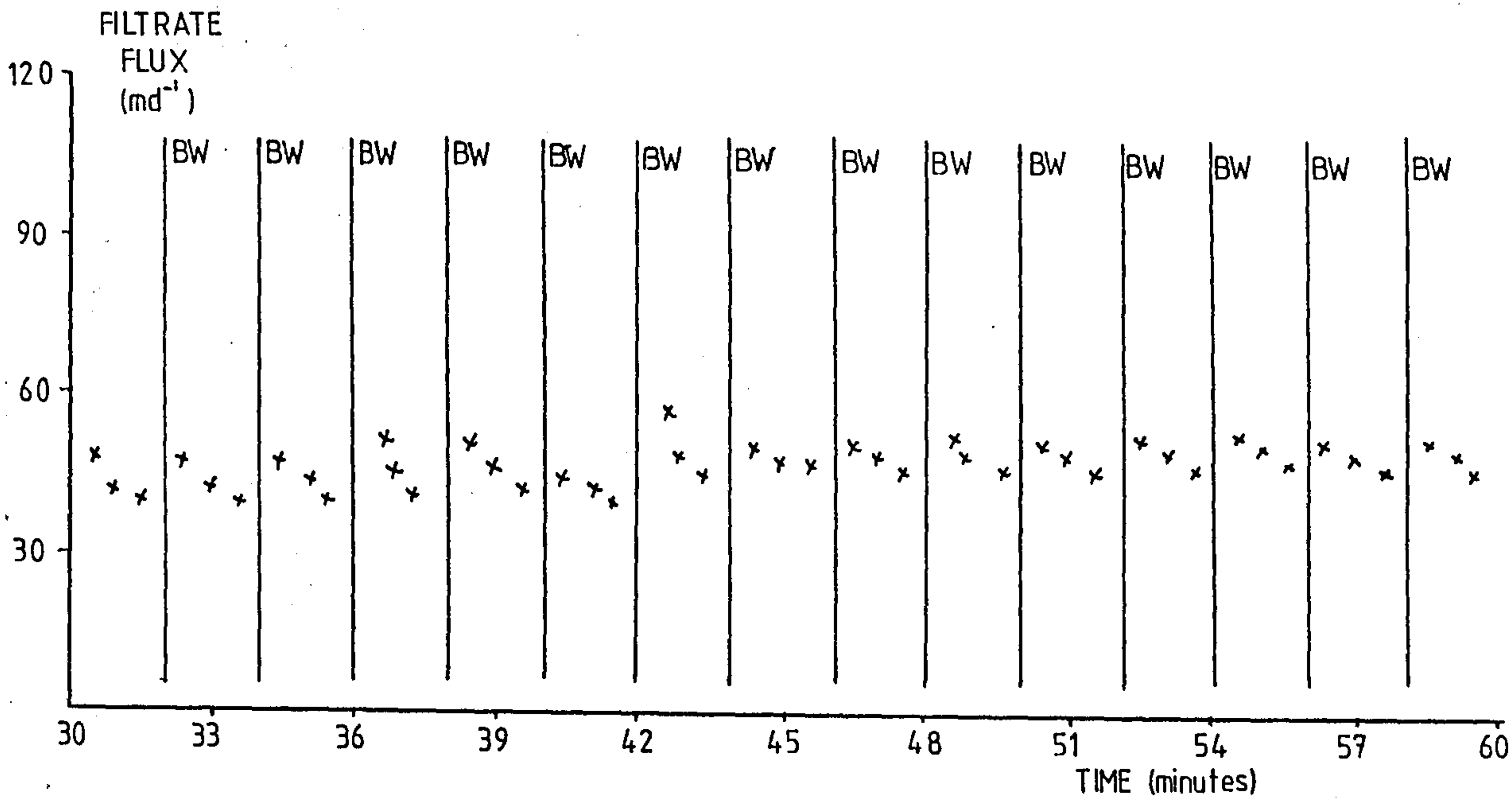
GRAPH 5.3 Flux vs time curve showing the effect of the introduction of purged air into the feed prior to the filtration channel during the filtration.



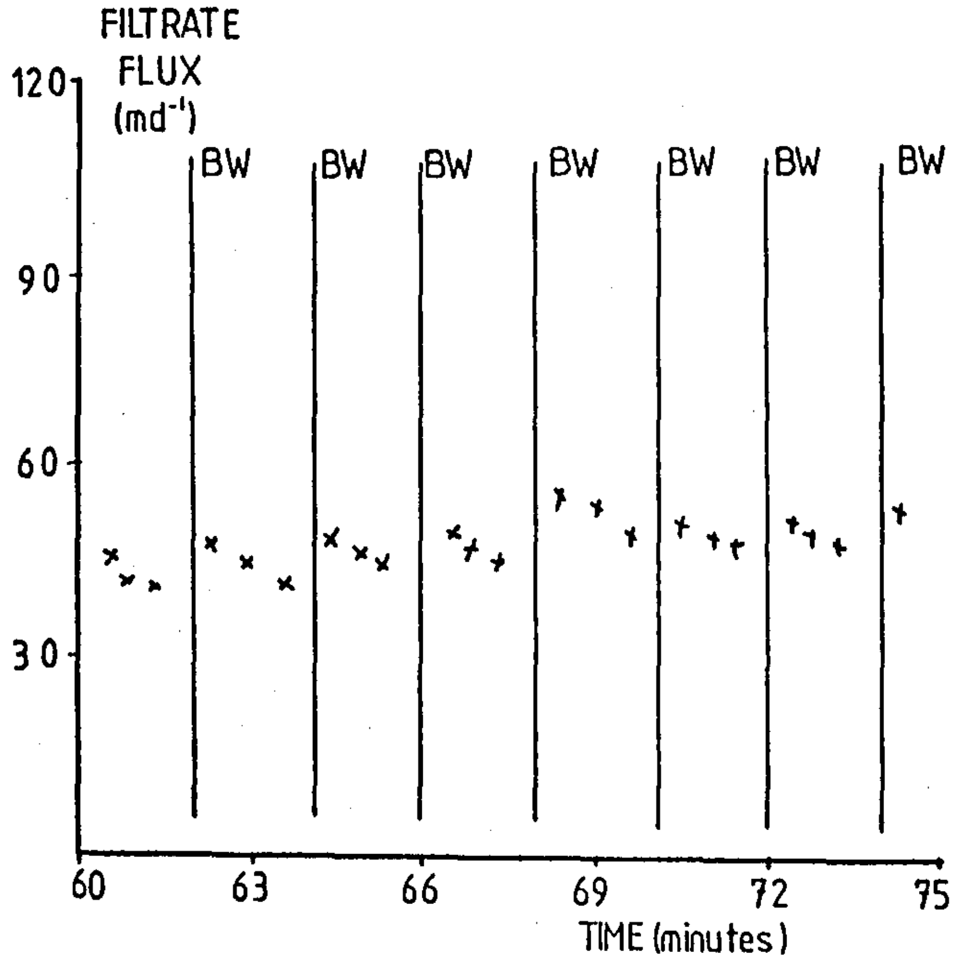
GRAPH 5.4 Flux vs time curve showing the effect of the introduction of purged air into the feed prior to the channel during a filtrate backflush. All backflushes were 41 kPa and 10 seconds long.



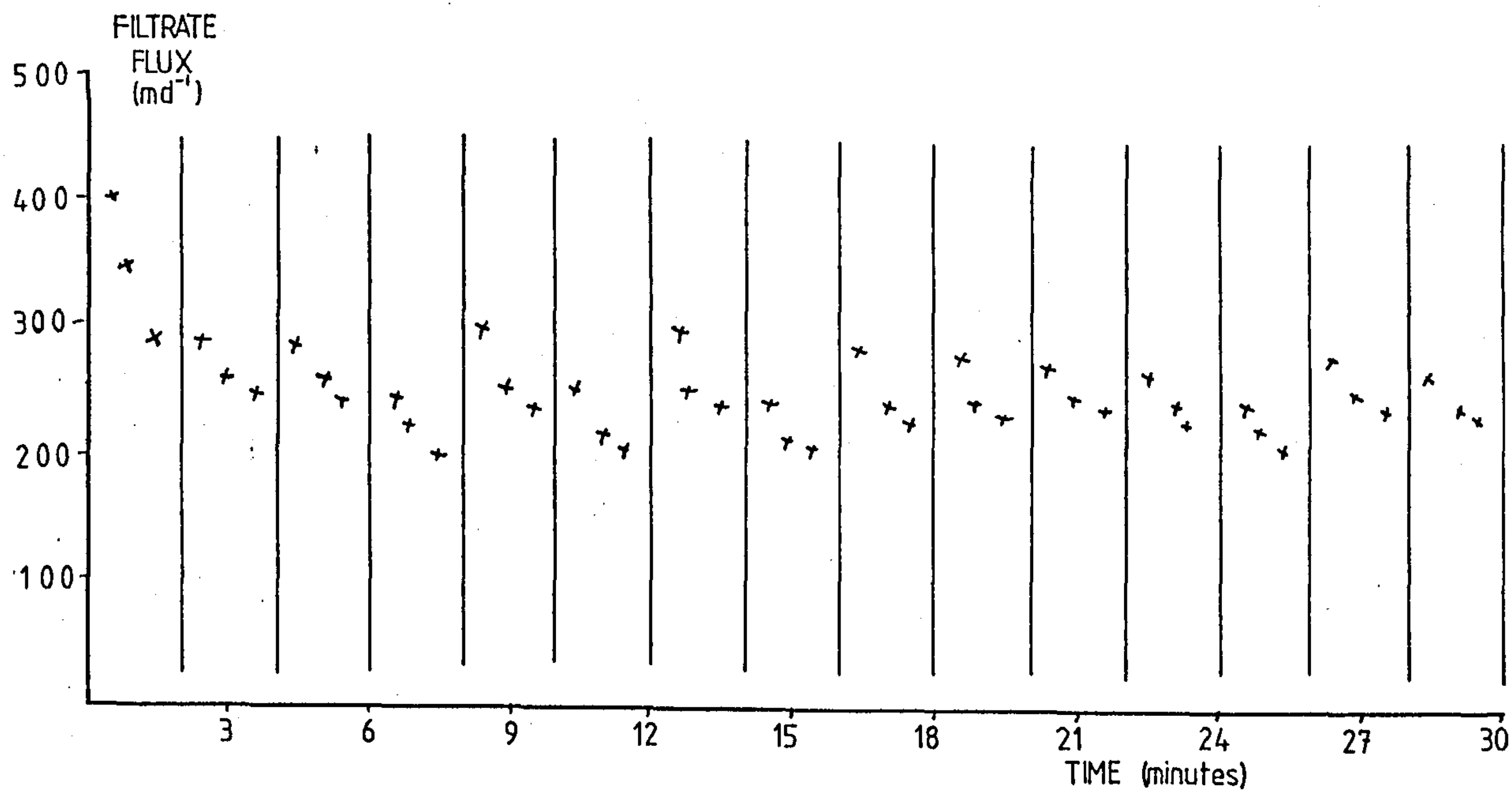
GRAPH 5.5(a) Flux vs time graph with several different compressed air backflushes. Continued overleaf.



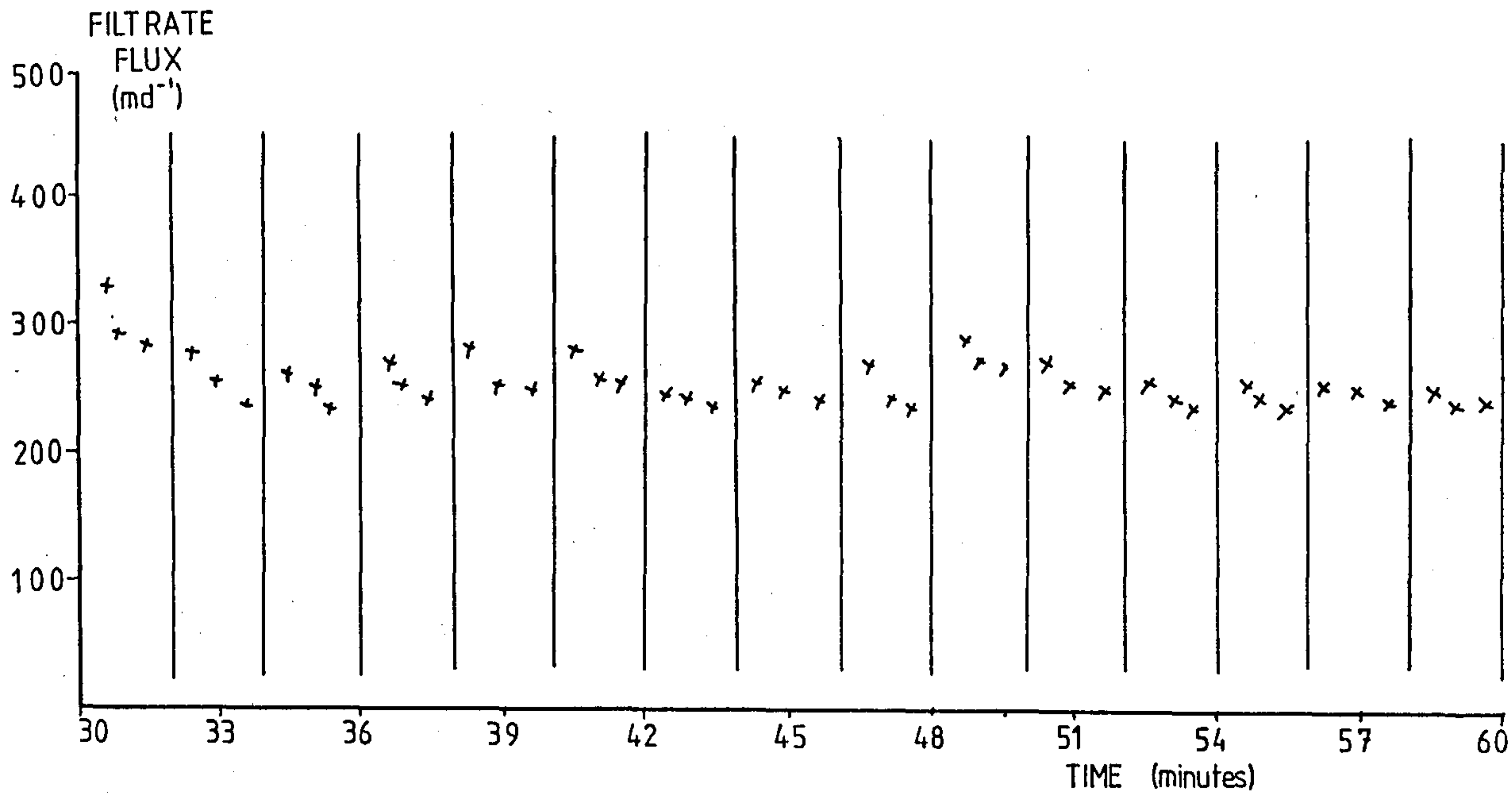
GRAPH 5.5(b) Continuation of graph 5(a). All backflushes are compressed air with a pressure difference of 272 kPa and a duration of 1 second. Continued overleaf.



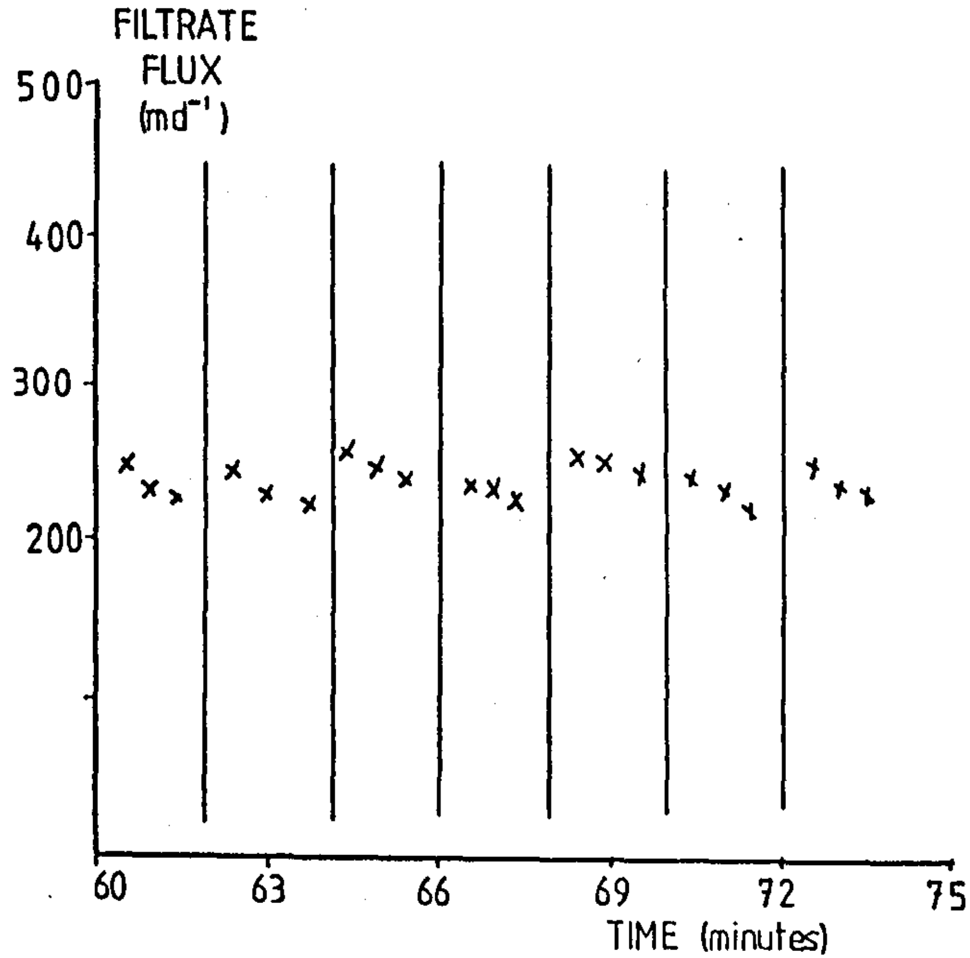
GRAPH 5.5(c) Continuation of graph 5(b). All backflushes use compressed air. The pressure difference for a backflush is 272 kPa and a duration of 1 second.



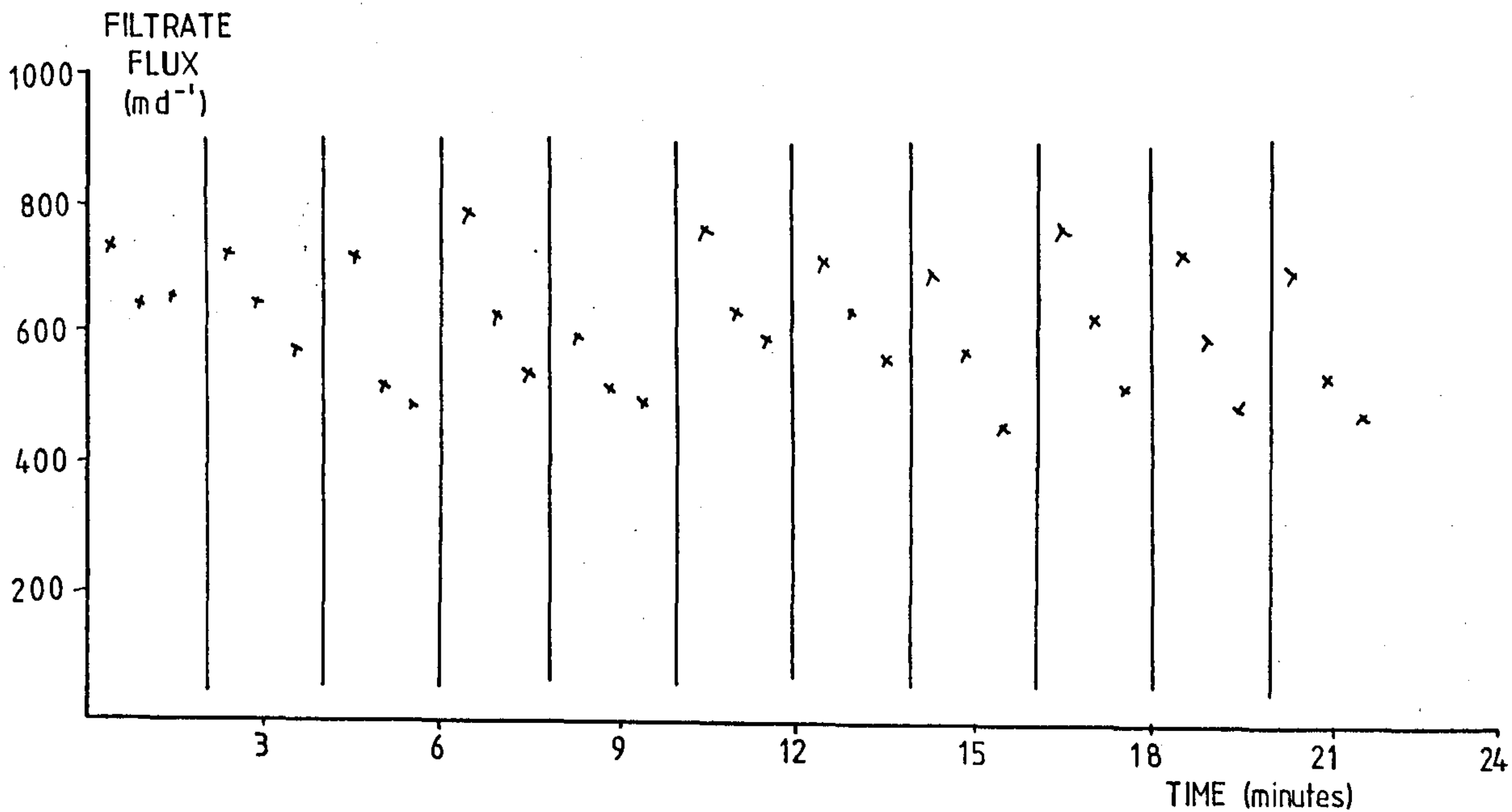
GRAPH 5.6(a). Flux vs time graph utilizing air backflushes at a pressure difference of 272 kPa and a duration of 1 second. Continued overleaf.



GRAPH 5.6(b). Continuation of graph 6(a). Air backflushes with a pressure difference of 272 kPa and a duration of 1 second are used. Continued overleaf.



GRAPH 5.6(c). Continuation of graph 6(b). Air backflushes with a pressure difference of 272 kPa and a duration of 1 second are used.



GRAPH 5.7 Flux vs time curve using air backflushes with a pressure difference of 272 kPa and a duration of 1 second.

DISCUSSION

The aim of the experiments described in this chapter was twofold. Firstly it was intended to gain information regarding the variables relating to a backflush, and secondly to try to find the optimal operating cycle. Any optimum, to be commercially viable, should allow a constant, or nearly constant, average filtrate flux over a long period of time.

As is shown in Graph 5.1, the filtrate flux declines over a period of time. One reason for this decline is that the particulate material in the feed progressively blocks the pores in the membrane. As time passes the amount of material blocking the membrane increases, the area available for flow through the membrane decreases, and consequently the filtrate flux decreases. Some workers in the field have found that the filtrate flux does not decrease to zero, but stabilizes at some equilibrium value. No evidence of this has been found in these experiments. However if the particulate matter in the feed was substantially larger in size than the pores in the membrane, then the mechanism of blocking described above, is no longer applicable. In these circumstances the filtrate flux becomes a diffusion controlled process, and an equilibrium flux could well be reached. Hermia in his paper describes various models of blocking filtration and an analysis of the various models on the results in Graph 5.1 are given in Appendix 2. In this particular application the feed material has some fine particulate matter in it and this would explain why no equilibrium value was reached.

Therefore it is necessary to continually, or at frequent intervals, clean the membrane so that the average flux is kept as nearly constant as possible. There are many methods that have been proposed that might achieve this. They are described elsewhere in this thesis. The simplest of these is reverse flow, or backflushing.

All the backflushing experiments described were performed with the feed material still flowing across the membrane. If the crossflow is turned off during a backflush, then the particles that are dislodged from the membrane will not be transported away from the membrane as readily. Thus it is likely that particles dislodged from the membrane during the backflush will be trapped again in the membrane when normal filtration resumes. Indeed the minimum time for a backflush should be equal to the length of the membrane divided by the crossflow velocity. If the duration of the backflush is shorter than this then particles dislodged from the membrane at the feed end of the channel will be deposited towards the other end of the channel. The backflush might have to be longer than this criteria, but should never be less.

Graphs 5.2 and 5.5 give data for backflushes of varying pressure differences and durations. On Graph 5.2(a) the backflushes at 160 minutes and 175 minutes were no more effective in restoring the filtrate flux, than were the preceding backflushes, although they lasted for 30 seconds as opposed to 5 seconds and the pressure differentials were the same. However, the backflushes at 245 minutes and 265 minutes were substantially more effective than the preceding and succeeding backflushes. These effective backflushes were shorter in duration (20 as opposed to 30 seconds) but the pressure difference was double (136 as against 68 kPa). This tends to suggest that the flux recovery due to backflushing is nearly independent of backflush duration, but strongly dependent on the effective pressure difference during the backflush.

Further evidence for this hypothesis is found in Graph 5.5(a). The 136 kPa, 5 second and 1 second backflushes were equally ineffective, but the 272 kPa, 1 second backflushes produced a constant average flux for 45 minutes. Graphs 5.6 and 5.7 used 272 kPa, 1 second backflushes every 2 minutes and in both cases constant average fluxes were produced. In Graph 5.7 the average flux was a high 600 m/d. Thus, it seems reasonable to suggest that the critical variable in backflushing is the pressure difference and that, provided a minimum backflush duration is observed, any longer duration gains only a minimal improvement in flux recovery.

This evidence suggests that a backflush is an instantaneous, or at least very rapid, operation and once any initial dislodgement has taken place, prolongment of the backflush condition does not cause any further material to be dislodged.

As mentioned in the introduction to this chapter, it was anticipated that the introduction of purged air would cause the membrane to be scoured and result in a less rapid decline in filtrate flux. Graph 5.3 shows an experiment where air was purged into the feed channel to test this hypothesis. The experiment consisted of periods where a slug of air was introduced every 30 seconds alternating with periods of normal filtration. In the periods where the air was introduced, there was an increase in flux, but there was no evidence that the flux declined less rapidly. The increase in flux is a pressure effect, the introduction of the air raising the pressure in the feed channel by 25%, and consequently increasing the pressure difference across the membrane. Thus there is no evidence to suggest that the air purging has a beneficial effect on the rate of flux decline and membrane blockage.

Another strategy for the introduction of purged air was to introduce the air into the crossflow during a filtrate backwash. It was anticipated that the scouring effect might assist the backflush to dislodge any trapped material on the membrane surface. Graph 5.4 shows an experiment that tests this. Purged air was introduced as a slug into the feed during alternate filtrate backflushes. There was no indication that backflushes in which the air was purged, were any more effective than the normal filtrate backflushes. These results indicate that the air scour principle is not effective in crossflow filtration.

There are three obvious backflushing fluids that could be used; a) feed water, b) filtrate water, and c) compressed air. Feed water was not used in these experiments, although some authors do use this method (ref Harwell report). This author believes such a strategy to be unwise. The risk of contamination of the filtrate by the feed is an obvious danger, but in this low concentration application it is not a major consideration. Far more serious is the risk of blocking the membrane during backflush, and the possibility that this blockage will become immovable. During the backflush the membrane is unsupported and thus is stretched by the pressure difference. Any particle that lodges on the membrane during the backflush may become trapped when the membrane relaxes. Such a blockage could not subsequently be removed, and will thus permanently damage the membrane. For this reason such a policy of backflushing with an unfiltered fluid is ill-advised.

It was found during these experiments that filtrate fluxes were substantially higher during a backflush than they were during normal filtration. Indeed in one case it was found to be 5 times higher. This means that if a significant fraction of the filtration cycle was spent backflushing, a very high proportion of the filtrate production could be used for backflushing, leaving a smaller amount of the filtrate as product. The reason for this higher flux is that the membrane area is substantially enlarged by the stretching of the membrane, and also the apparent pore size of the membrane is increased. Graphs 5.2, 5.5, 5.6, and 5.7 indicate that the compressed air backflush is at least as effective, and probably more effective than the filtrate backflushes. This seems to indicate that compressed air backflushes are more viable than filtrate backflushes. It also involves a lower capital cost. Compressed air is available on many sites whereas filtrate backflushes require an additional pump. Where compressed air is not available, air cylinders could provide an inexpensive and portable supply. Thus the optimum backflushing strategy appears to be frequent, short, high pressure compressed air backflushes.

From an examination of the graphs included in this chapter, it is obvious that the filtrate fluxes recorded vary markedly from run to run. The reason for this is the history of the membrane prior to the commencement of the run. In some cases the same membrane is used for several different experiments. Replacement of the membrane in the rig used for these experiments is a complicated and tedious procedure that takes several hours. The results presented in this chapter are only a sample of the total experimentation, and to replace the membrane prior to each run would have been impractical. The run described in Graph 5.7 was for a new membrane and should provide an indication of the fluxes that are obtainable.

It was noticed that when the pump was switched off, the filtrate flux was sometimes higher on the recommencement of the run than it was prior to the stoppage. A plausible explanation for this behaviour could be that the relaxation of the membrane that occurs when the pressure is released, frees and dislodges some of the material blocking the membrane. Thus switching off the pump has an effect analagous to a low-pressure backflush.

This chapter has examined the variables that influence backflush performane. A high-pressure, short duration, compressed air backflush was found to be optimal. These backflushes should be used frequently.

EXPERIMENTS WITH DIFFERENT CHANNEL GEOMETRIES

INTRODUCTION

The majority of experiments in this thesis, and indeed in the literature, have been performed using crossflow channels of rectangular cross-section. Although rectangles are perhaps the most obvious shape, other geometries do have some advantages. The design section of this thesis details some of the advantages of a triangular channel. The prototype design developed in this thesis specifies a triangular channel. An assumption of the analysis that led to this choice was that the channel shape does not influence the filtrate flux behaviour, especially with respect to time. It is necessary to know therefore whether the value of the average filtrate flux chosen as a design parameter, which was obtained on a rectangular channel, is appropriate for a triangular channel.

In order to justify this assumption, an experiment was performed to test this hypothesis. Flux vs time curves were obtained for five different channel shapes, and a comparison was made to see if the above assumption was valid.

EXPERIMENTAL PROCEDURE.

An obvious requirement for these experiments was an experimental rig that had a number of different crossflow channels of varying shapes. Rig C was designed with this in mind. All the experiments were performed with a constant velocity and pressure differential. It was not possible to keep the Reynolds number constant (to do so directly contradicts the constant velocity assumption), but there were two channels that had the same shape but varied in dimensions. If a shape dependence of filtrate flux is noticed, it should be possible to establish whether the important factor is channel shape or Reynolds number.

The five channels provide a spectrum of shapes that might be practicable. There are three triangular and two rectangular cross-sections. A description of the five channels is given below.

CHANNEL	DESCRIPTION
1.	Triangle 75, 45, 15 5mm deep. Membrane width=6.3mm
2.	Triangle 45, 90, 45 5mm deep. Membrane width=10mm
3.	Triangle 45, 90, 45 7.5mm deep. Membrane width=15mm
4.	Rectangle . 5mm deep. Membrane width=5 mm
5.	Rectangle. 5mm deep. Membrane width=10 mm

The procedure for these experiments was as follows. Each channel was connected in turn and the valves adjusted to deliver a constant velocity and pressure. A new membrane was installed prior to these experiments and as each channel was isolated from each other, effectively a new membrane was used for each channel. The velocity for each run was approximately 4 m/s and the pressure differential was 136 kPa.

RESULTS.

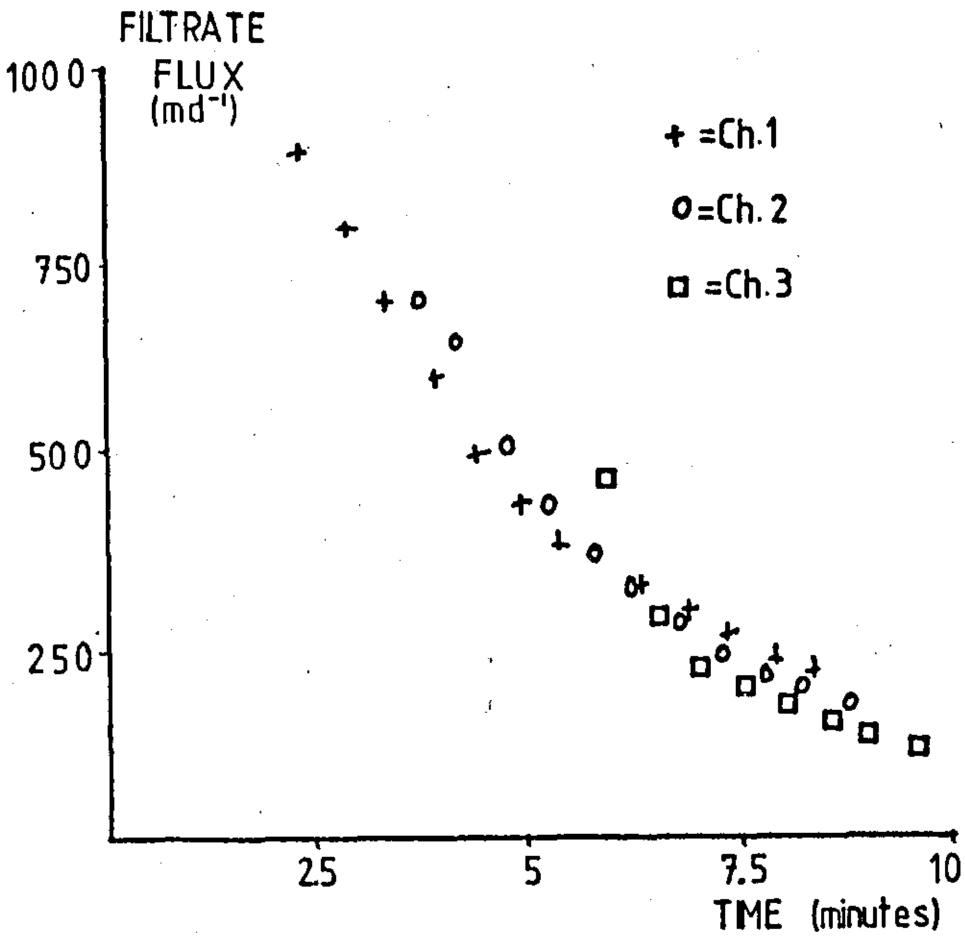
To present the flux time data for all the channels on one graph, would make that graph so cluttered as to be unreadable. Thus the results are presented in two graphs, one for channels 1,2, and 3 and the other for channels 1,4, and 5. Channel 1 provides a basis for comparison on both graphs.

DISCUSSION.

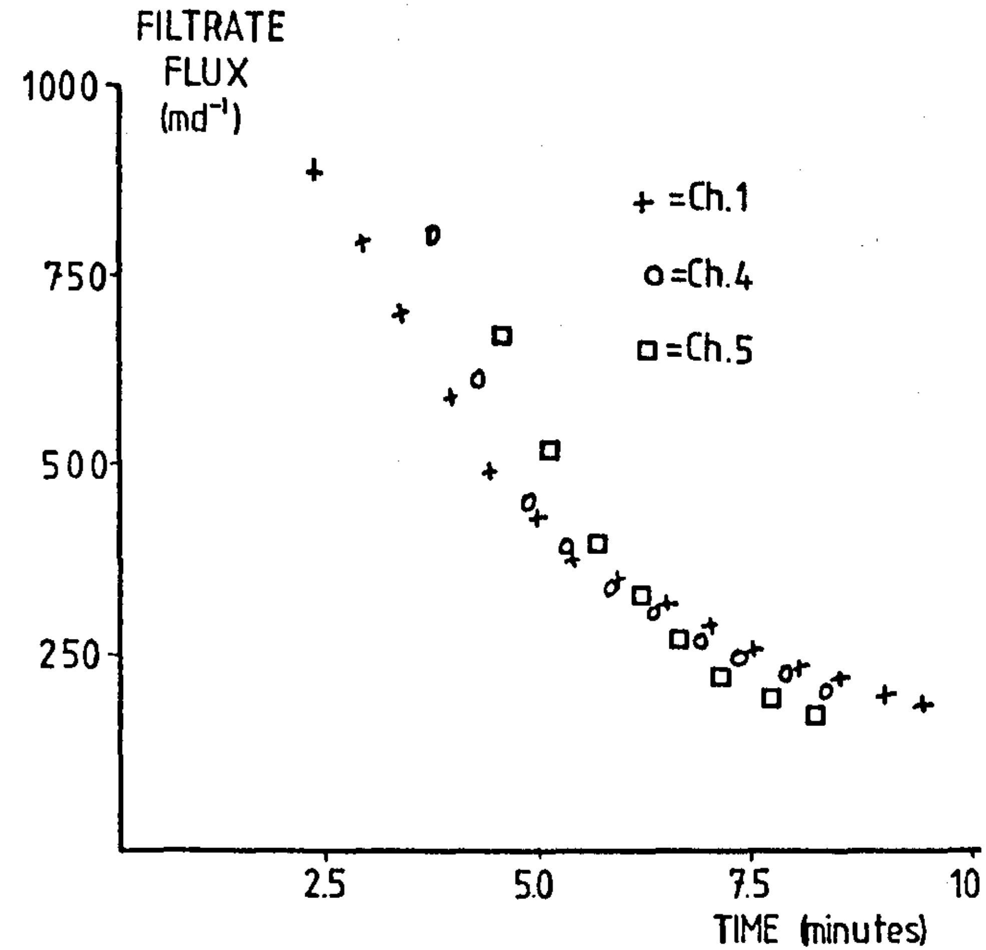
Graphs 6.1 and 6.2 support the assumption that the filtrate flux behaviour is independent of channel shape. There is no significant difference in the flux-time curves for the different channel geometries. Thus it is reasonable to assume that the channel shape is an important factor in the filtration behaviour.

The Reynolds number for the different channels varies between 13,500 amd 26,700. The evidence from these experiments is that Reynolds number is not an important factor in crossflow filtration. Harrison et al (37) give data in their paper that comes to a similar conclusion. Thus the most important factor seems to be velocity, or perhaps velocity gradient (or shear stress), and not the Reynolds number.

These experiments seem to justify the choice of triangular channel in the prototype design, and more particularly the assumption that data obtained on a rectangular channel can be applied to a crossflow channel of a different cross-section.



GRAPH 6.1 Flux vs time curve for different channel geometries. See text for details.



GRAPH 6.2 Flux vs time curve for different channel geometries. See text for details.

FILTRATE QUALITY TESTS

INTRODUCTION

An important factor in the design and evaluation of any separation process is the degree of separation achieved. In this case this is manifested as the quality of the filtrate that is delivered by the filter. It is not valid to compare two devices on their filtrate flux alone; the quality of filtrate is as important a criteria.

There is very little literature as to what is an acceptable filtrate. Collins (3) states that all suspended solids must be removed, whereas others claim that particles above 1 micron must be rejected. There is no data on the concentration of material that is allowable. There is no data on the maximum size of the particles that is critical, or the concentration of an intermediate sized particle. Thus there is a situation where a filter has to be designed to a very loose quality specification.

There is also very little information as to the likely concentration and particle size distribution of the feed to the filter. It is assumed that the concentration of suspended solids in the feed is low; the literature tends to suggest values of only a few parts per million.

The measurement of concentration and particle size in these very dilute suspensions is very difficult. Three possible methods which give quantitative results may be applicable : turbidity measurement, Coulter counter, and a laser difractometer. These are described in the following paragraphs.

Turbidity measurement does not measure particle size, but the concentration of suspended particles. The method relies on the obscuration of light by these suspended solids, the amount of obscuration being related to the concentration of the suspension.

The Coulter counter is able to determine both a relative concentration, and a particle size distribution. The principle of this device is that the electrical resistance across a narrow tube varies as particles flow through that tube. The variation in electrical resistance is proportional to the volume of the particle.

The Malvern Submicron Sizer is a laser defraction device. It can give both a relative concentration, and the mean size of the particles in that suspension. The operating principle of this sizer is a correlation of the signal received by reflection of particles in a laser beam. The magnitude of the Brownian motion of the particles is determined, and it is possible to derive a mean particle size from this.

Each of the above devices has its disadvantages. Turbidity measurement gives no information as to particle size. The Coulter counter cannot measure a particle with a size below 0.6 microns (0.3 microns is the absolute limit with the smallest tube available, but the noise of the system and the complexity of technique required impose a pratical limit of about 0.6 microns). The Malvern Submicron Sizer will only give a mean size. However it was thought that the maximum amount of information could best be obtained using the Malvern machine.

There are several variables that are likely to influence the quality of filtrate produced. These include the particular membrane in the filter, the nominal pore size of the membrane, and the history of the membrane (eg the number of backflushes to which it has been subjected). These can be classed as membrane variables. Process variables will also influence the quality of the filtrate; these include the crossflow velocity, the applied pressure, and the nature and concentration of the particulate material in the feed.

Ideally it is desirable to test each of these variables in turn, and determine how, and to what extent, each influences the filtrate quality. In these tests it was decided to test only three of these variables; the nominal pore size of the membrane, the crossflow velocity, and the effective pressure difference across the membrane.

EXPERIMENTAL PROCEDURE.

Two tests are described in this section. The first is to determine the effect of crossflow velocity and applied pressure, and the second the effect of nominal pore size. The feed material for both experiments was carefully controlled. Some fine AC (Feedspar) dust was mixed with water and Sodium pyro-phosphate and left in an Andreason jar to settle. After three days, the top section of this liquid was removed with a pipétte, and this very fine suspension was used as an additive to the feed. On one sample of this suspension a disc centrifuge test was run, so as to obtain a particle size distribution of this material. Unfortunately the centrifuge developed a fault during the test and no data was collected. However sedimentation theory predicts that the material will all be under 1 micron. A small crossflow filter with a 0.2 micron Nucleopore membrane was used to filter the water that was used to fill the feed tank. The suspension, described in the above paragraph, was then added to the feed tank, so that it gave a concentration of suspended solids of approximately 100 ppm.

For these tests Rig C was used. As it is entirely constructed of aluminium it is easier to keep the system clean, especially on the filtrate side of the membrane. Obviously the rig and the samples must be kept free from extraneous contamination so as to give the correct results. The samples were kept in small sample tubes, which were first washed out with filtered water, and prior to the introduction of the actual sample, the tube was rinsed with the filtrate. Where dilution of the sample was required, distilled and filtered water was used.

For the first test the following procedure was used. The valves were adjusted so that a crossflow velocity of 1.66 m/s and a pressure of 126 kPa was delivered by the pump. A 60 second period was allowed so as to give steady state conditions, and the samples were taken. The valves were then adjusted so as to give the next measurement condition. Four operating conditions were used for these tests and these are summarised below.

RUN No.	VELOCITY	PRESSURE	CONDITION
1.	1.66 m/s	136 kPa	Low Pressure Low Velocity
2.	1.66 m/s	272 kPa	High Pressure Low Velocity
3.	4.0 m/s	272 kPa	High Pressure High Velocity
4.	4.0 m/s	136 kPa	Low Pressure High Velocity

All the tests were performed on Channel 2, which has a cross-section that is a right-angle triangle, 10mm wide by 5 mm deep.

The procedure for the second experiment was as follows. First two membranes were joined along their edges, with the left half of the membrane being Versapor with a 1.2 nominal micron pore size, and the right half having a 5 micron nominal pore size. The membrane was then inserted in the rig, so that channel 2 had the 1.2 micon membrane, whereas channel 4 had the 5 micron membrane. The lines were then attached to channel 2 , the pump was switched on, and the valves adjusted to give a crossflow velocity of 1.66 m/s and a pressure of 136 kPa. Sixty seconds were then allowed to elapse to allow the filter to achieve steady state, and then the filtrate samples were taken. Channel 4 was then connected and the procedure repeated with the same operating conditions.

Three samples were taken for each operating condition. This was to ensure that each sample was representative, and contamination had not occurred. When each sample was analysed, the results were averaged over the three samples so that a representative value was obtained. The samples were analysed on the Malvern Submicron Sizer. The sample time was chosen as 50 micro-seconds and the experimental duration as 10 seconds. Approximately 20 analyses were run for each sample, and the results were written down. Later the results were put into a digital computer and the mean, standard deviation and standard error of the mean was calculated for each operating condition.

RESULTS.

The results are presented in tabular form. The number of counts is a measure of concentration of suspended solids. The standard deviation of each result is due both to the variation within each sample, and also the variation between the three replicate samples for each operating condition.

RUN 1.

CONDITION		No.OF COUNTS.	SIZE (microns)
Tank Sample.	Mean	2,350,000	1.51
	Std Dev.	928,600	0.12
	Std Error.	218,900	0.06
Low Pressure	Mean	117,000	1.46
Low Velocity	Std Dev.	23,000	0.58
	Std Error.	4,200	0.15
High Pressure	Mean	95,900	2.33
Low Velocity	Std Dev.	21,900	0.75
	Std Error.	4,400	0.34
High Pressure	Mean	98,500	1.83
High Velocity	Std Dev.	54,600	0.44
	Std Error.	11,400	0.18

Low Pressure	Mean	92,300	1.14
High Velocity	Std Dev.	25,300	0.79
	Std Error.	4,600	0.24

RUN 2

CONDITION		No OF COUNTS	SIZE (microns)
Tank Sample	Mean	4,251,000	1.66
	Std Dev.	1,151,000	0.41
	Std Error.	257,000	0.09
1.2 Micron	Mean	170,100	3.55
Nominal Pore Size	Std Dev.	48,800	5.67
	Std Error.	10,400	1.57
5.0 Micron	Mean	173,100	1.94
Nominal Pore Size	Std Dev.	55,000	0.81
	Std Error.	12,000	0.18

Table of Filtrate Quality Results

DISCUSSION

The results given above are both interesting and disappointing. It was hoped that the mean particle sizes for all the filtrate suspensions could be determined accurately, and a meaningful comparison made. Unfortunately the size data are both inconsistent and in some cases nonsensical. For example, in Run 1 the tank sample often has a smaller mean size, than does the filtrate at the other operating conditions. This is even more noticeable in Run 2. The standard deviation of the size measurements are all rather large, and in one case actually exceeds the mean. This implies that the size data are erroneous; there is no conceivable mechanism where a feed with an average size of 1.66 microns, can be filtered through a 1.2 micron pore size membrane, and the filtrate emerge with an average size of 3.55 microns.

There are two possible explanations for this behaviour: a) flocculation, or b) errors in the analysis. When the suspension was manufactured (as described in the Experimental Procedure section) a dispersing agent was used. However when this suspension was added to the feed tank, the concentration of dispersant was then too low to be effective. Thus it is possible that at this stage, the dispersant reflocculated. However, the evidence from the analysis does not support this view. The analysis of the tank samples in both runs are the only ones that give consistent readings, and the individual machine runs are reasonably reproducible. It is possible that the filtrate flocculated after it passed through the membrane. This is a plausible hypothesis, and a possible mechanism for this behaviour would be a change in the electric charge on the particles as they passed through the membrane.

The other source of this discrepancy could be in the Malvern Submicron Sizer. The operating principle of this machine is a mesurement of Brownian motion. Brownian motion is random, and any systematic motion is filtered out of the signal. When large particles (in this case about 1 micron) settle in a fluid where there is a large density difference, the settling movement is many times larger than the Brownian motion. Thus it is very difficult to extract the true Brownian motion out of the signal received from these larger particles. Normally this sizer is used for suspensions of neutral density. Indeed the experience obtained in this department with its machine is confined to latices with density matched closely to that of water. These latices are also very small; usually they have a maximum size of about 0.1 microns. Thus the application for this thesis was outside the previous experience that the department had in operating this machine. Whether the machine is capable of being used successfully for these suspensions is open to question; however it could take many months to find a suitable procedure to operate it successfully and thus is outside the sphere of this thesis.

There is another procedure that could be used to gain more suitable results. This is to use suspensions that can be measured successfully on the sizer. The fine AC dust was chosen as the contaminant because it more accurately represents the particles that are likely to be found in a real application. There are two reasons why latices are unrealistic; a) their neutral density and b) their sphericity. The latter is especially likely to lead to anomalous results. Spherical particles seem less likely to block the membrane than do those of irregular cross-section. However the use of these latices could well give accurate results, even if the interpretation of these results is handicapped somewhat by the difficulties detailed above.

Although the size measurements are unreliable, the reasons above do not in any way invalidate the concentration reading (as given by the number of counts). For both Runs 1 and 2 there is no evidence that the operating conditions effect the quality of filtrate. In Run 1 the discrepancies in the number of counts between the two low pressure runs, and the two high pressure runs, were slight. There seems to be no reason why the applied pressure should affect the quality of the filtrate .For it to do so, the increased pressure must cause a separation, and increase in pressure will have an equal effect on the particles and on the fluid.

In Run 1 there is no apparent discrepancy between the two velocity conditions. As mentioned in the literature survey, Carter and Hoyland (24) reported that the rate of deposition of a fouling layer was independent of the tangential flow velocity. This is consistent with the results found above. It seems that two effects might be at work here: first the increased velocity promotes turbulence which might force more particles towards the membrane, and the second, the greater turbulence causes particles, once they get into the boundary layer, to have an increased chance of re-entering the mainstream, rather than going through the membrane. Neither of these effects appears to predominate, and it appears that filtrate quality is independent of crossflow velocity.

Run 2 tests whether the nominal pore size of the membrane influences the filtrate velocity. The concentration for the two membranes are within 1.5% of each other, and since this is well within experimental error, there is no indication that the pore size does effect the filtrate quality. The maximum pore size of the feed was about 1 micron, and thus if the membranes were perfect separators this result would be expected. (A perfect separator is one where any particle above a certain cutoff size is retained; any particle below the cutoff size is allowed to pass.) However membranes are not perfect separators (they have a range of pore sizes, not one absolute pore size) and this result is surprising. The 1.2 micron membrane is a "tighter membrane" and thus it would be expected to retain more particles than the looser 5.0 micron membrane. It is possible that the nominal pore sizes for the membranes are rather arbitrary, and that the larger pore size membrane is not as loose (or conversely the 1.2 micron membrane is not as tight) as the values attached to them seem to indicate. It would certainly seem worthwhile to do some more tests on this, and also to ascertain the flux increases that could be obtained with a larger pore size membrane.

Although all the experiments in this chapter were analysed by the Malvern Submicron Sizer, the other analysis methods (Coulter Counter,and turbidity measurement) should not be dismissed. Indeed, if a large pore size membrane with a feed suspension of suitable size distribution was to be used, analysis by Coulter Counter should provide some accurate results. Turbidity measurement provides a portable and easy method of analysis, and is especially useful for field, as opposed to laboratory, experiments.

Another method of filtrate quality measurement that has been used is the use of porous rock. The rate at which the filtrate blocks the rock is a measure of the particulate size and concentration in the filtrate. This method mimics the damage that will be done to an oil-bearing formation by injection water that has not been properly filtered. The method has three disadvantages however. Firstly the tests are tedious and not particularly accurate. Secondly the data obtained are not able to be translated into size and concentration measurements, and lastly, any irregularities in the rock samples will cause anomalous and confusing results.

QUALITATIVE OBSERVATIONS

There are some experiments that should be included in this thesis, that are qualitative, not quantitative. These observations are recorded in this chapter. Although comparative data are not recorded in these observations, and it is not possible to calculate the effect on filtrate flux, these observations are still useful. From them, it is often possible to speculate as to whether a particular phenomenon has a neutral, beneficial, or deleterious effect on filtrate flux.

Th High Speed Photography, as detailed in the Velocity Measurement (Section 9), showed two interesting effects. Both observations were made on a film that was photographed looking down onto the membrane surface (as opposed to the films taken for velocity measurement that were taken looking parallel to the membrane). The first showed that some of the particles on the membrane surface tended to move along the surface, often rotating as they moved. These particles, in order to be visible, had to exceed 20 microns in size, and it is not possible to ascertain whether a similar movement holds for much smaller particles. The second effect that was noticed was that when large particles (over 100 microns) were moving near the membrane surface, they tended to collide with smaller particles. These smaller particles then re-entered the mainstream. This observation then led to speculation about deliberately introducing larger particles into the feed and using these collisions as a method of minimising flux decline. However, the particles that were seen to re-enter the main stream had a size greater than 20 microns, and it is questionable whether these collisions would occur with much smaller particles, which are more likely to cause blocking of the membrane. The air scour, as detailed in the Backflushing (Section 5), relies on a similar effect to that detailed above.

One method that has been proposed to minimise the rate of filtrate flux decline, is to use obstructions to promote turbulence near the membrane surface. It was expected that the increased turbulence would act to clean the membrane continuously. To test this hypothesis, a suitable obstruction (a nylon thread) was placed on the membrane surface during a filtration run. When the membrane was removed after the run was concluded, it was noticed that there was a layer of dirt near the obstruction, extending for approximately 2 mm each side of the obstruction. It was also observed after several experiments, that where there was a crease in the membrane, a heavier concentration of particles was deposited near this crease than on other parts of the membrane. This evidence tends to suggest that the obstruction tends to cause more blocking of the membrane, not less.

Another experiment was run to reinforce this observation. Fluorescent powders of different colours were manufactured so that a colour represented a particular size fraction. Red particles had a size of approximately 5 microns, while the yellow particles were 1-2 microns in size. These powders were then mixed with the feed in equal concentrations, and the above experiment repeated. It was found that the area in the immediate vicinity (2 mm) of the obstruction was substantially redder in colour than the rest of the membrane. This implies that larger particles are deposited on the membrane as a result of the obstruction. It appears that the obstruction, rather than promoting turbulence, tends to create a region where particles are more likely to be deposited on the membrane. As mentioned in the Literature Survey, Sheppard and Thomas (12) used rough and smooth support plates for their membranes. They found that the rough plate gave no evidence of increased turbulence promotion, and that the rough plate was more susceptible to fouling. Their conclusions therefore are in agreement with the observations detailed above.

VELOCITY MEASUREMENT

INTRODUCTION

The literatue seems to indicate that the important factor in the prevention of flux decline, is the shear stress at the membrane surface. The shear stress is proportional to the velocity gradient at the surface. In order to ascertain this gradient, experiments were performed to measure the velocity in the region of the surface.

There are several methods of measuring velocities, but the circumstances in this instance dictated that the method used must satisfy three requirements. Firstly the method used must measure velocities remotely; that is measuring the velocities at a distance rather than in-situ. Secondly the technique had to determine velocities in a small element of volume, rather than spatially-averaged velocities. Finally the method had to be able to assess velocities in the close proximity of the membrane.

Two methods that seem to fulfil these requirements were a) Laser Anemometry and b) High-Speed Photography. Both these methods have the intrinsic disadvantage that the velocity measured is a particle velocity rather than a fluid velocity. If the particle is sufficiently small, then it provides a reasonable estimation of the fluid velocity, but the larger the particle the greater the deviation from the fluid behaviour.

LASER ANEMOMETRY

The laser anemometer is a modern velocity measurement device, which can provide very accurate estimates of particle velocities. It consists of a laser, a photon-detector, and a correlator. The laser beam is split into two separate beams which intersect in the zone where the velocity is to be measured. The two laser beams intersecting create an interference pattern. When particles flow across the measurement zone, they alternate between light and dark bands, which give the impression of very rapid flashes of light. These flashes can then be detected by the photon-detector which is focused on the measurement zone. By correlating the information from the photon-detector, the average time between flashes can be established. The velocity of the particles in the fluid is related to this time and also to the fringe spacing in the interference zone, and the velocity can be determined by a simple formula.

The actual value determined is not stictly speaking a velocity but rather a velocity component. It is not possible to determine which way a particle is travelling across the interference pattern, nor is it possible to determine the velocity components in the orthogonal directions simultaneously. (It is possible to determine them at a later stage by rotating the beams and hence the interference pattern.) Thus what is actually measured is the velocity component in either of two directions, one forwards across the interference pattern and the other backwards. Since the velocity components that are eventually determined are time averaged, the distinction made above tends to be rather academic.

If most of the photons detected are reflected from particles in the interference zone, the signal is a strong one and velocity measurement is precise. Frequently noise is also detected, and the resultant signal leads to a less precise velocity determination. In these particular experiments the closer the interference zone to the membrane surface, the more noisy was the signal from the photon-detector. When a measurement within 0.5 mm from the membrane surface was attempted, the signal was almost random, and it was not possible to derive any sensible velocity measurement.

These experiments were performed using Rig A (as described elsewhere). The arrangement of the anemometer is shown in Fig 9.1 (overleaf). The circulating fluid was water, seeded with a small concentration of Titanium Dioxide. These Titantium Dioxide particles are sufficiently

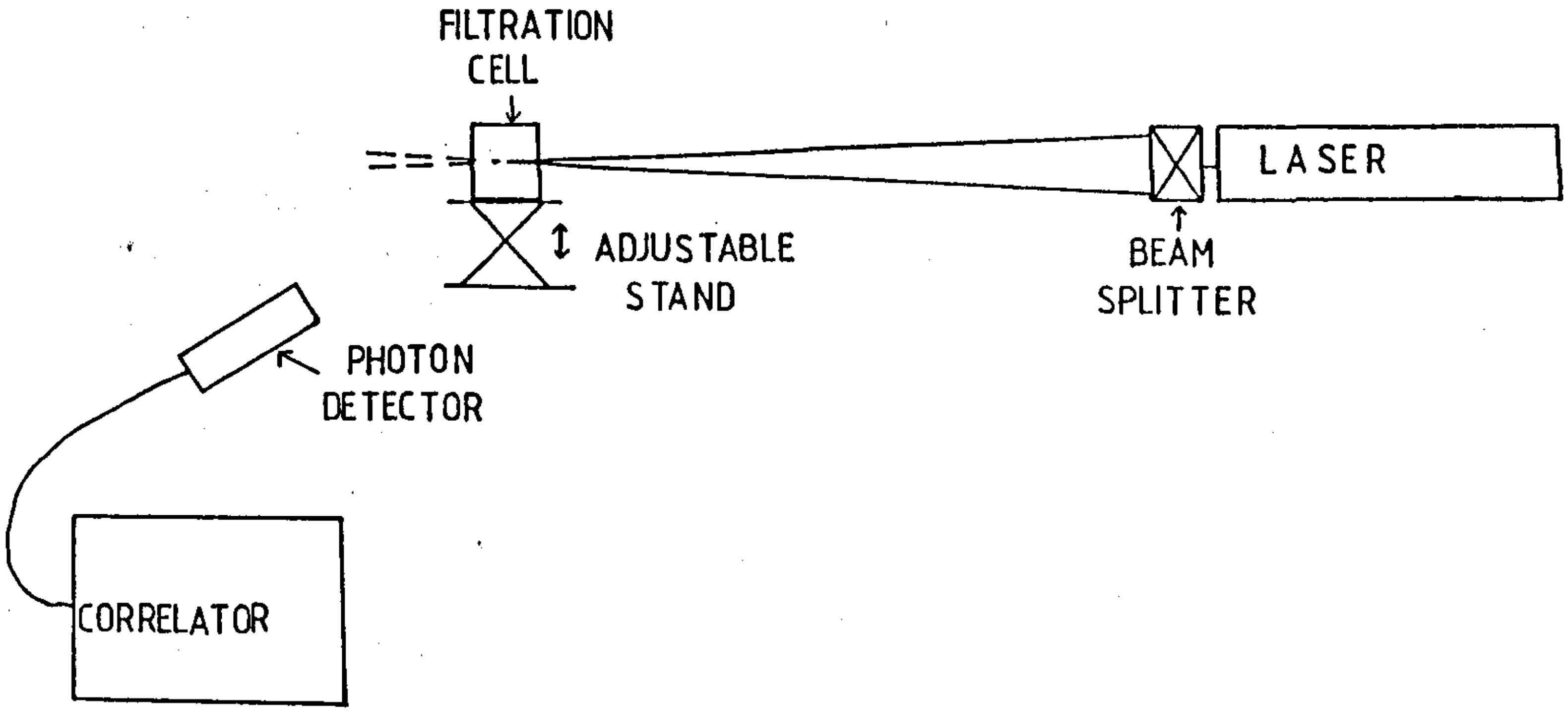


FIG 9.1 Diagram of laser anemometer and filtration cell showing the experimental arrangement.

small (less than 0.5 microns), so that they mimic the fluid flow very closely.

RESULTS

Three sets of results are presented in Graphs 9.1 - 9.3. Each is described briefly below.

Graph 9.1 This profile gives velocites across most of the crossflow channel. The average velocity is approximately 1.3 m/s with a Reynold's number of 13,300.

Graph 9.2 This profile has an average velocity of approximately 2 m/s, and a Reynold's number of 20,000. The profile is very much "flatter" than that in Graph 9.1.

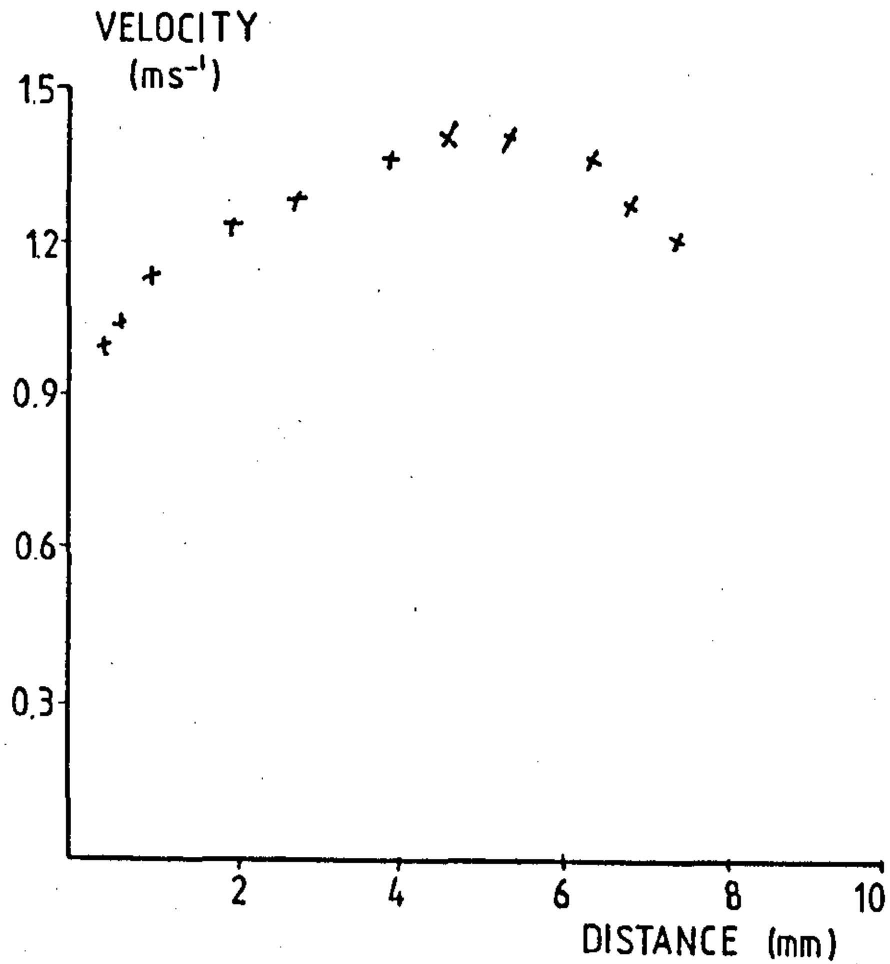
Graph 9.3 This profile has an average velocity of 0.7 m/s. The profile is a lot more curved than that in the previous graphs. The Reynold's number is 6,700.

HIGH SPEED PHOTOGRAPHY

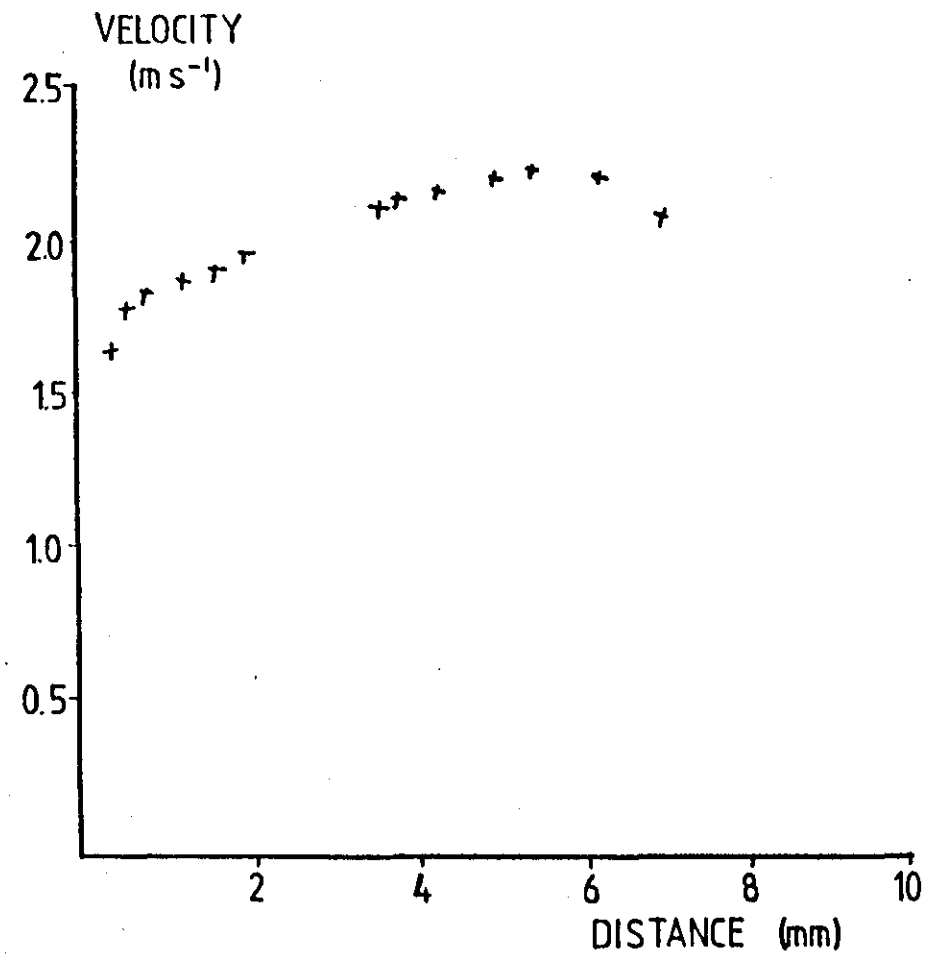
The principle of velocity measurement by high speed photography is relatively simple. The trajectories of the particles are recorded on the film, and these can be measured when the film is projected. Since the distance travelled by a particle can be measured, and the number of frames taken for this can be counted, the velocity can be determined simply, provided the frame speed is known.

This method has some drawbacks. It is necessary to have as great a magnification as possible, for two reasons. Firstly, tracer particles should be as small as possible, so that they mimic the flow of fluid closely, and secondly velocities close to the membrane have to be measured. In order to resolve both the particles and the membrane surface, a high magnification is required.

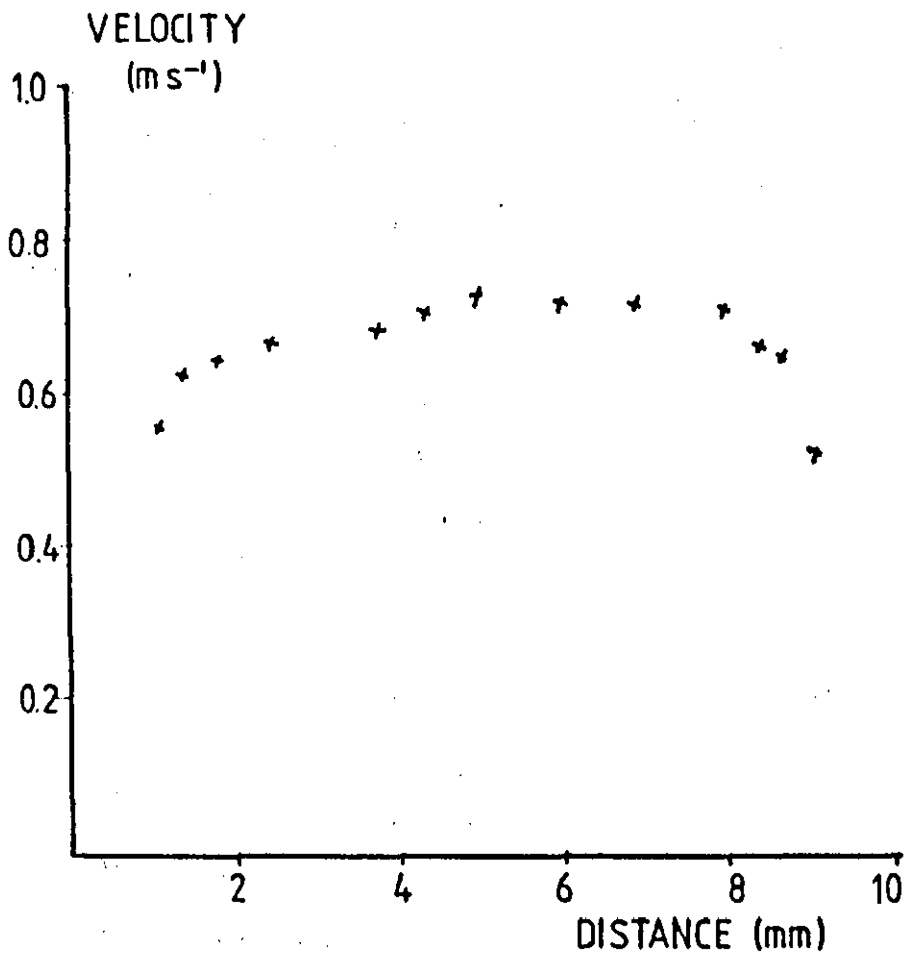
It is also essential to have a fast frame speeed, so that accurate particle trajectories can be found. The greater the magnification, the greater the frame speed that



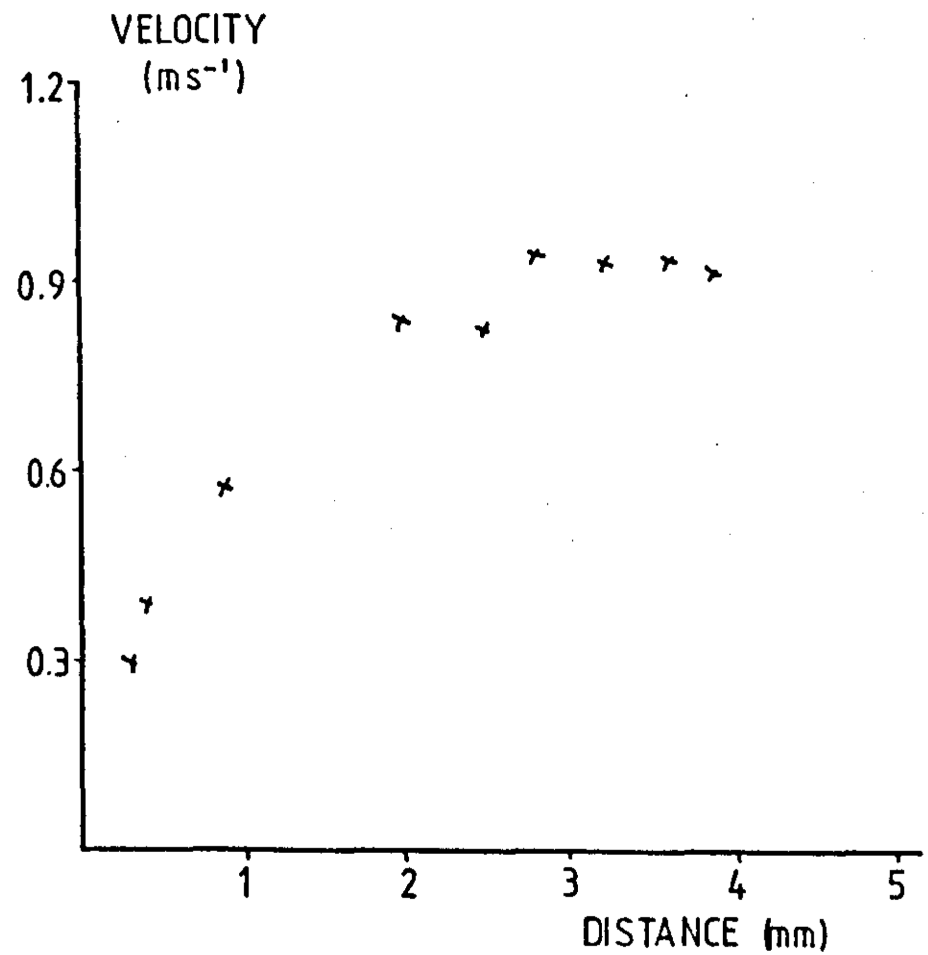
GRAPH 9.1 Laser anemometer velocity profile across the filtration cell. (Re = 13,300)



GRAPH 9.2 Laser anemometer velocity profile across the channel. (Re = 20,000)



GRAPH 9.3 Laser anemometer velocity profile. (Re = 6700)



GRAPH 9.4 Velocity profile as measured by high speed photography. (Re = 11,400)

is required. This is because the greater the magnification, the smaller the actual area that is covered by the camera, and the shorter the time that a particle takes to traverse that area. Thus a faster frame speed is needed, so that particles can be traced accurately. This fast frame speed dictates the intensity of light that must be shone into the filming zone, and in these experiments a high intensity light source had to be focused so as to adequately light the filming zone.

It was found necessary in these experiments to use a frame speed of between 4000 and 6000 frames per second. With this frame speed and using the most intense light source available (7.5 kW), it was possible to get an image to film ratio of 1.25 to 1. This means that the minimum particle size that could be resolved satisfactorily was between 20 and 40 microns in diameter. A C dust (Feldspar) of this size was used as a seed in the circulating fluid. For optical reasons, the glass rig (Rib B) was constructed and used for these experiments. Two alternative filming arrangements are shown in Fig 9.2.

RESULTS.

One graph (Graph 9.4) of the velocity profile obtained by the high speed photography is included. It was possible to get a velocity measurement within 0.25 mm of the membrane by this method. Closer than this the errors were too great, chiefly because of the lack of resolution of the membrane surface and the particles.

DISCUSSION

Particles over 10 microns in diameter are very unlikely to block a membrane with a nominal pore size of 1.2 microns. The critical part of the velocity will be within 10 particle diameters of the membrane surface, and thus this will be within 100 microns of the surface. Since, in this study it was not possible to measure velocities within this distance from the membrane surface, the data

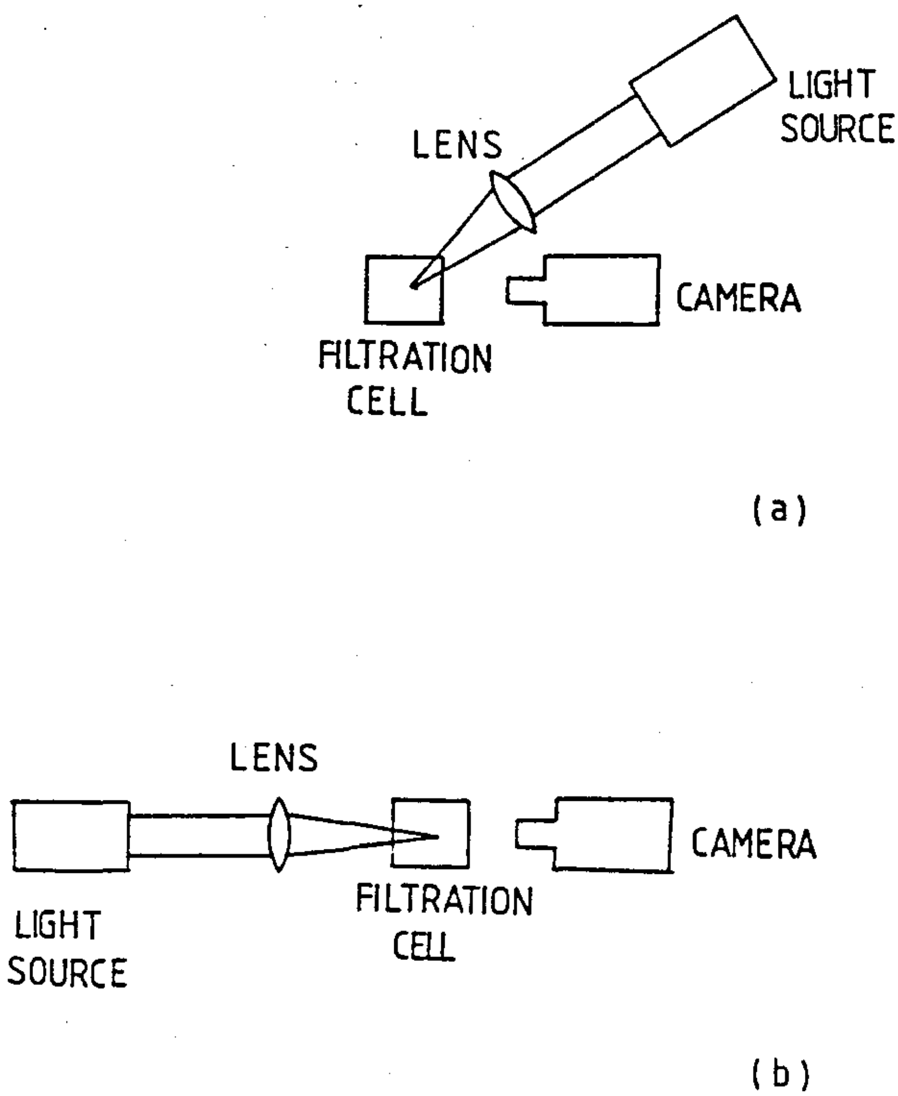


FIG 9.2 Diagram showing the arrangements for high speed photography. Fig 9.2(a) shows the forelit arrangement and Fig 9.2(b) the backlit configuration.

that were determined are not definitive, but can only assist as an indication of the velcocity profile near the membrane.

The reason the laser anemometer was unable to provide better data was that the true signal was drowned in the excess light that was reflected from the membrane surface. When the measurement zone was close to the membrane, the beam became partially obscured. Light due to reflection from the membrane surface progressively became greater than the light reflected from the tracer particles as the measurement zone was moved closer to the membrane. The "velocity" part of the signal received by the photon detector became more and more difficult to extract from the background signal, and consequently the velocity measured became less and less precise. There is also a possibility that the reflected light tended to distort the interference pattern in the measurement zone.

One way in which this problem might be overcome, is to reduce the diameter of the laser beams, by some lens arrangement. This would reduce the reflection of extraneous light, and enable measurements to be taken closer to the membrane surface.

The reason it was not possible to get better measurements from the high speed photography is a matter of resolution. It was not possible to resolve the membrane surface or the particles sufficiently well, to enable measurements to be taken closer to the membrane. If suitable lenses were available, it would be possible to increase the magnification. More light would then be needed, but this could be provided by focusing of a suitably intense source. It may not be necessary to increase the frame speed by a proportional amount, as the velocities will be somewhat lower near the membrane surface.

The results obtained are much as expected. All are for turbulent flow conditions, and the profiles exhibit the characteristic "flatness" of that regime. The two methods give results consistent with each other. It is apparent that the higher the velocity, the much greater the velocity gradient near the surface.

The velocity component towards the membrane (the filtrate velocity) was found not to be measureable. It is likely to be at least two, and probably three, orders of magnitude less than the component along the membrane surface. This would indicate that it will be very difficult indeed to determine this component of the velocity.

Measuring the velocity profile is only one way of determining the velocity profile: the other way is to calculate the velocity theoretically. Some approches were detailed in the Literature Survey. This might be the easier way of finding the profile very close to the membrane.

MEMBRANE TESTS

INTRODUCTION

The single most important factor in the viability of a crossflow filter is the choice of membrane. It is not only the permeability aspects of the membrane that should be considered. The mechanical aspects are of equal importance.

For this reason some tests on the stability and strength of a particular membrane were performed. The membrane was the one most frequently used for experiments reported in this thesis, Versapor. This membrane is a nylon-acrylonitrile copolymer, marketed by Gelman Sciences Ltd. It is described in the Membrane Survey (Section 12). Two tests are described, a swelling test and tensile test.

SWELLING TEST.

Nylon is a very strong plastic material but it swells in contact with water. To determine the extent of this swelling, a 75mm square sample of membrane was cut out and inserted in a beaker of water. Twenty four hours later, the sample was measured and it was therefore possible to determine the extent of any swelling. It was found that the membrane had swollen approximately 3mm or 4%. There was some evidence that this swelling was anisotropic, although the extent of this was relatively minor. Two different nominal pore size membranes were used for these tests, and there was no evidence of this influencing the extent of the swelling.

TENSILE TESTS.

Tensile tests give an indication of the mechanical properties of a material. A tensilemeter stretches a material at a constant, and predetermined rate. The force required for this stretching is measured and recorded on a chart recorder. It is possible to obtain the Ultimate

Tensile Strength of a material, and also the elongation for a particular applied tensile force from this method.

There are three factors which might influence the mechanical properties of this membrane. These are 1) whether the membrane is wet or dry, 2) the direction of the sample in relation to the membrane roll, and 3) the nominal pore size of the membrane. Two conditions of each of the factors were permutated, to give eight tests in all.

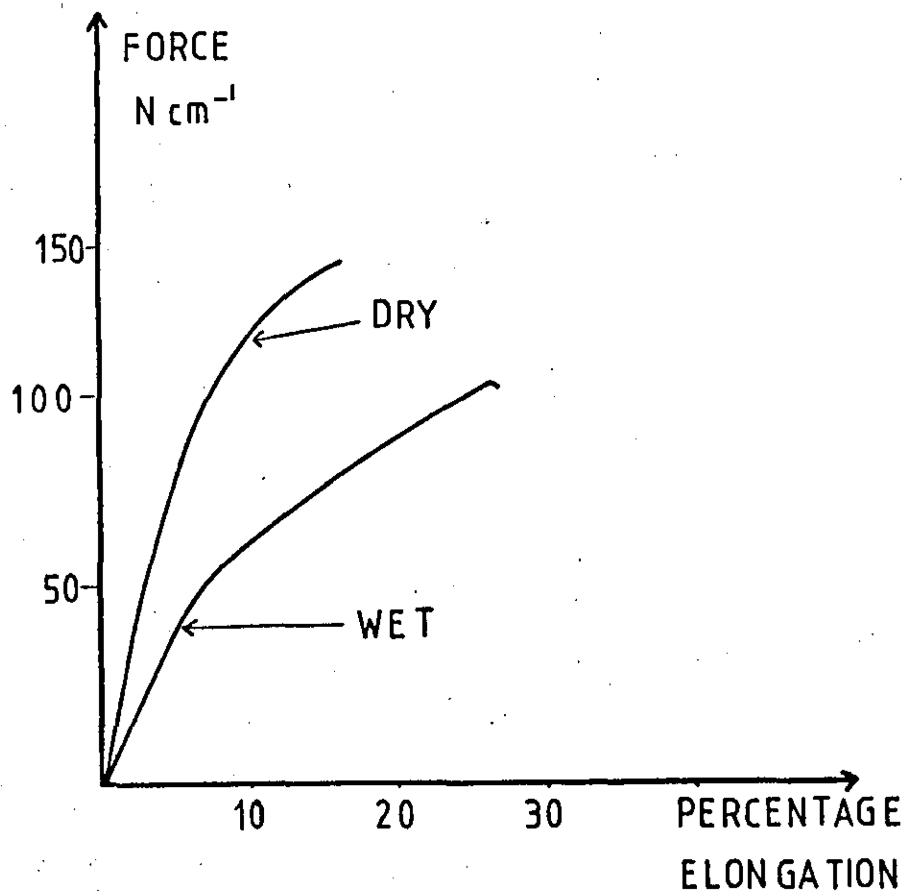
Two different nominal pore sizes were used (1.2 microns and 5.0 microns). For each pore size, samples were taken across and along the membrane roll. For each of these conditions one sample was tested dry and another was soaked in a beaker of water prior to testing. Finally, each test was repeated five times, to ensure a representative test.

Each sample was 1cm wide by 10cm long. When inserted in the tensilemeter, the distance between the jaws of the device was 50mm. The tensile tests were performed using a tensilemeter in the Institute of Polymer Technology, Loughborough University. The assistance of the Institute is gratefully acknowledged.

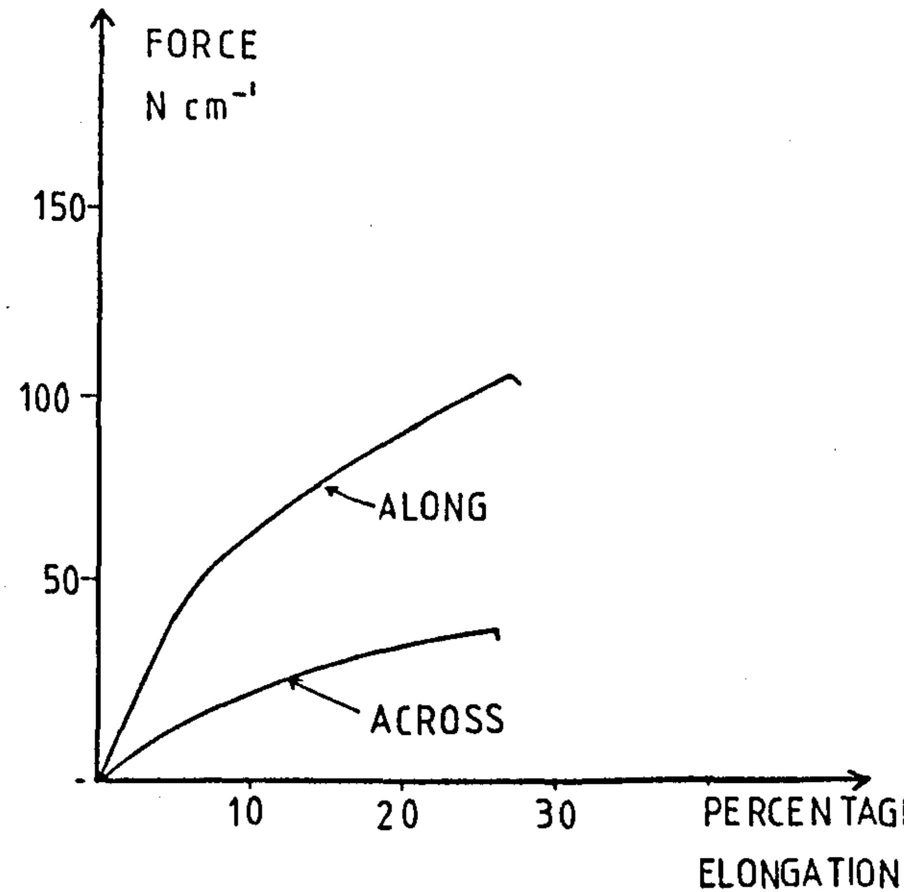
Three graphs of results are included. Graph 10.1 shows the effect of the wetness of the membrane on the tensile properties. Graph 10.2 demonstrates the extent to which the material is anisotropic and Graph 10.3 shows the influence of the nominal pore size of the membrane on the mechanical properties.

DISCUSSION.

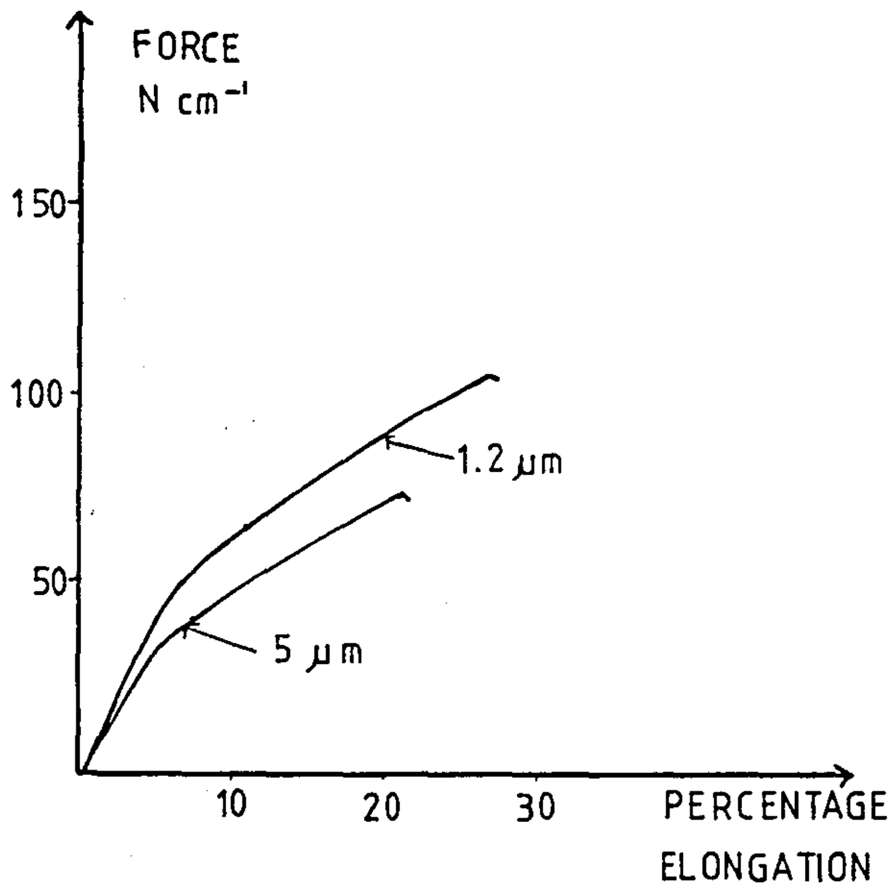
The swelling tests demonstrate the problem that one has with a nylon membrane. The membrane is usually inserted into the filter dry, and therefore when it is wetted it becomes loose. In Appendix 2, it is shown how a defection can be calculated from the percentage elongation. In a 5mm channel, the 4% swelling corresponds to a deflection in the centre of the channel of 0.6mm. This is not insignificant, and represents a major difficulty in the application of this membrane.



Graph 10.1 Tensile test curve for wet and dry membranes.



Graph 10.2 Tensile test curve for membrane taken along and across the roll.



Graph 10.3 Tensile test curve for membranes with nominal pore sizes of 1.2 and 5.0 microns.

An alternative to inserting the membrane dry is to wet the membrane prior to installation. If the membrane should become dry during operation, then a tension will be left in the membrane which will obviously be deleterious to it. If this course is followed, care should be taken to prevent the membrane from drying out.

The interpretation of the tensile tests is relatively straightforward. Graph 10.1 shows the effect of the wetness on its tensile characteristics. The wet membrane has an Ultimate Tensile Strength approximately 30% lower than the dry membrane. The elongation at break is nearly double for the wet membrane, and the modulus of elasticity is obviously far higher in the wet case. Obviously the membrane will be wet when in use, so the data that should be applied is that relating to the wet membrane.

Versapor is sold in roll form, each roll being 4.27 metres long by 254mm wide. It was suspected that the material might be anisotropic. To test this hypothesis, samples were taken along and across the roll. Graph 10.2 shows that the material is anisotropic, and is considerably stronger in the direction along the membrane roll than across the roll. This is almost certainly due to the method of manufacture. The length of the channels exceeds the width of material, and thus the material is clamped across the weaker direction.

Graph 10.3 shows the effect of the nominal pore size of the membrane on tensile properties. The larger pore size material is slightly weaker, and more easily stretched than the finer material. The smaller pore size material has the fibres closer together. This is what leads to its finer pore size. The closeness of the fibres gives it its increased strength.

From the tensile data it is possible to predict the deflection one would observe in an unsupported membrane. This is applicable during the backflushing of the filter. An analysis of this type is included in Appendix 2.

PROTOTYPE DESIGN

INTRODUCTION

This chapter presents the design of a prototype crossflow filter. It is expected that when the prototype is constructed it will undergo field trials. This chapter presents the design and explains the rationale of the design, and it is hoped to show that what might appear to be arbitrary specifications, are part of a design philosophy.

There are a number of principles that apply to a prototype design, that do not apply to the design of a product in full-scale production. One of the most important of these is adaptability. This principle, if embodied in the design, has two consequences; a) a number of particular configurations may be tried and the optimum solution can be determined, and b) if a feature of the design is incorrect, it can be modified after the prototype has been constructed.

The prototype is to be used both in laboratory experiments, and also in field tests under conditions that would be encountered if the filter was developed commercially. Thus, it is necessary for the filter to be both as self-contained and as portable as possible. The design should be such that the filter can be easily assembled and disassembled. It is assumed that the filter will be modified somewhat after construction, and thus it is essential that the prototype be as robust as possible. All these factors have been considered in this design.

OPERATING CYCLE

As indicated in the backflushing chapter, the most effective backflushing strategy consists of frequent, short, high pressure backflushes. The optimum backflushing fluid is compressed air. The minimum duration for a backflush is the crossflow velocity divided by the channel length. Since a velocity of 5 m/s and a channel length of approximately 1 m are envisaged, this minimum time is only about 0.2 seconds. This is obviously too short for practical reasons, (the valves cannot work that quickly), so the following backflush specifications were made.

Fluid	Compressed air
Duration	1-2 seconds
Frequency	Every 1-2 minutes
Pressure	272 kPa.

Table of Backflushing Specifications

FLOW DIAGRAM.

A schematic diagram is given in Fig 11.1 which shows the proposed flow arrangement. The filter will be controlled by an ITT controller. Full details of the instrumentation are given later in this chapter.

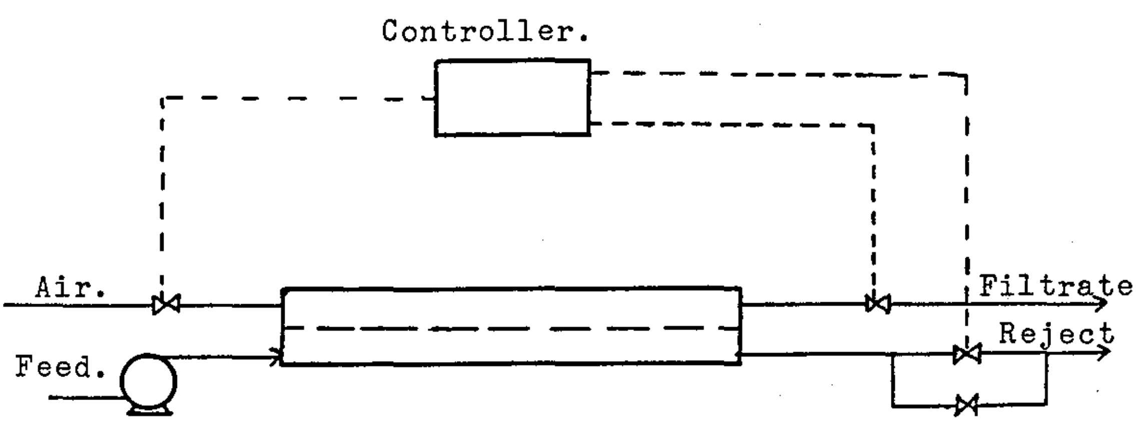


Fig 11.1 Flow Diagram of prototype filter.

MEMBRANE

During each backflush cycle, the membrane is subjected to very high mechanical stresses. It is possible to support the membrane during the filtration mode, but not during the backflush mode. Thus the membrane must be sufficiently robust to be able to withstand the full backflush pressure unsupported. The repeated nature of backflushing leads to frequent stress reversals, placing an additional load on the membrane. Therefore the choice of membrane is considered to be the single most important factor in the operation of a successful filter.

At this stage, the membrane has not been specified. An indication of the available range of membranes is given in the Membrane Survey (Section 12). A porous plastic product, "VYON F" produced by PORVAIR LTD. is specified as the membrane support material. This material has been used throughout this course of experiments, and appears to be very suitable for crossflow filtration. When the prototype is constructed, a number of different membranes will undergo tests in the filter. Thus a suitable membrane should be determined in this way. In the design, special attention has been given to allowing the easy instalment of many different membranes.

MATERIALS OF CONSTRUCTION

Originally it was intended to construct the filter out of metal, probably either stainless steel or aluminium. However there are several reasons why a polymeric material is more suitable. Firstly there are no corrosion problems. Secondly it is easier to work with and bond together. Finally it is both lighter in mass, and less expensive.

With respect to strength and chemical inertness, polypropylene is probably the best material. However it is difficult to work with, and is very difficult to bond to other materials, or to itself. Poly Vinyl Chloride (PVC) is a readily available material, with good strength properties. It is relatively inexpensive, and is easily machineable. It was therefore decided to specify PVC as the material of construction.

For the piping, either ABS or PVC would be suitable, but these materials are not very robust. Galvanised mild steel will corrode, but the corrosion will be limited by the galvanising, and the advantage of the materials robustness led to its adoption for all the piping. For the bolts and clamps, it was decided that stainless steel was too expensive, and a coated mild steel would suffice.

CHANNEL GEOMETRY.

At the start of this project, it was envisaged that any filter would consist of long narrow channels. It is a critical part of the design to optimise the channel geometry.

For the design of this prototype, a pump was available, and it seemed to be suitable for the flowrates and pressures envisaged. Normally the filter would be designed around pre-determined specifications, and a pump would be specified at the completion of the design which fulfilled the requirements of the application. In this case the opposite occured. The pump was specified at the outset, and the filter was designed around the operating curve of this pump.

In order to evaluate the effect of channel geometry, a computer programme was developed. A listing of this program is give in Appendix 3. The input parameters for this programme were:

1/ Channel Geometry (Shape and dimensions of the channels)

2/ Number of channels.

3/ Filtrate Flux.

4/ Pump's operating curve.

The following values were specified in the programme.

1/ Reject flowrate was 20% of the feed flowrate. (Thus 80% of the feed is filtered)

2/ Pressure at the reject end of the channels to be 100 kPa (this seems to be a reasonable value so as to provide a worthwhile filtrate flux at the end of the channel).

From this information it is possible to calculate the following:

1/ Effective channel length required (so as to satisfy reject criteria)

2/ Crossflow and filtrate flowrates.

3/ Entry, exit, and mean crossflow velocities.

4/ Pressure drop along the channel, and the pressure of the feed at the channel entrance.

The most restrictive assumption of the programme concerns the filtrate flux. It was assumed to be constant over all flowrates, pressures, crossflow velocities, and channel geometries. This is obviously an extreme simplification. However there are three factors which mitigate the effects of this assumption. Firstly, average values of the filtrate flux that have been obtained in experiments described in this thesis were used for the analysis. Secondly, each iteration was carried out with two different filtrate flux values; one higher than that normally obtained, the other lower than that obtained previously (for example, in Graph 5.7). Finally, it has been shown in Section. 6 that there is no evidence to suggest that the channel geometry influences the filtrate flux. Since it is the object of this program to optimise the channel geometry, this constant flux assumption is not as restrictive as it might first appear.

A flow diagram of the programme is given in Fig 11.2. Using this programme it is possible to calculate the effect of the various parameters. Some typical print-out from the program is given in Appendix 4. The effect of the various input parameters are discussed below.

Channel Shape.

Four different channel cross-sectional shapes were used in the programme to compare the effect of shape by using shapes of equal area. The four were; a) square, b) rectangle (2x1), c) equilateral Triangle, and d) right-angled triangle. In each case the membrane was considered to be along the longest side.

The shape which required the shortest channel length was the right angled triangle. This is because this shape has the largest membrane area to crossflow area ratio (for a constant length). Triangles have the additional advantage of being able to be "stacked", so minimising the overall size of each plate. Also triangles tend to have higher rigidity than do rectangles (which can fold under stress). Therefore, because the right-angled triangle will lead to the most compact filter, this shape is specified for the crossflow channel cross-section.

Channel Dimensions.

The larger the channel, the longer the channel length required to achieve the 20% reject ratio specified. It was decided that a channel length of 1m is about optimum. If the channels were much shorter than this, the channel dimensions become so small, and so many are required, that the cost is exorbitant. Much longer than this, the fabrication becomes too difficult and unwieldy.

Number of channels

In the region where the pump's operating curve is flat (at low flowrates), the number of channels does not effect

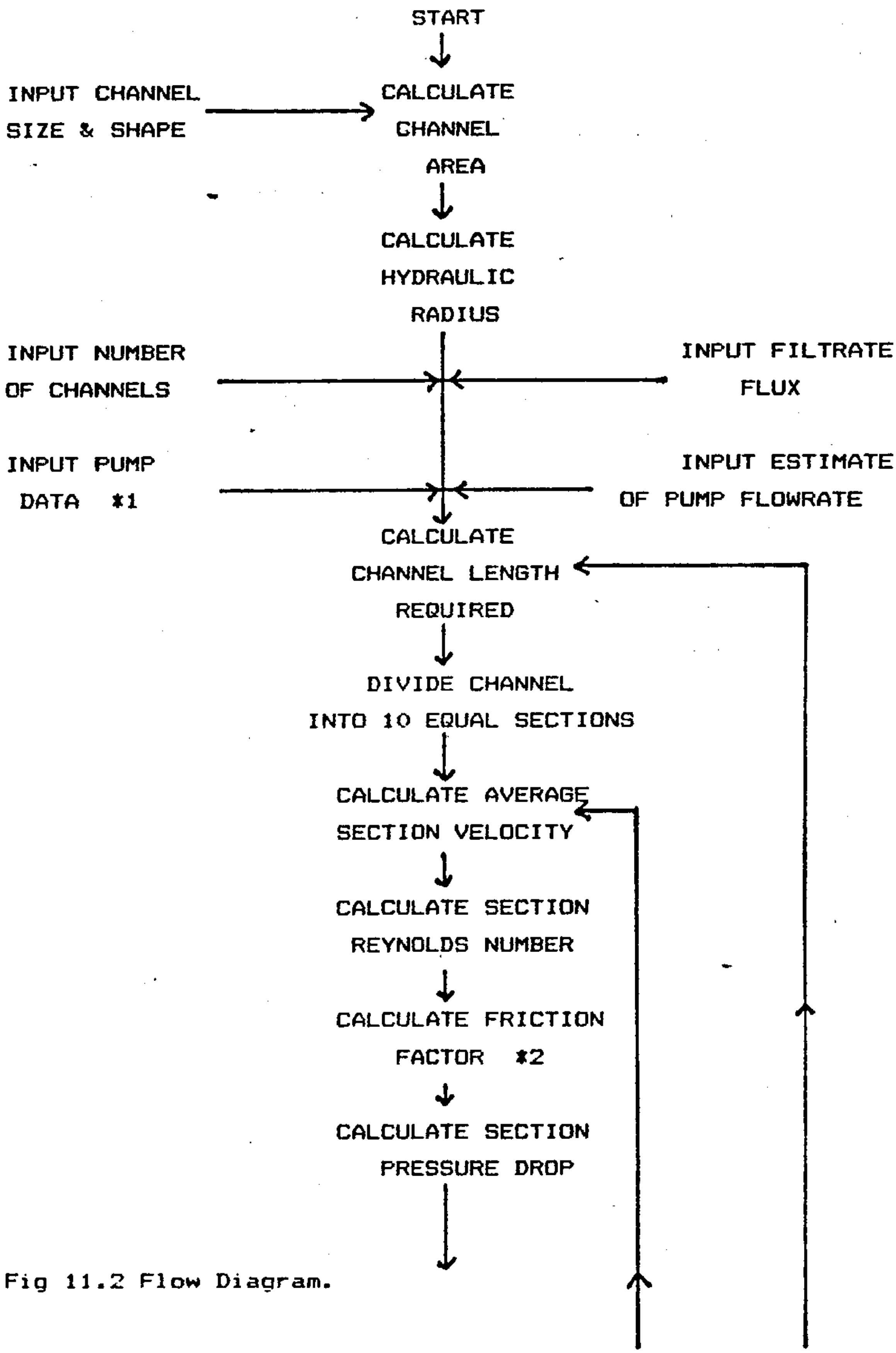


Fig 11.2 Flow Diagram.

CALCULATE VELOCITY
OF NEXT SECTION

ITERATE FOR
ALL 10 SECTIONS

CALCULATE CHANNEL
PRESSURE DROP

CALCULATE PUMP PRESSURE
AT THAT FLOWRATE

CALCULATE
NEW
FLOWRATE

ARE PRESSURE DROP
AND PUMP PRESSURE EQUAL

no

yes

PRINT OUT RESULTS

STOP

NOTES

*1. The pump's operating curve was divided into six sections, and linear interpolation was used within each section. This gives a reasonably accurate value for the pressure delivered at each flowrate.

*2. The Fanning friction factor was calculated by the Nikuradse equation.

$$1/ f = 4*\log(Re* f) - 0.4$$

This iterative procedure was used with a 1% difference as a convergence criteria. The formula is applicable for smooth pipe, and an arbitrary 20% was used to account for surface roughness.

Fig 11.2 Flow Diagram.

the required channel length or pressure drop. At the high flowrate end of the curve, the pressure drop, the crossflow velocity and the required channel length decrease, with increasing numbers of channels. Providing the velocities remain high enough, and the pressure delivered by the pump is sufficient, it is optimal to operate at the highest number of channels. This maximises the production of the filtrate for pumping energy required.

Filtrate flux.

Two values were used in this analysis. The lesser value is below that obtained in Graph 5.7 and the larger was similar to that obtained at the commencement of the runs described in Graphs 6.1 and 6.2. These values influence the length required and it seems reasonable to design towards the lower flux value (which was maintained for over 30 minutes with no evidence of decline). In doing this, the design is conservative, and the flowrates that are aimed at should be achieved.

Taking all the above factors into consideration, it was decided to specify the following geometry for the prototype filter.

CHANNEL SHAPE	Right angled triangle
CHANNEL HEIGHT	2.5 mm
CHANNEL WIDTH	5.0 mm
NUMBER OF CHANNELS	200
CHANNEL LENGTH	1 m

The following values are expected in the operation of the filter.

VELOCITY	
(at channel entrance)	6.1 m/s
(at channel exit)	1.2 m/s
(geometric mean)	3.7 m/s
PRESSURE DROP	
(along channel length)	155 kPa
PUMP FLOWRATE	7.57 m/s

FILTRATE CHANNELS

The previous tables specified the crossflow channels and their geometry. These channels are milled on both sides of a set of plates which for convenience are called crossflow, or feed, plates. The membrane and the membrane support then separate each of these plates from another set of plates. These plates carry the filtrate away and are referred to as filtrate plates. Thus a sandwich is constructed of filtrate plate, membrane and support, crossflow plate, membrane and support, filtrate plate etc.

The channel geometry on the crossflow plate has been optimised. Each crossflow channel must line up with a filtrate channel to allow a free passage of filtrate. Thus the width of each filtrate channel has to be 5 mm, the same as the crossflow channels. To maximise the pressure difference across the membrane, it is necessary to maintain the pressure on the filtrate side of the membrane as near atmospheric as possible. Thus it is desirable to minimise the pressure drop along the filtrate channels.

The cross-sectional area of the crossflow channel is 6.25 square mm. If the filtrate channels are made square in cross-section, then the cross-sectional area of each filtrate channel is 25 sq mm. The velocity is proportional to the inverse of the area, so the velocity in the filtrate channel is 4 times less (25/6.25). The pressure drop is proportional to the sqaure of the velocity, so the pressure drop along the filtrate channel will be 16 times smaller than that in the crossflow channel. This pressure drop is not excessive, so a filtrate channel 5mm * 5 mm is specified.

LAYOUT OF PLATES.

Each channel is 5 mm wide. It seems reasonable to suggest that 25 channels be milled on each side of each plate. This gives a minimum width of 125 mm. Assuming that the channels need to be spaced 2.5 mm from each other, this leads to a width of 175 mm. Then appromimately 50 mm will be required on each edge of the plate for sealing and gasketing. This leads to a total plate width of 275 mm which seems to be a reaonable width. If the width were to exceed this, then the device would be very hard to handle, while a narrower filter would have a disproportionate amount of material at the edges used for sealing. Thus 25 channels per side of each plate seems to be give a reaonable layout.

The number of crossflow plates required is 200/(25*2) which is 4. One extra filtrate plate is required as the top and bottom plates have channels on only one side.

MECHANICAL DESIGN OF PLATES

Design of crossflow plate.

The crossflow plate will consist of channels on each side of the plate, the apex of the channel on one side of the plate being directly between two channels on the other side of the plate. This is shown in Fig 11.3

Fig 11.3 Expected cross-section of crossflow plate.

During the filtration mode there is no pressure difference across the webbing, because all the channels are at the same pressure. However during a backflush, the plate is in compression. The mode of failure for the plate, will be shear across the minimum cross section.

Force due to = Pressure * area
compression

Pressure = 2.5 * Backflush pressure. (the 2.5 is
= 2.5 * 544 kPa a safety
= 1360 kPa factor)

Area = Width * length
= 5 mm * 1 m
= 0.005

Force = 1.360E6 * 5E-3
= 6800 N for each channel.

Cross-section reqd.= Force/ (Max Shear Stress)

The maximum shear stress for PVC is 5500 psi (7)
37.4 kPa

Cross-section reqd = 6.8E3 / 37.4E6
= 0.18E-3 sq m

Since the channel is 1 m long, the minimum thickness of the webbing is 0.2 mm. This is far too thin for practical purposes, and any realistic design will have a thickness far greater then this.

Another mode of failure will be the porous support failing due to compression. This decides the required flat distance channels. If the channels are mounted too close together, then the Vyon membrane support will fracture.

Area = Force / Max permissable stress.

The maximum permissable compressive stress of the Vyon is not available. However it is certain to exceed the max tensile stress which from the Vyon catalogue is 16 MPa for 4.75 mm for Vyon F. Thus to design to this value will be conservative.

Area = 6.8E3/16E6
= 0.425E-3 sq m.

Since the channel length is 1m, the minimum flat width required is 0.5 mm

The end result of the above calculations is that the minimum thickness of the plates will not be dictated by stress calculations, but by ease of handling. PVC is a very flexible material, and is very floppy in sheet form. Thus it is necessary to have it fairly thick to prevent this flexibility from being excessive. If the sheet is too thin, assembly and machining will be very difficult. It was decided to manufacture the crossflow plate out of PVC 12 mm thick. Although this is far too thick for production filters, it lends robustness to the prototype. It was also decided to machine the channels at 7.5 mm centres, leading to a 2.5 mm wide flat between the channels.

Design of Filtrate Plate

In order to align the channels in both the filtrate and crossflow plates, the thicknesses in the filtrate plate must be the same. Since the pressures in the filtrate plate are less than those in the crossflow plate (normal filtration pressure as opposed to backflushing pressure), the crossflow plate will fail prior to the filtrate plate.

Thus the thickness of the filtrate plate is decided not by stress calcualtions, but by flexibility considerations prior to assembly. Obviously the filtrate plate must exceed 10 mm to enable some material to separate the channels, and for the plate to remain integral. If the plate is milled out of a sheet 18 mm thick, the minimum cross-sectional thickness is 8 mm. This is sufficiently thick to enable the plate to be easily handled.

SEALING AND GASKETING.

The design as developed so far, consists of interspersed sheets of impermeable PVC and porous Vyon separated by the membrane. In the centre of the plates this aspect is essential. Towards the edges of each plate, it is necessary for all the material to be impermeable, so that the filter is sealed. The obvious way of doing this is to enclose the Vyon sheets in an envelope of PVC, so that the centre of the sheets are permeable and the edges impermeable. This method has the difficult problem of sealing the PVC to the Vyon, over a very narrow cross-section. Thus the two sheets must be cut to very fine tolerances, and this leads to expensive machining.

An alternative method is to make the Vyon impermeable at its edges. This has been done in the laboratory on a small sample, immersing the porous material in a liquid neoprene rubber compound. If this can be done on a larger scale, and no difficulties are forseen, then the Vyon will both support the membrane, and also assist in sealing the filter.

There are several advantages in doing this. Firstly, there is no need to join the plastics to each other. Secondly wastage is minimised. A layer of excess rubber is built up on the surface on the Vyon during the sealing operation. This surplus rubber will assist in three ways; a) it will eliminate the need for a gasket, b) it will assist in absorbing irregularities in the surface, and c) it will help to clamp and seal the edges of the membrane.

It should be stressed that the normal direction of flow through the Vyon is across the minimum thickness. It is not very permeable, but not impermeable in the direction which would lead to the filter leaking. Thus this method does not subject the Vyon to the full flow, but merely seals off any of the tortuous paths that might lead to leakage. This method of sealing is recommended in the construction of the filter.

CLAMPING

There are two obvious ways of clamping the filter. Bolts can be inserted through the filter at frequent intervals. This also ensures that the plates are aligned in the correct manner. Clamps can be used externally to hold the filter together. The prototype will use a combination of these. Bolts will be used, and as well, clamps will be used at the end of the filter where the fluids enter and exit the device.

The overall length of the filter is 1m, plus what is required at each end of the filter. If 150 mm is required at each end, to allow the pipes to be attached to the filter, the overall length will be 1.3 m. Assume that bolts are spaced every 50 mm along the edges, with the first bolt being 25 mm from each end. This seems to be a reasonable distance; if the bolts were closer, too many would be required and assembly and disassembly would be extremely tedious, and if the bolts were spaced at greater intervals, the filter would distort under pressure and leak. The number of bolts required therefore is

1.3 / 0.05 = 26 bolts per side.

The toal force exerted on the bolts is the pressure times the area.

Area = Area per channel * No. of channels per side of each plate

= 1 * 0.005 * 25

= 0.125 sq m.

Force = Pressure * Area

= 1.36E6 * 0.125

= 170 kN

From BS.4882 (bolting specifications) the maximum allowable tensile strength for mild steel bolts is 400 MPa.

Area of = Force / Allowable stress in bolting material.

Bolts reqd. = 170E3 / 400E6

= 0.425E-3 sq m.

Assuming the force is evenly distributed over all the bolts along the side opposite the channels.

Area per = Total Area / No of bolts opposite channels
Bolt = 0.425E-3 / (2*1/0.05)
= 10.6E-6 sq m.
= 10.6 sq mm

This leads to bolts about 4 mm in diameter. However this is too small for practical reasons. According to BS.4882 an M10 bolt has a stressed area of 58 sq mm. This is over 5 times in excess of the minimum required, and will be specified for the prototype.

This analysis is strictly only valid for that section of the filter adjacent to the channels. However the pressures are the same at the ends of the filter, and since the area exposed to this pressure is similar, and certainly not 5 times larger, the same bolt sizes will suffice. The bolts will also be placed across the width of the filter, with the same spacing, and these will assist in sharing the load at the fluid entrances and exits.

An auxiliary clamp is provided at each end of the filter. This consists of two lenghts of mild steel bar (25 mm * 12.5 mm) which are placed acoss the top and bottom of the filter. The bars are then joined by an M10 threaded fastener. This external clamp will assist in sealing the filter near the ends.

PIPE SIZING

The pump outlet has a size of 1 3/8" (35 mm). This is very small for the size of flow envisaged. A table is given below which shows the pressure drop dependance on the pipe size for the flowrate required. This particular table is for the feed flowrate, but the filtrate flowrate is only 20% lower, so the table is broadly applicable to that flowrate as well.

PIPE SIZE (mm)	VELOCITY (m/s)	PRESSURE DROP (kPa per metre length)
35	7.9	16
38	6.6	10.4
44.5	4.5	4.8
51	3.7	2.5
63.5	2.39	0.8

By inspection of this table, a pipe diameter in excess of 50 mm will not lead to excessive pressure drop. Assuming between 5 and 10 metres of pipe length are required for the feed and filtrate pipes, the total pressure drop will be less than 25 kPa, which is less than 20% of the pressure drop along the channels. A standard size for which valves and other fittings is readily available is 2" BSP. This will be specified as the pipe size for the feed, filtrate, and reject. For the air line, a size of 1/2" BSP is specified. This should be sufficient, but as no data on the air flowrates are available, this pipe can not be sized in the usual way.

FLOW ARRANGEMENT

One of the most difficult aspects of the design of the prototype, was how to get the water into and out of the filter. There are four flows involved; feed in, reject out, filtrate out, and compressed air out.

The method that will be used, utilizes the impermeable layer of Vyon around the edges of the filter. Four pipes come into the filter, one at each corner. These four pipes are connected to conduits, which have been drilled through all the plates. These conduits are sealed from each other, and from the channels, by the impermeable Vyon. A wedge shaped section is then cut out to connect each conduit to the appropriate set of channels, the feed and reject conduits being connected to the crossflow channels, etc. The conduits have been placed in such a position, that when the wedge shape sections are cut out, the conduits remain isolated from each other, and thus the channels are only connected to the appropriate conduits, and isolated from the other fluid-carrying conduits.

This flow arrangement can be seen more clearly in the drawings for the prototype which are included in Appendix 5.

INSTRUMENTATION.

An ITT controller will control the operating cycle. To enable this to operate correctly, a reasonably complex circuit is required. This is shown in the line diagram Fig 11.4.

As can be seen from the diagram, there are four valves required. Details of these valves are:

A/ This is a 1/2" BSP solenoid valve capable of switching compressed air at a pressure of 680 kPa.

B/ This is a 2" BSP solenoid valve capable of switching water at a pressure of 680 kPa.

C/ Same as B above.

D/ This is a 2" BSP gate valve, used to set and tune the reject pressure.

There are three pressure gauges required as follows:

1/ 0 to 1 MPa air pressure gauge used to register the line air pressure.

2/ 0 to 1 MPa water pressure gauge, used to determine the pressure delivered by the pump.

3/ 0 to 1 MPa water pressure gauge used to determine the pressure of the reject

Gauge 3 can be tuned by adjusting Valve D. The difference between the pressures measured by gauges 2 and 3 gives the pressure drop along the channel. The mean of the pressures determined by gauges 2 and 3 gives the mean pressure difference across the membrane.

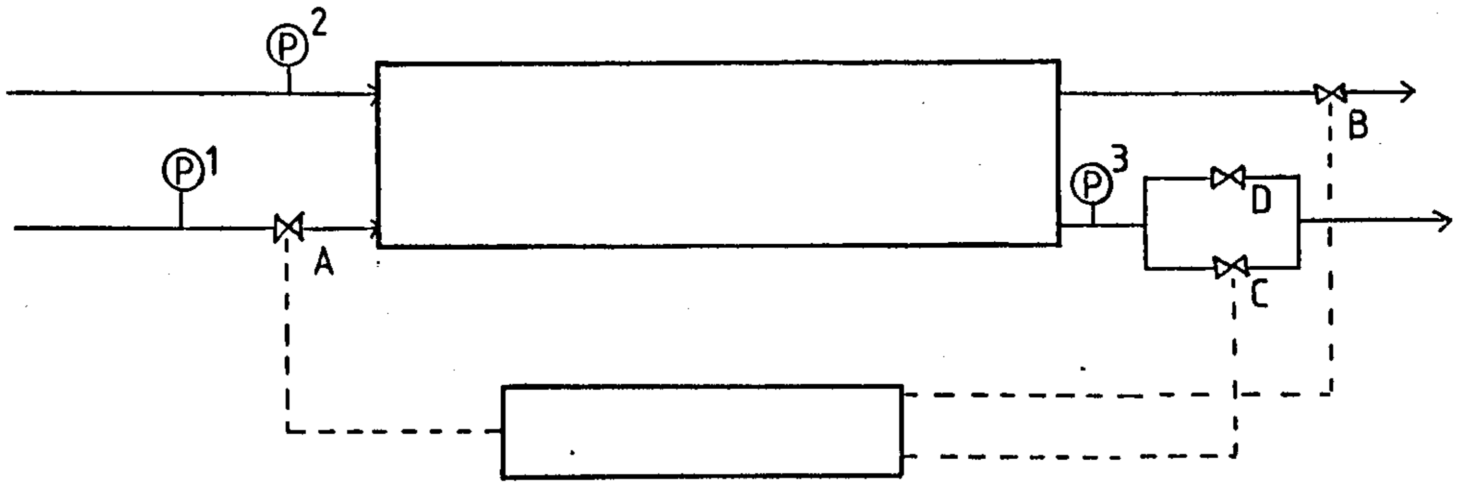


Fig 11.4 Flow diagram of prototype filter.

Although not shown on the line diagram, a flow measurement device will be needed in the filtrate line, to record the production of filtrate. An orifice plate and manometer is probably the most suitable method, although some continuous method of recording would be an advantage.

OPERATING PROCEDURE

The sequence in which the valves are switched on and off is very important to the overall operation of the filter. A schematic diagram of the valve sequence to initiate a backflush, and to return to normal filtration mode is given in Fig 11.5.

The backflush cycle can be divided into five sections (A-E on the diagram). Each section performs a specific function, details of which are given below.

Section A: These valve settings are for the filter in the normal filtration mode.

Section B: This valve opening lowers the pressure on the crossflow side of the filter, prior to the introduction of the compressed air.

Section C: The opening of the air valve removes all the filtrate before the backflush commences. This maximises the yield of filtrate, while ensuring that the backflushing fluid is only compressed air.

Section D: The closing of the filtrate valve raises the pressure on the filtrate side of the filter, initiating the actual backflush condition.

Section E: In this mode the filter resumes filtration prior to the increase in pressure caused by the closing of valve C. After the closure of valve C the valves settings revert to those described in section A.

The valve sequences have been arranged to maximise the filtration rate and the effectiveness of backflushing,

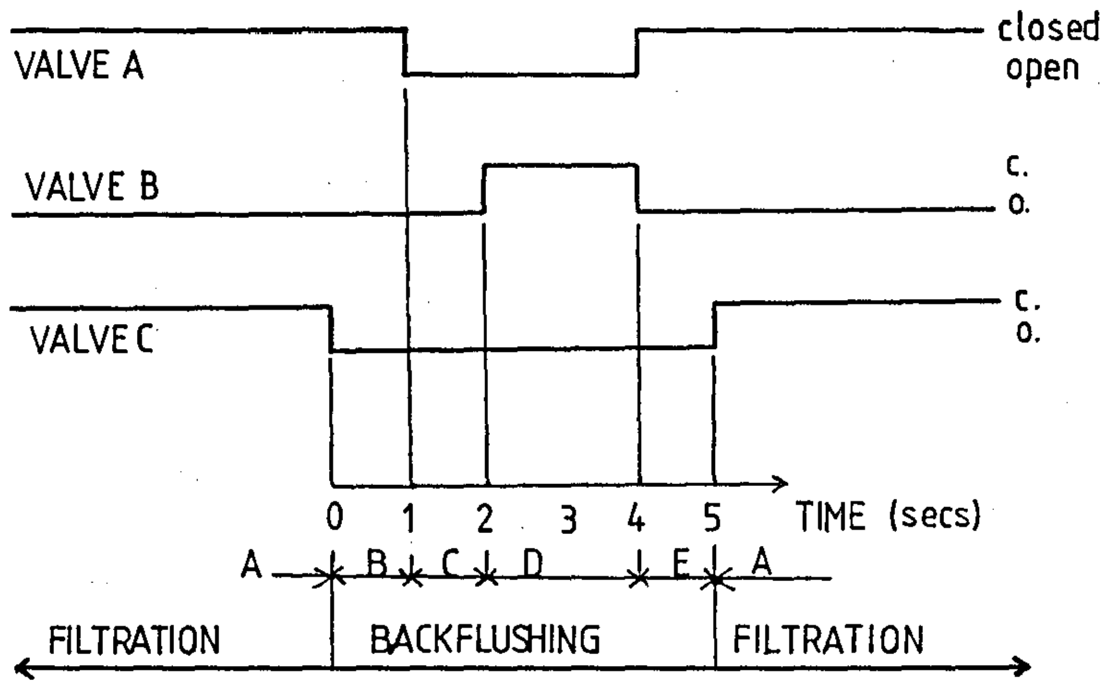


Fig 11.5 Valve sequence to initiate a backflush.

while minimising the downtime of the filter.

TANKS.

Obviously, to enable the filter to be tested in the laboratory, a tank must be provided. Although the flowrates are high, the residence times in the filter are low, so a tank of 200 litres will suffice. Both the reject and filtrate must be directed straight back to the tank for this strategy to work.

COMPRESSED AIR

In the laboratory there is a ready supply of compressed air available. However, it might be advantageous for the filter to be self-contained. Thus the filter could require its own air supply. One way of doing this, is to include a compressed air cylinder with the filter. It is envisaged that a small cylinder would last for a considerable number of backflushes.

One aspect of safety must be considered. There are very considerable pressures in an air cylinder and it is essential to include a safety release valve in the air line, just past the throttling valve, to prevent an accidental increase in pressure within the filter. Such an increase could well cause a catastrophic failure of the filter.

FRAME.

When the prototype is assembled, it will be necessary to install it on a suitable frame. This frame has not been designed, but it will consist of a baseplate which will hold the pump, and some suitable lattice structure to which the filter, and piping can be attached. The filter could be installed either horizontally or vertically in this frame. Although there seem to be no advantages from a filtration viewpoint (the particles are too small for gravity forces to be significant), the vertical arrangement might result in a more compact prototype.

DRAWING.

Detailed drawings of the filter are included in Appendix 5.

MEMBRANE SURVEY

The eventual success or failure of crossflow filtration will depend mainly on the availability of a suitable membrane. An indication of the membranes that are available will obviously be advantageous. For this reason a letter was circulated to a number of membrane manufacturers and filtration equipment suppliers. A copy of this letter is given in Appendix 3. This survey is not meant to be a definitive one, but to provide an indication of what membranes are available which are likely to be of use in crossflow filtration.

The first membranes were not artificially produced, but were naturally occuring. Indeed, many natural processes are dependent on these semi-permeable membranes. The first artificial membranes had extremely low permeabilties, and the low filtrate fluxes prevented membrane processes from gaining widespread adoption. The first breakthrough in membrane technology came when a very thin active layer was attached to a passive substrate. This lead to an order in magnitude increase in permeability, but the new membrane had as high a selectivity as the previous membrane. This new membrane encouraged people to state that membrane processes would become very popular, but this early optimism was ill-placed, and membranes did not continue to improve at this rate.

There are two main requirements of a membrane which is to be used in a crossflow filter. Firstly it must have a reasonably high permeability. Secondly it must have a high tensile strength to withstand repeated high pressure backflushing that are required for this particular application. These two criteria were stressed in the letter that was sent out to the membrane manufacturers, and these two criteria are foremost in the appraisal of the replies.

There are different membranes for different purposes. Reverse Osmosis membranes have very fine pore structure, while ultrafiltration membranes are much looser. Both these membranes are commonly manufactured out of cellulose acetate. Unfortunately this material, although quite strong under compression, is very weak under tension. This is not important for reverse osmosis or for ultrafiltration, because these processes do not involve backflushing. The membranes which might be suitable for crossflow filtration can be divided into two main categories: membranes that are constructed out of an homogenous material, and laminates with a substrate provided for strength. The latter must not only be of suitable strength, but the adhesion must also be strong.

The membrane that was used for all the experimental work in this thesis, was Versapor, manufactured by Gelman Sciences Ltd. This membrane is an acrylic copolymer in a non-woven nylon substrate and is available in a number of different pore sizes. Although it is a reasonably inexpensive membrane, its wet strength is less than its dry strength, and it is anisotropic (see Section 10). It is doubtful whether this membrane will survive the frequent stress reversals to which it will be subjected. Versapor superseded another membrane called Acropor. This latter membrane is an acrylic copolymer on a woven nylon backing. Gelman have indicated that this material might have more suitable tensile properties. They have also indicated that limited quantities of this membrane might be available.

Schleicher and Schull suggested that their membrane laminates might be suitable. These membranes are manufactured by bonding a thin film of cellulose nitrate to a cardboard support. These membranes are recommended for use in filter presses, and are said to be very robust. These membranes are available in the following nominal pore sizes; 0.2, 0.45, 0.6, and 3.0 microns.

Pall Group manufacture a number of membranes. Although negotiations are still continuing, they have indicated that their pure nylon, or polypropylene membranes might be suitable for this application. These membranes have a very open structure, which should give them very high initial fluxes. However no knowledge of their tensile properties is available at present.

Carl Freudenberg Ltd, manufacture a number of non-wovens for reverse osmosis. These materials are normally used as a support material for reverse osmosis. However their tensile properties appear to be suitable for crossflow filtration and an examination under a microscope reveals a pore size that might be suitable. Especially suitable is the material designate FO 2407, which consists of non-woven polyester fibres and polyethylene adhesive. These materials should be kept in consideration.

The Swiss Bolting Cloth Co. Ltd. manufacture very fine meshes out of woven fibres. The particular meshes they recommend were manufactured out of polyester fibre. These meshes have a very uniform structure, and from a manual examination, their tensile properties seem to be excellent. The main drawback of these materials is that the open area appears to be low, as shown by electron micrographs. It is certainly worthwhile to give these membranes further consideration.

Another approach is to use porous ceramics as the membrane. These ceramics are essentially depth filters, with a high dirt-holding capacity. The tensile strength of this material would probably be satisfactory, but it is doubtful if backflushing would be as effective with this material.

Battery separators have been suggested as being suitable as a membrane for this application. These materials are available in pore sizes from 0.05 microns, to 5 microns. However, according to Chloride Batteries, the tensile strength is only obtained at the lower pore sizes. These materials might be worth investigating as a last resort.

This survey is not meant to a definitive survey of all the membranes that are available, but to provide an indication of some of the membranes that are available, which might be successfully applied to crossflow filtration with backflushing.

DISCUSSION

Each experimental section in this thesis has a discussion of the results included. Thus it is not the function of this section to discuss the experimental results. This section will briefly summarise those individual discussions, to show what has been achieved, and to indicate what has still to be investigated.

The most important shortcoming of this thesis, is that the prototype design does not specify a particular membrane. Indeed there are no experiments reported in the thesis that show the effects of different membranes on filtrate flux, backflushing effectiveness, etc. The project cannot be considered successfully concluded until this information has been collected.

The tensile test described in the Membrane Tests (Section 10), provide a ready and accurate way of ascertaining a membrane's tensile properties. Thus it would seem desirable to repeat these tests for as many of the membranes described in the Membrane Survey (Section 12), as possible. However the tensile data obtained is for a new membrane. It could well be that repeated backflushing damages a membrane, and that a membrane which seems to be sufficiently strong, as indicated by the tensile test data, has only a short life. To test this some membrane duration tests will be needed. This would involve constructing a rig which will expose a membrane to repeated stress reversals. These last tests will take a long time, and a suggested strategy is to use tensile tests in the first analysis, and subject the promising membranes to the endurance tests.

Tensile strength is not the sole criteria for the selection of a membrane. There are two other criteria that are equally important: a) flux behaviour, and b) filtrate quality. The backflushing tests (Section 5) suggested that backflush effectiveness is dependent on backflushing pressure and independent of the duration of a backflush. It was also indicated in that section that 272 kPa, 1 second backflushes every 2 minutes gave a constant average filtrate flux over a sustained period. These conclusions only apply when the membrane was Versapor. It is quite possible that the effectiveness of backflushing is very membrane dependent. If a membrane was more of a depth filter, as opposed to a surface filter, backflushing would be expected to be less effective. Thus, the evaluation of a membrane must include some backflushing experiments. Perhaps the best way of doing this is to test each membrane for an hour long run, backflushing at 1 or 2 minute intervals. In this way both filtrate flux behaviour, and the effectiveness of backflushing can be ascertained in the same experiment.

The variation of filtrate quality with the choice of membrane must also be determined. In Section 7, no evidence of variation of the concentration of suspended particles with the pore size of the membrane was found. This anomalous result could have been due to the rather arbitrary way in which membrane manufacturers determine and specify pore sizes. The Malvern Submicron Laser which was used in Section 7 was not particularly satisfactory in determining filtrate quality, and the analysis method has to be improved. Another factor which has to be determined is the effect of the number of backflushes on the filtrate quality. It is possible that a membrane of sufficient tensile strength, and good flux-time characteristics (with backflushing), and good initial filtrate quality, is not a suitable membrane because the filtrate quality declines with the number of backflushes endured. Thus it would be advantageous to do a series of filtrate quality measurements at various times in a particular membrane's history.

The approach outlined above has assumed that the operating cylce involves high pressure backflushes, and that it is necessary to find a membrane that is suitable. The alternative is to specify a membrane, and then optimise the backflushing to suit the membrane. This is a feasible alternative, but the experiments in Section 5 indicate that high pressure backflushing is required. If a suitable membrane is not found, it may be necessary to examine other methods of preventing flux decline. One method that was examined (Section 5), and found not to be effective (with Versapor), was air scour. There are other possibilities, which are described in the following paragraphs.

Some of the methods that have been proposed for minimising flux decline rely on increased turbulence promotion. These include placing barriers just above the membrane surface, and the use of latex spheres in the feed. However there is considerable doubt as to whether increased turbulence has a beneficial or a negative effect. In reverse osmosis, increased turbulence could well assist what is essentially a mass transfer controlled process. In Section 8, evidence is given which suggests that for crossflow filtration, increased turbulence has a negative effect.

Ultrasonics have been used to minimise flux decline in membrane processes. Although this method has been successful (16,17), it has not been widely used. This is probably because of the capital cost of installing ultrasonics in any commercially sized equipment. Although it is probable that the application of ultrasonics would have a beneficial effect in crossflow filtration, it is also probable that the membrane life would be lessened by the vibration. It seems improbable that any membrane, which was not sufficiently strong to withstand repeated backflushing, would be able to withstand the vibration that would result from the application of ultrasonics for any practicable period.

Another method of flux decline minimisation which has been proposed, is the use of mechanical devices to scour the membrane. Tiller (29) devotes most of his paper to describing scrapers that clean the membrane periodically. These devices are more suited to thickening operations, as opposed to clarification. It was observed in Section 8 that large particles often disturbed smaller particles on the surface, and caused there re-entrainment. It is possible that the seeding of the feed to the filter with larger particles might have a beneficial effect. This method seems to be worth trying, although this author feels that any beneficial effect is likely to be so small as to be negligible.

Of all the methods of flux decline minimisation, backflushing seems to be the most effective. The alternatives seem unlikely to be as effective. The one method that might be a valid alternative is ultrasonics, and it appears that this is as likely to damage the membrane as is backflushing. Thus it would appear that the viability of crossflow filtration rests almost entirely upon finding a suitable membrane, which can withstand the repeated backflushing, as well as satisfying the other criteria.

This thesis has reported experiments that determine the effect of backflushing variables and differing channel geometries on filtrate flux. No experiments are reported that demonstrate the effect of crossflow velocity, pressure differential across the membrane, and feed concentration. As mentioned in the Literature Survey, the effect of crossflow velocity has been well reported. The velocities that are used in the prototype design are similar to those most commonly reported in the literature. These velocities have not been optimised, but evolved as the design proceeded. As it does not seem to be a critical variable, provided the flow is turbulent, the velocities which will be obtained in the prototype appear reasonable.

The effect of pressure difference on filtrate flux was also reported in the Literature Survey. There is considerable variation amongst authors as to the effect of pressure, and even whether an increased pressure is beneficial. The prototype will operate at pressures similar to those most commonly reported in the literature. Since the literature agrees that it is not a critical variable, the specified pressures are likely to give reasonable results.

The effect of feed concentration on filtrate flux is probably the most neglected variable in crossflow filtration research. For thickening operations this is likely to be a critical parameter. For clarification, the operator has little control over the feed concentration, so, provided the filter can meet its specification at the highest likely concentrations, the effect of this variable is unimportant.

The effect of these variables on filtrate quality has not been reported prior to this thesis. The indication is that the effects of pressure and crossflow filtration on filtrate quality are minimal. Provided this is true for all membranes, the operating variables of the filter can be changed to optimise the filtrate flux, with the knowledge that the changes will not influence the filtrate quality.

All engineering research has as its final objective, the deriving of a mathematical model of the process. Although only a minor attempt at this is reported in this thesis, there are some results that indicate the approach that might be profitably pursued. As mentioned earlier in this section, there is evidence to suggest that turbulence promotion has a negative effect on filtrate flux decline. A plausible explanation for this is that the increased turbulence causes more particles to reach the membrane surface, and thus block the membrane, whereas an increased crossflow velocity helps to remove particles that are already on the membrane surface. This is in line with the findings of Carter and Hoyland (24) who reported that the rate of deposition of particles is independent of the crossflow velocity, but that the equilibrium thickness of any particle layer is strongly dependent on the crossflow velocity. Thus there appears to be two mechanisms involved;a) deposition, which is independent of crossflow velocity but is increased by greater turbulence, and b) particle re-entrainment, which is strongly dependent on crossflow velocity, or more strictly, the shear stress at the surface. This indicates that the type of model required involves two mechanisms, and this approach is worth pursuing further.

An alternative approach is offered by blocking filtration. Appendix 1 examines this, and tests the models that Hermia uses over two runs. By far the most applicable of Hermia's models is the Standard Blocking Filtration Law. This model assumes that the pore volume decreases with the volume of filtrate. This implies that there is at least a tendency towards depth filtration.

A possible extension of Hermia's technique, is to assume a random pore structure, with specified mean size, and a random particle size distribution in the feed. In a computer programme, with some criteria for blocking, it should be possible to simulate the blocking of the membrane, from the interaction of the two random distributions. Obviously, to be of any practical significance, the model must predict the effect of velocity and pressure. This could perhaps be implemented by making the blocking criteria dependent on these two variables.

An equally important aspect is to model backflushing. At present no comprehensive data exist on the variation of the backflushing flux (the flux of fluid flowing through the filter during backflushing) with pressure. If this data were to be collected, it might be possible to correlate backflushing effectiveness, with the velocity in the membrane pores. This author believes that the pore velocity is the critical factor in backflushing.

In Section 9, velocity measurements were made near the membrane surface. It was thought that the velocity gradient close to the membrane surface was an important variable in crossflow filtration. Although it was not possible to determine this gradient close to the membrane, it would be desirable to know what effect mainstrean velocity, and other variables would have on this gradient. A practical approach as described in the discussion in Section 9 might provide better data, but a theoretical approach might provide a less tedious method. Some methods are described in the Literature Survey, and the approach of Rekin (46) seems to be as promising as any.

Crossflow filtration is not just applicable to clarification. Filtration is essentially a batch operation, although rotary drum filters and belt filters do allow continuous operation. A feature of batch operations is their inherently high labour cost. Crossflow filtration would seem to provide a easy method of semi-continuous operation. It is especially applicable where, for some reason, space is short. The particular operation to which this project was dedicated was a rather difficult one, due to the magnitude of the flows involved, and the unusual cost parameters that apply in offshore technology. Although it cannot be said to have been successfully applied, it seems that there is a reasonable prospect of success. To extend the principle of crossflow filtration to other operations, both large-scale, and especially small scale, would not seem to present any insurmountable obstacles.

SUGGESTIONS FOR FURTHER WORK

Most theses end with a number of suggestions for further research, which may assist in bringing the research to a more complete state. This thesis is no exception and a number of suggestions are presented in this section.

The next step in this research is to construct and test the prototype filter. Also it is essential to find a suitable membrane. The way in which this should be fulfilled is described in the Discussion (Section 13). When the most suitable membrane is ascertained, it will be possible to perform field tests on the prototype filter. These tests can be carried out under operating conditions that would be met if the filter were adopted for this application.

Following these tests, a careful evaluation must follow. There are three possible courses of action that could result from this evaluation. Firstly, it could be decided that crossflow filtration does not offer a viable alternative for the filtration of injection water, and thus the project would be discontinued. Secondly, it could be decided that crossflow filtration was viable for this application. Following this decision, a production design would be commenced, and production and marketing aspects would be considered. The third possibilty would be that insufficient data had been obtained from the field tests, and new tests were required before the decision to proceed could be taken.

The above strategy is the one that would be employed to fulfil the first objective given in the introduction (the application of crossflow filtration to the filtration of injection water used in oil extration). There are a number of tests that can be suggested which will assist in the acquisition of knowledge concerning the process of crossflow filtration.

The backflushing experiments are reasonably complete. The optimal cycle has been found, and the important variable has been identified. It would be desirable to derive an equation relating the effectiveness of a backflush to the backflush variables. This equation might then be only applicable for a particular membrane, so the effort required to derive this equation might outweigh the benefit obtained.

The filtrate quality tests require a great deal more work. The effect of pressure and velocity has been measured, although the reliabilty of this data is questionable. The effect of membrane history (number of backflushes), feed concentration and the actual membrane have still to be determined. It is also necessary to find a more suitable method of measuring filtrate quality, and suggestions for this are given is Section 7.

The effect of the two major process variables, pressure and crossflow velocity, on filtrate flux has not been studied for this thesis. Although several different groups have studied this, the data reported is diverse and, in the case of pressure, sometimes contradictory. Eventually these crucial variables will have to be studied in greater detail. A less important variable is feed concentration. Although there is some data on this in the literature, more study is required. An attempt should also be made to study the effect of other suspension parameters, for example the particle size distribution of the feed.

As mentioned earlier, the membrane for the prototype filter has not been specified. Eventually the effect of the other variables should be made independent of the membrane. Thus, it will be necessary to classify the membranes. This might well be done by a pore size distribution, and a series of other parameters which specify the behaviour of the membrane (for example, a blocking propensity).

The alternative objective of any academic research is to achieve a comprehensive theoretical understanding of the process being researched. In the Discussion (Section 13), a number of approaches that might be used to derive a mathematical model of crossflow filtration are described. All these approaches are worth investigating further, but the most promising is the "deposition/re-entrainment" approach. It is suggested that this approach be investigated further. The deposition of particles on the surface is probably quite easy to characterize, but it is going to be very difficult to simulate experiments where this mechanism is operative without the re-entrainment. The re-entrainment of particles will be more difficult to describe mathematically, but it is almost certain that the shear stress at the surface will be the important variable in any relation that is proposed.

The velocity measurements described in Section 9 were not very successful. For the approach described in the previous paragraph to be successful, it will be necessary to find an easy way of determining the shear stress at the surface. The meausement of this stress will not be easy, and perhaps the easiest way is to persue the theoretical methods outlined in the Literature Survey (Section 3).

The other theoretical approaches described in the Discussion (Section 13), are also worth pursuing. The extension of Blocking Filtration theory could well be of assitance. One extension that might be considered is to simulate blocking of a random distribution of pores, by an equally random distribution of particle sizes.

Finally, it would also be desirable to gain a theoretical understanding of the backflushing process. As stated in the Discussion, the effectiveness of a backflush is likely to be related to the pore velocity. A set of experiments should be performed measuring these pore velocities, and correlating them with backflushing effectiveness.

CONCLUSIONS

The major achievement of this thesis is the design of the prototype filter. This prototype encompasses most of the experience gained by the experimental work reported in this thesis. The prototype includes an optimised operating cycle, and an optimised channel geometry. The major omission from the design is that the membrane has not been specified.

Two conclusions can be drawn from the backflushing experiments. Firstly, the effectiveness of a particular backflush is very dependent on the pressure of the backflush, and nearly independent of the backflush duration. Secondly, frequent, high pressure, and short duration backflushes give an optimal operating strategy, and a constant average flux can be maintained. Also reported in Section 5 is that air scour is ineffective in controlling filtrate flux decline.

The channel geometry experiments indicated that the filtrate flux is not influenced either by the channel geometry, or by the Reynolds number of the crossflow. These conclusions are for a constant crossflow velocity.

The filtrate quality experiments examined the effect of pressure difference, crossflow velocity, and membrane pore size on concentration of particles in the filtrate. There was no evidence to suggest that any of these variables influenced the filtrate quality, although the reliability of this data cannot be vouched for. It was intended to measure the mean particle size of the filtrate, but these experiments provided diverse and often contradictory data. Suggestions are made which might assist in obtaining more reliable data.

In Section 8, some qualitative observations are reported. The most significant of these is that turbulence promotion leads to a greater deposition of particles on the membrane. The inference which can be drawn from this is that increasing the turbulence without increasing the crossflow velocity (by placing obstructions to the flow near the membrane surface) is likely to have a deleterious effect on the filtrate flux. Another interesting observation reported in this section is that larger particles rolling along the membrane surface can dislodge smaller particles and cause them to be re-entrained.

The Membrane Tests (Section 10) describe a quick way of ascertaining a membrane's tensile strength. Appendix 2 shows how the data obtained may be used to calculate the membrane deflection during backflushing. It was recommended that a new membrane be subjected first to these tests, to see if it is likely to be suitable, followed by the more exhaustive tests described in the Discussion (Section 13). The Membrane Survey (Section 12), is a survey of available membranes that might be suitable for this application.

Finally, some ideas as to the theoreticall description of crossflow filtration have been developed, and are described in the Discussion (Section 13).

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APPENDIX 1

BLOCKING FILTRATION

Blocking filtration occurs when the medium resistance changes with time. In cake filtration it as assumed that the extra resistance that is found during a filtration run is due entirely to the resistance of the cake and that the membrane resistance is a constant. Hermia (38) describes 3 different models of blocking filtration. They will be described below, along with a brief description of cake filtration.

Cake Filtration.

The theory of cake filtration rests on two assumptions; a)the resistance of the cake is proportional to the volume of filtrate, and b) the resistance of the filter medium is constant. From these assumptions it is possible to derive the following equation for constant pressure filtration.

$$K*V = t/V - 1/Q_o$$

Since K and Q_o are constants the equation is linear and a graph of t/V vs V will be a straight line.

Complete Blocking Filtration Law.

This model is based on the assumption that every particle in the feed, capable of blocking the membrane, when it reaches the membrane finds an open pore and seals it completely. Particles cannot be superimposed on each other and cannot rest on a non-active area of the membrane. With this assumption the following equation can be derived

$$Q = Q_o - P*s*V/(u*R)$$

where Q = volumetric flowrate
Q_o = initial volumetric flowrate
P = applied pressure
s = blocked area per unit filtrate volume
V = volume of filtrate
u = viscosity
R = filter resistance

Since Q_o, s, P, u, and R are constants the equation is linear and a graph of Q vs V will be a straight line.

Intermediate Blocking Law

This model is an extension of the previous model. When a particle reaches an open pore it will completely seal the pore. However it is possible for a particle to reach an area of the membrane where there is already a particle, or for a particle to fall on a non-active area of the membrane. For this model Hermia derives the following equation

$$Q = Q_o/(1 + (sP/uR)*t)$$

where the nomemclature is as in the previous model. From this equation, it is possible to rearrange it as follows

$$kt = 1/Q - 1/Q_o$$

This equation is linear, and a plot of $1/Q$ vs t should yield a straight line.

Standard Blocking Filtration Law.

The assumption for this model is that all the particles in the feed are absorbed onto the walls of the pores. As the material is deposited the volume of the pores decreases by the amount of material filtered out. Hermia assumes that the proportional volume reduction is equal to the proportional area reduction. By using a mass balance and the Poiseuille equation the following relation may be derived

$$Kt = t/V - 1/Q_o$$

This equation is also linear and a plot of t/V vs t will yield a straight line.

APPLICATION TO THIS STUDY.

It is possible to take results obtained from crossflow filtration and try and fit the data to these models to see which, if any, is applicable to the experimental evidence. This was done for some of the data obtained in this study. Two runs are presented whose flux vs time curves are given in Graph A1.1.

A digital computer was used to plot the data shown on Graph A1.1 in 4 different ways. Table 1 summarizes the plots required.

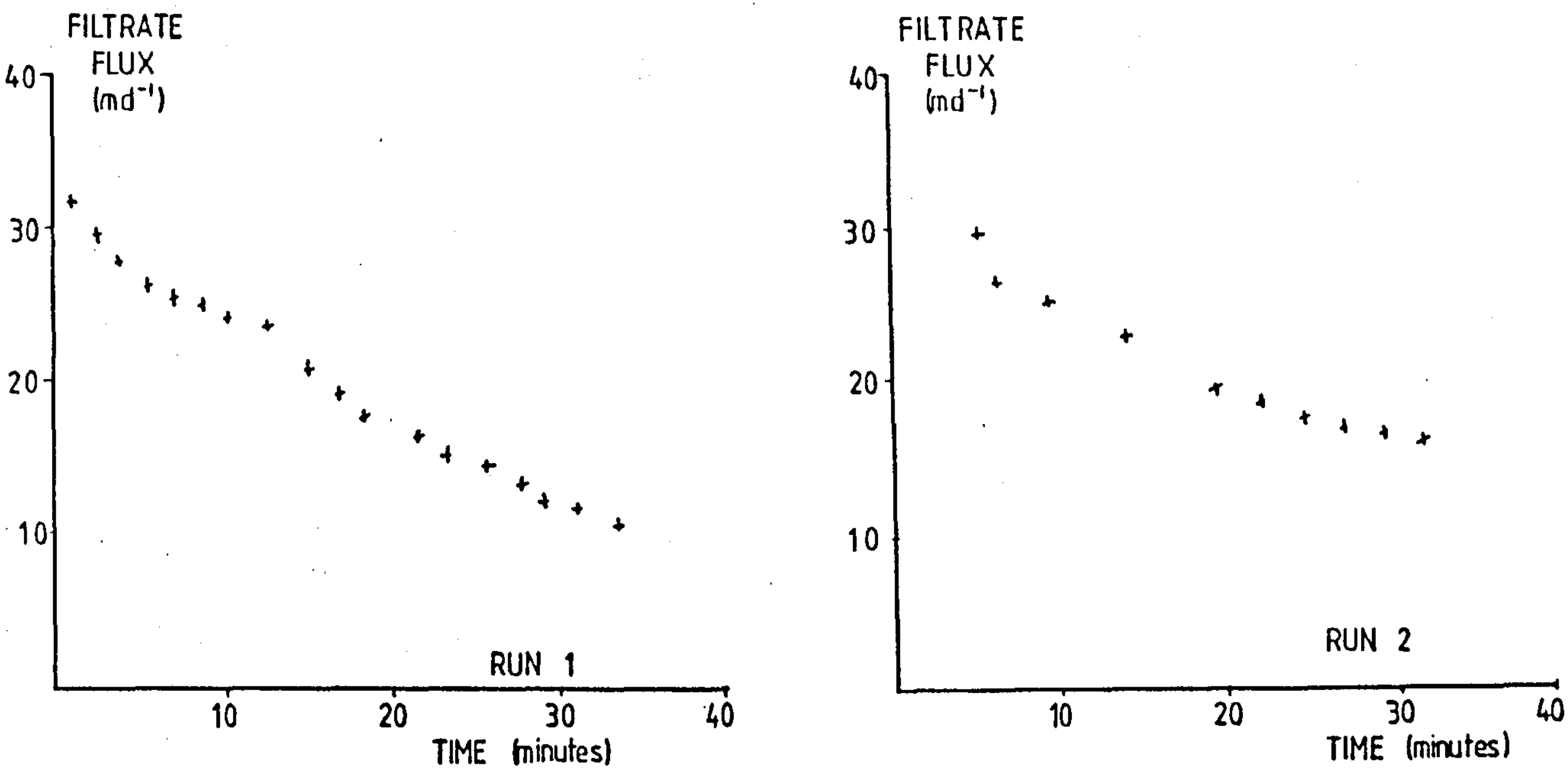
TABLE 1

MODEL	PLOT REQUIRED
Cake Filtration	t/V vs V
Complete Blocking	Q vs V
Intermediate Blocking	1/Q vs t
Standard Blocking	t/V vs t

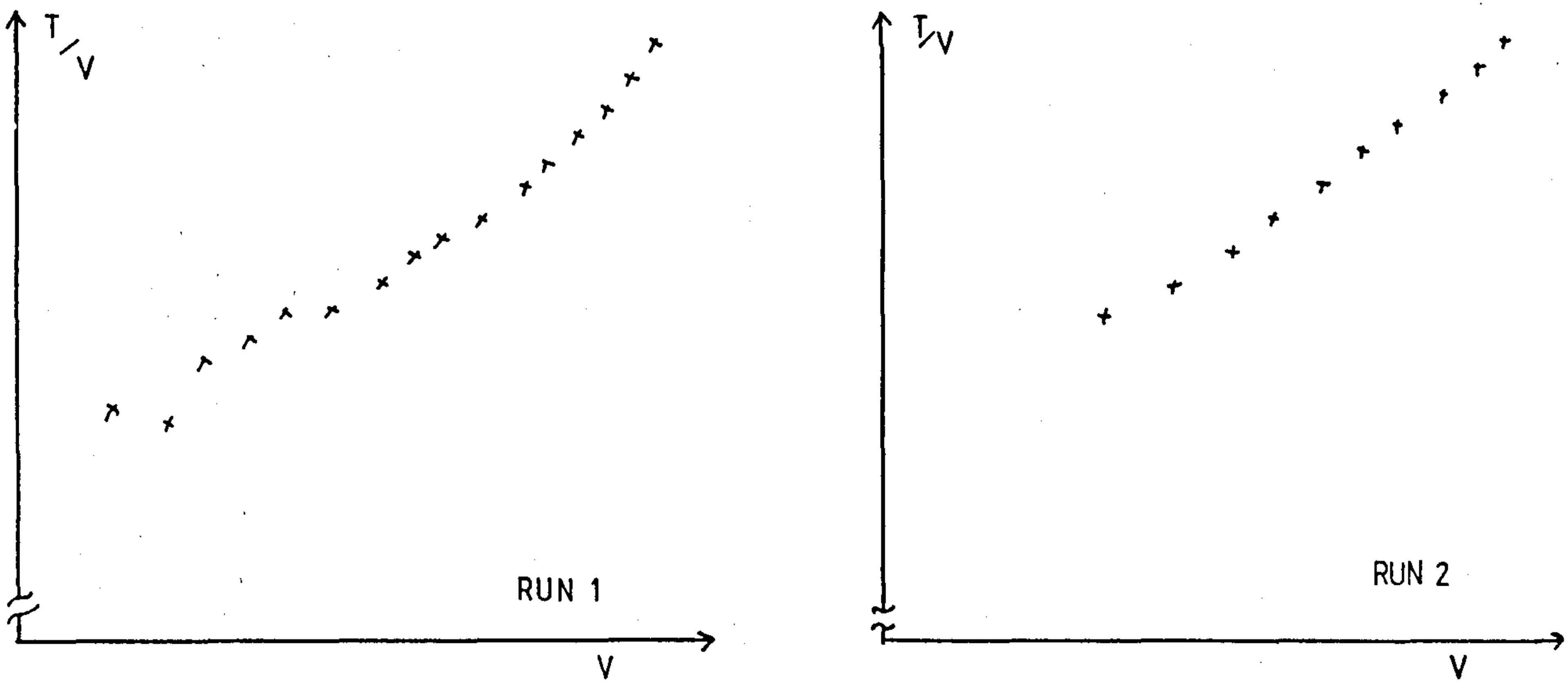
Each plot, if it gave a straight line would indicate that one of the models described above would be applicable.

RESULTS

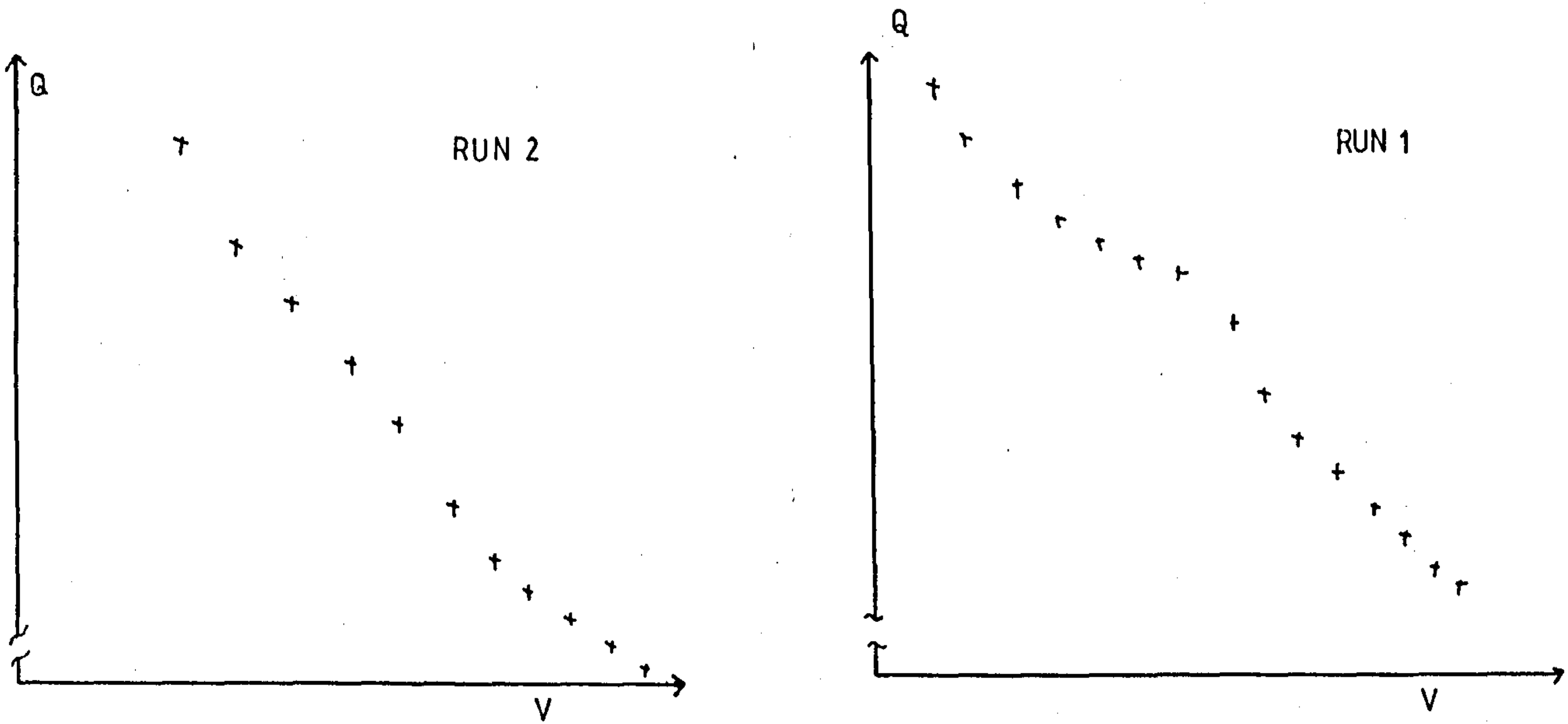
Each of the plots detailed above are given in Graphs A1.2 - A1.5. Since only the qualitative aspects of these graphs will be considered, the scales of the graphs have not been marked. The y-axis does not have its origin at zero, but has been broken so that a better appreciation of the linearity of the graph can be gained. These results are discussed in the Discussion chapter in the main body of the thesis.



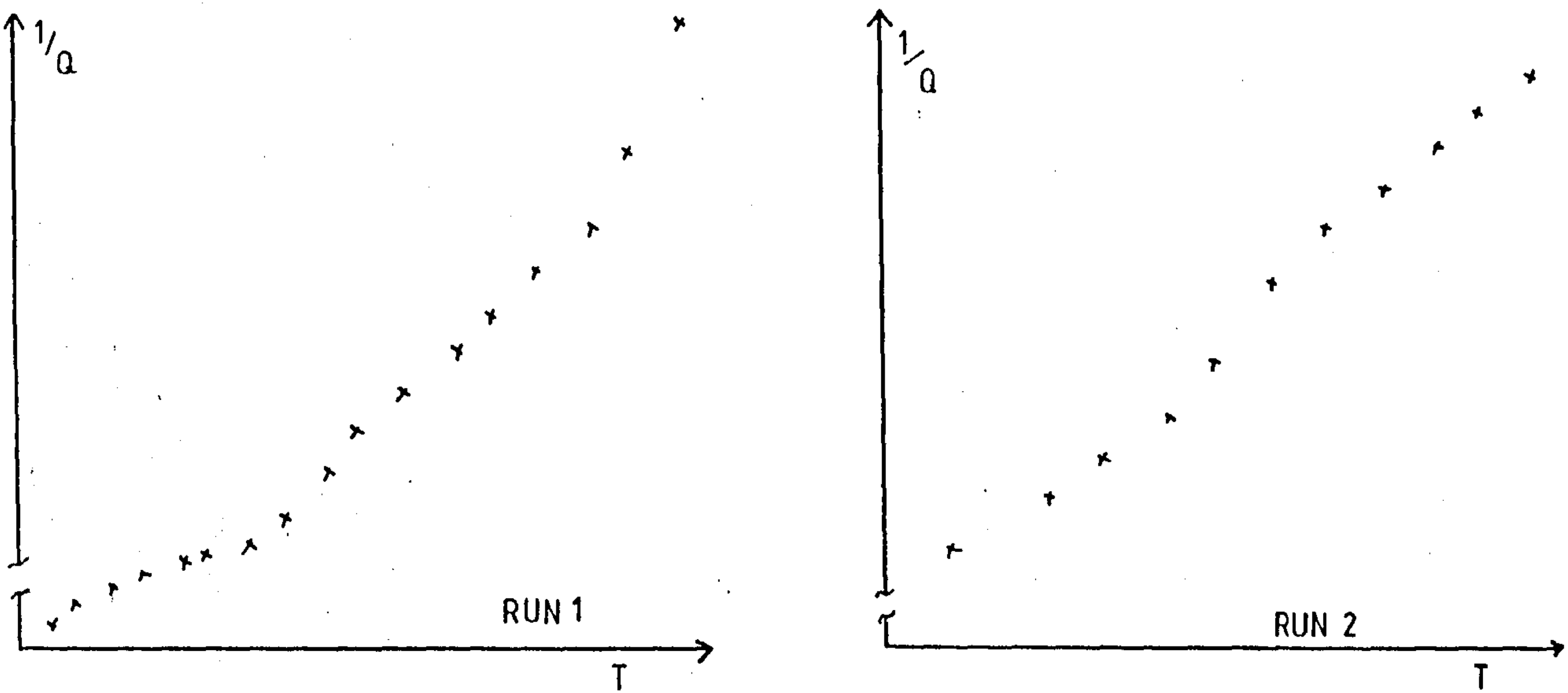
GRAPH A1.1 Flux vs Time curve for Runs 1 and 2



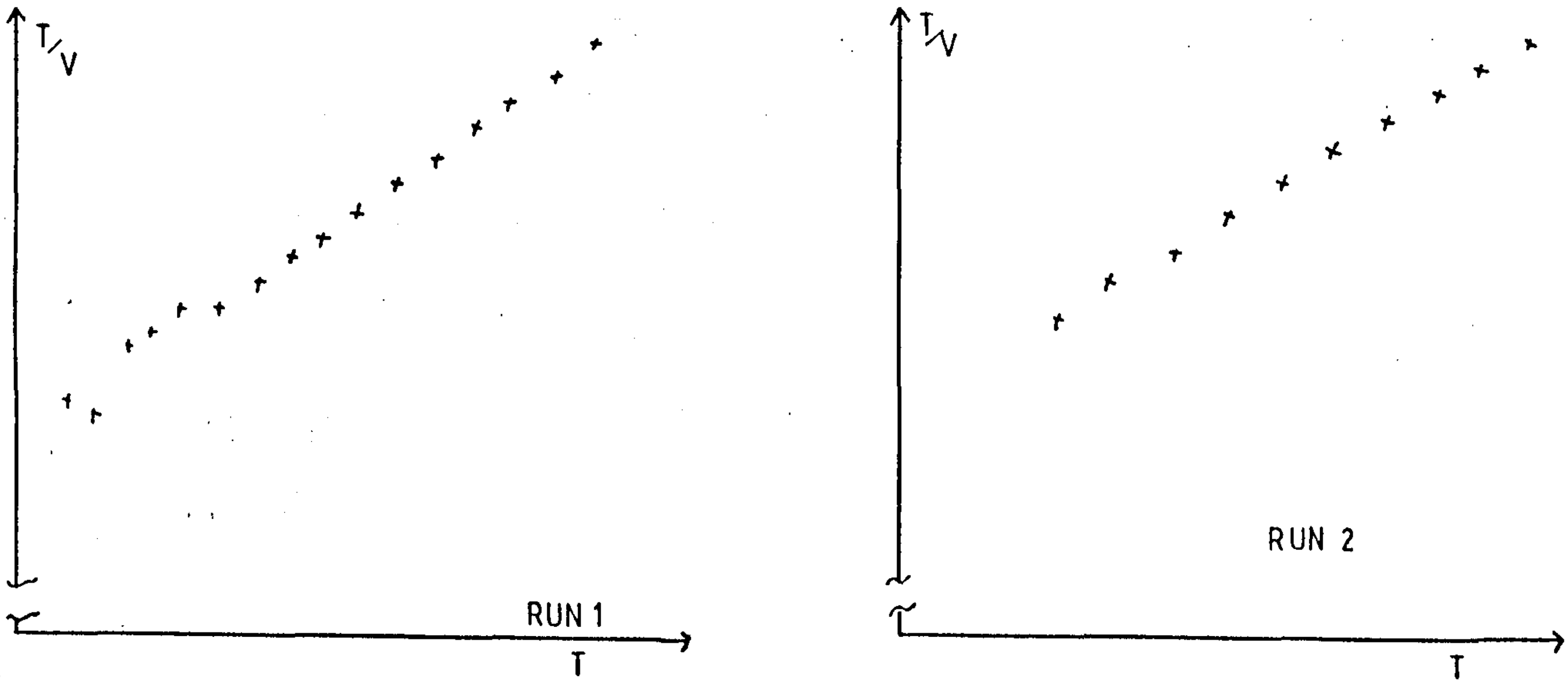
GRAPH A1.2 t/V vs V curve for Runs 1 and 2.
Linearity is measure of Cake Filtration.



GRAPH A1.3 Q vs V curve for Runs 1 and 2.
Linearity is measure of Complete Blocking Filtration.



GRAPH A1.4 1/Q vs t curve for Runs 1 and 2.
Linearity is measure of Intermediate Blocking Filtration.



Graph A1.5 t/V vs t curve for Runs 1 and 2.
Linearity is measure of Standard Blocking Filtration.

APPENDIX 2

1. CALCULATION OF DEFLECTION OF MEMBRANE FROM THE ELONGATION.

Figure A2.1 represents the situation. The width represents the channel width and the h is the value of the membrane deflection. By simple trigonometry;

$$0.5 * w = r \cos(a)$$

$$r = w/(2 \cos(a)) \qquad (1)$$

Also $r - h = r \operatorname{Sin}(a)$

$$h = r(1 - \sin(a)) \qquad (2)$$

The elongation of the membrane above its length when undistorted is the ratio of the length of the arc subtended by the angle $2*(90 - a)$ to w.

The length of the arc is

$(2*(90 - a)/360) * 2*PI*r$ where a is expressed in degrees

$$\text{Elongation} = \frac{2*(90-a)*2*PI*r}{360} - w \quad /w$$

$$= \frac{2*(90-a)*2*PI*w*\cos(a)}{360 \qquad\quad 2} - w \quad /w$$

Since w is known, this equation then relates to a. Although this equation is not explicit in a, it is easy to iterate using a digital computer, and obtain a table of values relating elongation to a. From equation 1, r can then be determined and from equation 2, the deflection h is easily determined for a particular elongation.

This was done on a digital computer, using a channel size of 5mm (w=5) and the data are given in Graph A2.1

CALCULATION OF MEMBRANE DEFLECTION DURING BACKWASH.

Consider a strip of membrane 1cm wide by w long, across the channel. Let the channel width w=5mm.

The force extended on this membrane during a backwash of pressure P is

F = P*area = P*10mm*5mm

Assume P = 272 kPa.

F = 272E3*10E-3*5E-3.

= 13.6 N per cm width.

From Graph 10.2, this would cause a deflection of approximately 6%. Then from Graph A2.1 this elongation would result in a deflection of approximately 0.75 mm.

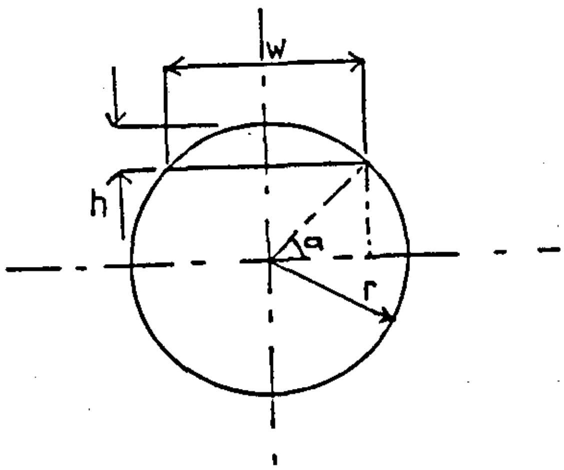
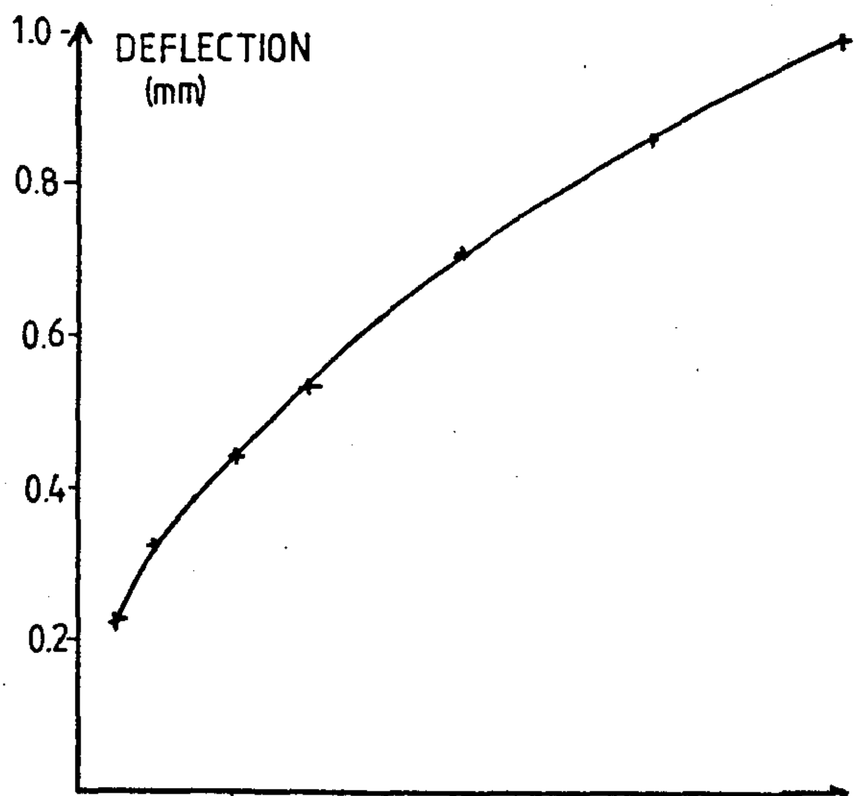


Fig A2.1 Diagram showing how a membrane distorts during backflushing



Graph A2.1 Graph relating deflection of membrane, to the percentage elongation.

APPENDIX 3

MEMBRANE SURVEY LETTER

This appendix is a copy of the letter that was sent out to various membrane manufacturers to survey available membranes. The results are collated in Section 12.

University of Technology

LOUGHBOROUGH LEICESTERSHIRE LE11 3TU Tel: 0509 63171 Telex 34319 Telegrams Technology Loughborough

DEPARTMENT OF CHEMICAL ENGINEERING
Professor and Head of Department
D. C. Freshwater

AJC/JAB February, 1982

The Technical Director,

Dear Sir,

At present we are designing a cross flow filter for injection water in the North Sea. This is a high flux application and the prospects for its adoption on North Sea installations are extremely good. It is anticipated that this project will lead to a lucrative commercial venture.

Our design involves frequent high pressure short duration reversals of flow. This reversal exerts a high mechanical stress on a membrane. It is easy to support the membrane during the filtration mode. However, during the backwash mode it is impossible; if it is supported then one no longer has cross flow.

A product is required with a maximum particle size of about 1 μm. Utilizing the cross flow principle we believe we can use a membrane with a maximum pore size of between 3 and 5 μm.

What we need therefore is a robust membrane, which does not swell in water. It should have a good tensile strength and retain its tensile properties when wet. The more rigid the membrane the better.

Do your company make any membranes which would meet these requirements? If not, do you have any ideas as to how we might obtain a suitable membrane.

Thank you for your assistance.

Yours sincerely,

A. J. Carter
Research Assistant

APPENDIX 4

CHANNEL GEOMETRY PROGRAMME

This appendix includes both sample printout, and a full listing of the programme used to optimise the channel geometry. A full explanation, and a flow diagram for the algorithm are given in the Prototype Design (Section 11).

```
CHANNEL GEOMETRY
                   CHANNEL SHAPE  TRIANGULAR
                   CHANNEL HEIGHT 1.5 MM
                   CHANNEL WIDTH  3 MM

NUMBER OF CHANNELS 500

FILTRATE FLUX 40 M/H             .0111111 M/S

CALCULATED VALUES
                   VELOCITY AT CHANNEL ENTRANCE 7   M/S
                   VELOCITY AT CHANNEL EXIT     1.4  M/S
                   AVERAGE VELOCITY             4.3  M/S

                   EFFECTIVE CHANNEL LENGTH REQD
                   FOR 20% REJCTION              .38

                   PRESSURE DROP ALONG CHANNEL 142.733 KPA
                                               20.6 PSI

                   PUMP OUTPUT    7.83098E-03 M**3/SEC
                                  103.4 GPM


FILTRATE FLUX 20 M/H             5.55556E-03 M/S

CALCULATED VALUES
                   VELOCITY AT CHANNEL ENTRANCE 5.9  M/S
                   VELOCITY AT CHANNEL EXIT     1.2  M/S
                   AVERAGE VELOCITY             3.6  M/S

                   EFFECTIVE CHANNEL LENGTH REQD
                   FOR 20% REJCTION              .64

                   PRESSURE DROP ALONG CHANNEL 183.762 KPA
                                               26.6 PSI

                   PUMP OUTPUT    6.68127E-03 M**3/SEC
                                  88.2 GPM
```

```
CHANNEL GEOMETRY
                   CHANNEL SHAPE  EQUILATERAL TRIANGLE
                   CHANNEL HEIGHT 2 MM
                   CHANNEL WIDTH  2.3 MM

NUMBER OF CHANNELS 500

FILTRATE FLUX 40 M/H             .0111111 M/S

CALCULATED VALUES
                   VELOCITY AT CHANNEL ENTRANCE 6.6  M/S
                   VELOCITY AT CHANNEL EXIT     1.3  M/S
                   AVERAGE VELOCITY             4  M/S

                   EFFECTIVE CHANNEL LENGTH REQD
                   FOR 20% REJCTION              .48

                   PRESSURE DROP ALONG CHANNEL 151.571 KPA
                                               21.9 PSI

                   PUMP OUTPUT    7.64226E-03 M**3/SEC
                                  100.9 GPM


FILTRATE FLUX 20 M/H             5.55556E-03 M/S

CALCULATED VALUES
                   VELOCITY AT CHANNEL ENTRANCE 5.6  M/S
                   VELOCITY AT CHANNEL EXIT     1.1  M/S
                   AVERAGE VELOCITY             3.4  M/S

                   EFFECTIVE CHANNEL LENGTH REQD
                   FOR 20% REJCTION              .81

                   PRESSURE DROP ALONG CHANNEL 190.805 KPA
                                               27.6 PSI

                   PUMP OUTPUT    6.46679E-03 M**3/SEC
                                  85.4 GPM
```

```
CHANNEL GEOMETRY
                    CHANNEL SHAPE  RECTANGULAR
                    CHANNEL HEIGHT 1.06 MM
                    CHANNEL WIDTH  2.1 MM

NUMBER OF CHANNELS 500

FILTRATE FLUX 40 M/H             .0111111 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 6.7  M/S
                    VELOCITY AT CHANNEL EXIT     1.3  M/S
                    AVERAGE VELOCITY             4.1  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              .51

                    PRESSURE DROP ALONG CHANNEL 154.603 KPA
                                                 22.4 PSI

                    PUMP OUTPUT    7.57682E-03 M**3/SEC
                                   100 GPM


FILTRATE FLUX 20 M/H             5.55556E-03 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 5.7  M/S
                    VELOCITY AT CHANNEL EXIT     1.1  M/S
                    AVERAGE VELOCITY             3.4  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              .86

                    PRESSURE DROP ALONG CHANNEL 193.645 KPA
                                                 28 PSI

                    PUMP OUTPUT    6.38069E-03 M**3/SEC
                                   84.2 GPM
```

```
CHANNEL GEOMETRY
                    CHANNEL SHAPE  RECTANGULAR
                    CHANNEL HEIGHT 1.5 MM
                    CHANNEL WIDTH  1.5 MM

NUMBER OF CHANNELS 500

FILTRATE FLUX 40 M/H             .0111111 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 6.3  M/S
                    VELOCITY AT CHANNEL EXIT     1.3  M/S
                    AVERAGE VELOCITY             3.8  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              .68

                    PRESSURE DROP ALONG CHANNEL 171.539 KPA
                                                 24.8 PSI

                    PUMP OUTPUT    7.09647E-03 M**3/SEC
                                   93.7 GPM


FILTRATE FLUX 20 M/H             5.55556E-03 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 5.3  M/S
                    VELOCITY AT CHANNEL EXIT     1.1  M/S
                    AVERAGE VELOCITY             3.1  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              1.14

                    PRESSURE DROP ALONG CHANNEL 209.554 KPA
                                                 30.3 PSI

                    PUMP OUTPUT    5.92111E-03 M**3/SEC
                                   78.2 GPM
```

```
CHANNEL GEOMETRY
                    CHANNEL SHAPE  EQUILATERAL TRIANGLE
                    CHANNEL HEIGHT 4 MM
                    CHANNEL WIDTH  4.6 MM

NUMBER OF CHANNELS 125

FILTRATE FLUX 40 M/H            .0111111 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 6.9  M/S
                    VELOCITY AT CHANNEL EXIT     1.4  M/S
                    AVERAGE VELOCITY             4.2  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              .99

                    PRESSURE DROP ALONG CHANNEL 140.798 KPA
                                                20.3 PSI

                    PUMP OUTPUT   7.96259E-03 M**3/SEC
                                  105.1 GPM


FILTRATE FLUX 20 M/H            5.55556E-03 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 5.9  M/S
                    VELOCITY AT CHANNEL EXIT     1.2  M/S
                    AVERAGE VELOCITY             3.5  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              1.69

                    PRESSURE DROP ALONG CHANNEL 180.554 KPA
                                                26.1 PSI

                    PUMP OUTPUT   6.77213E-03 M**3/SEC
                                  89.4 GPM
```

```
CHANNEL GEOMETRY
                    CHANNEL SHAPE  TRIANGULAR
                    CHANNEL HEIGHT 3 MM
                    CHANNEL WIDTH  6 MM

NUMBER OF CHANNELS 125

FILTRATE FLUX 40 M/H            .0111111 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 7.2  M/S
                    VELOCITY AT CHANNEL EXIT     1.4  M/S
                    AVERAGE VELOCITY             4.5  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              .78

                    PRESSURE DROP ALONG CHANNEL 132.179 KPA
                                                19.1 PSI

                    PUMP OUTPUT   8.15135E-03 M**3/SEC
                                  107.6 GPM


FILTRATE FLUX 20 M/H            5.55556E-03 M/S

CALCULATED VALUES
                    VELOCITY AT CHANNEL ENTRANCE 6.2  M/S
                    VELOCITY AT CHANNEL EXIT     1.2  M/S
                    AVERAGE VELOCITY             3.8  M/S

                    EFFECTIVE CHANNEL LENGTH REQD
                    FOR 20% REJCTION              1.34

                    PRESSURE DROP ALONG CHANNEL 173.383 KPA
                                                25.1 PSI

                    PUMP OUTPUT   6.98953E-03 M**3/SEC
                                  92.3 GPM
```

CHANNEL GEOMETRY

CHANNEL SHAPE RECTANGULAR
CHANNEL HEIGHT 3 MM
CHANNEL WIDTH 3 MM

NUMBER OF CHANNELS 125

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6.6 M/S
VELOCITY AT CHANNEL EXIT 1.3 M/S
AVERAGE VELOCITY 4 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.43

PRESSURE DROP ALONG CHANNEL 162.418 KPA
23.5 PSI

PUMP OUTPUT 7.44193E-03 M**3/SEC
98.2 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 5.5 M/S
VELOCITY AT CHANNEL EXIT 1.1 M/S
AVERAGE VELOCITY 3.3 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 2.39

PRESSURE DROP ALONG CHANNEL 198.723 KPA
28.7 PSI

PUMP OUTPUT 6.22165E-03 M**3/SEC
82.1 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE RECTANGULAR
CHANNEL HEIGHT 2.13 MM
CHANNEL WIDTH 4.2 MM

NUMBER OF CHANNELS 125

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 7 M/S
VELOCITY AT CHANNEL EXIT 1.4 M/S
AVERAGE VELOCITY 4.3 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.07

PRESSURE DROP ALONG CHANNEL 142.933 KPA
20.7 PSI

PUMP OUTPUT 7.91135E-03 M**3/SEC
104.4 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 5.9 M/S
VELOCITY AT CHANNEL EXIT 1.2 M/S
AVERAGE VELOCITY 3.6 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.82

PRESSURE DROP ALONG CHANNEL 182.365 KPA
26.4 PSI

PUMP OUTPUT 6.71704E-03 M**3/SEC
88.7 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE EQUILATERAL TRIANGLE
CHANNEL HEIGHT 6 MM
CHANNEL WIDTH 6.9 MM

NUMBER OF CHANNELS 50

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 7.4 M/S
VELOCITY AT CHANNEL EXIT 1.5 M/S
AVERAGE VELOCITY 4.5 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.59

PRESSURE DROP ALONG CHANNEL 151.604 KPA
21.9 PSI

PUMP OUTPUT 7.64003E-03 M**3/SEC
100.8 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6.2 M/S
VELOCITY AT CHANNEL EXIT 1.2 M/S
AVERAGE VELOCITY 3.8 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 2.69

PRESSURE DROP ALONG CHANNEL 190.854 KPA
27.6 PSI

PUMP OUTPUT 6.45953E-03 M**3/SEC
85.3 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE TRIANGULAR
CHANNEL HEIGHT 4.5 MM
CHANNEL WIDTH 9 MM

NUMBER OF CHANNELS 50

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 7.8 M/S
VELOCITY AT CHANNEL EXIT 1.6 M/S
AVERAGE VELOCITY 4.8 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.26

PRESSURE DROP ALONG CHANNEL 144.705 KPA
20.9 PSI

PUMP OUTPUT 7.86937E-03 M**3/SEC
103.9 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6.6 M/S
VELOCITY AT CHANNEL EXIT 1.3 M/S
AVERAGE VELOCITY 4 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 2.14

PRESSURE DROP ALONG CHANNEL 183.708 KPA
26.6 PSI

PUMP OUTPUT 6.67378E-03 M**3/SEC
88.1 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE TRIANGULAR
CHANNEL HEIGHT 2.5 MM
CHANNEL WIDTH 5 MM

NUMBER OF CHANNELS 100

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 8.6 M/S
VELOCITY AT CHANNEL EXIT 1.7 M/S
AVERAGE VELOCITY 5.4 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION .77

PRESSURE DROP ALONG CHANNEL 219.949 KPA
31.8 PSI

PUMP OUTPUT 5.36295E-03 M**3/SEC
70.8 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6.8 M/S
VELOCITY AT CHANNEL EXIT 1.4 M/S
AVERAGE VELOCITY 4.2 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.23

PRESSURE DROP ALONG CHANNEL 236.011 KPA
34.2 PSI

PUMP OUTPUT 4.27187E-03 M**3/SEC
56.4 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE RECTANGULAR
CHANNEL HEIGHT 4.5 MM
CHANNEL WIDTH 4.5 MM

NUMBER OF CHANNELS 50

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 7 M/S
VELOCITY AT CHANNEL EXIT 1.4 M/S
AVERAGE VELOCITY 4.3 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 2.27

PRESSURE DROP ALONG CHANNEL 171.65 KPA
24.8 PSI

PUMP OUTPUT 7.08923E-03 M**3/SEC
93.6 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 5.9 M/S
VELOCITY AT CHANNEL EXIT 1.2 M/S
AVERAGE VELOCITY 3.5 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 3.8

PRESSURE DROP ALONG CHANNEL 209.762 KPA
30.4 PSI

PUMP OUTPUT 5.9299E-03 M**3/SEC
78.3 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE TRIANGULAR
CHANNEL HEIGHT 2.5 MM
CHANNEL WIDTH 5 MM

NUMBER OF CHANNELS 150

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 7.8 M/S
VELOCITY AT CHANNEL EXIT 1.6 M/S
AVERAGE VELOCITY 4.8 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION .7

PRESSURE DROP ALONG CHANNEL 168.189 KPA
24.3 PSI

PUMP OUTPUT 7.29181E-03 M**3/SEC
96.2 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6.5 M/S
VELOCITY AT CHANNEL EXIT 1.3 M/S
AVERAGE VELOCITY 4 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.17

PRESSURE DROP ALONG CHANNEL 206.5 KPA
29.9 PSI

PUMP OUTPUT 6.10168E-03 M**3/SEC
80.5 GPM

CHANNEL GEOMETRY

CHANNEL SHAPE TRIANGULAR
CHANNEL HEIGHT 2.5 MM
CHANNEL WIDTH 5 MM

NUMBER OF CHANNELS 200

FILTRATE FLUX 40 M/H .0111111 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6.8 M/S
VELOCITY AT CHANNEL EXIT 1.4 M/S
AVERAGE VELOCITY 4.1 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION .61

PRESSURE DROP ALONG CHANNEL 114.34 KPA
16.5 PSI

PUMP OUTPUT 8.43877E-03 M**3/SEC
111.4 GPM

FILTRATE FLUX 20 M/H 5.55556E-03 M/S

CALCULATED VALUES

VELOCITY AT CHANNEL ENTRANCE 6 M/S
VELOCITY AT CHANNEL EXIT 1.2 M/S
AVERAGE VELOCITY 3.6 M/S

EFFECTIVE CHANNEL LENGTH REQD
FOR 20% REJCTION 1.07

PRESSURE DROP ALONG CHANNEL 162.116 KPA
23.4 PSI

PUMP OUTPUT 7.4443E-03 M**3/SEC
98.3 GPM

```
1 REM ********************************************
2 DIM RE(10),V(11),P(11)
10 REM **** PROGRAM TO CALCULAtE VARIOUS
20 REM **** POSSIBLE HYDRODYNAMIC VARIABLES
30 REM **** IN THE DESIGN OF A CROSS FLOW
40 REM **** FILTER
50 REM **** WRITTEN BY A.J. CARTER (COPYWRIGHT)
60 REM ********************************************
70 REM
80 REM
90 REM *** INPUT PUMP DATA FROM TAPE
95 GOSUB 10000
100 REM ***  INPUT DESIGN PARAMETERS
110 REM ***                                    AREA
120 PRINT "IS CROSS FLOW CHANNEL AN EQUILATERAL TRIANGLE"
130 INPUT Z$
140 IF Z$ < "Y" THEN 200
145 R$="EQ"
150 PRINT " INPUT HEIGHT (VERTICAL) OF CHANNEL (MM)
160 INPUT H
170 H=H/1000
175 B=H/0.75
180 A=0.5*H*B
190 GOTO300
200 PRINT "IS CHANNEL RECTANGULAR"
210 INPUT Z$
211 R$="T"
213 IF Z$ >= "Y" THEN R$="REC"
215 Z=1
220 IF Z$ < "Y" THEN Z=0.5
230 PRINT " INPUT CHANNEL WIDTH (MM)"
240 INPUT B
250 B=B/1000
260 PRINT "INPUT CHANNEL HEIGHT (MM) (VERTICAL)"
270 INPUT H
280 H=H/1000
290 A=Z*B*H
295 GOSUB 2000
300 GOSUB2000
310 REM *** INPUT NUMBER OF CHANNELS"
320 PRINT "INPUT NUMBER OF CHANNELS REQUIRED"
330 INPUT N
335 GOSUB 3000
340 REM **
350 REM **  CALCULATE TOTAL AREA
360 A1=N*A
370 REM ** INPUT INITIAL FLOWRATE
380 PRINT "INPUT INITIAL FLOWRATE (FROM PUMP) (GPM)"
390 INPUT Q1
400 Q=Q1 *7.57682E-05
410 REM ***
420 REM *** CALCULATE INITIAL VELOCITY
430 V=Q/(A*N)
440 REM **  INPUT FILTRATE FLUX (M/H)
450 PRINT "INPUT FILTRATE FLUX (M/H)"
460 INPUT F1
470 F=F1/3600
475 GOSUB 4000
480 REM ***
485 CT=0
490 REM *** CACULATE EFFECTIVE CHANNEL LENGTH TO
500 REM *** OBTAIN 20% REJECTION
510 REM FLOWRATE ALONG CHANNEL IS Q/N
515 CT=CT+1
520 REM FLOWRATE THRU MEMBRANE IS F*B*L
530 REM *** THEN  F*B*L = 0.8 * Q/N
540 REM ***   L=0.8*Q/(*F*B*N)
550 L=0.8*Q/(F*B*N)
555 PRINT L,CT
560 REM *** ONE MUST NOW ITERATE CALCULATING
580 REM *** 1/ REYNOLDS NUMBER
590 REM *** 2/ FRICTION FACTOR
600 REM *** 3/ PRESSURE DROP
610 REM ***  FOR EACH LENGTH OF PIPE
620 REM ***  ONE MUST THEN CALCULATE
630 REM ***  1/THE AMOUNT OF FILTRATE FOR EACH LENGTH
640 REM ***  2/THE FLOW THRU EACH CHANNEL FOR THE NEW SECTION
650 REM ***  3/THE NEW VELOCITY IN THE NEW SECTION
660 REM ***
670 REM *** WE ASSUME THAT THE FLUX IS THE SAME FOR EACH
680 REM *** SECTION
690 REM *** IE THE FLUX IS INDEPENDANT OF PRESSURE
700 REM ***    DIFFERENTIAL AND VELOCITY
710 REM ***
720 REM *** Q2=FLOW IN EACH CHANNEL (Q/N)
730 Q2=Q/N
740 REM ****   USE L/10 AS ITERATION LENGTH
750 L1=L/10
760 REM
770 REM ***  CALCULATE HYDRAULIC RADIUS
780 REM ***   RH=4*CSA/(WETTED PERIMETER)
790 IF R$="REC" THEN WP=2*(B+H):     GOTO 850
800 REM *** CALC SLOPE HEIGHT
810 REM *** SH [2=H [2 + (0.5*B) [2
820 SH=SQR(H[2 +(.5*B) [2)
830 WP=2*SH + B
850 RH =4*A/WP
```

```
850 RH =4*A/WP
860 REM ***  ITERATE 10 TIMES
870 V(1)=Q/(N*A)
880 FOR J=1 TO 10
890 REM *** CALC REYNOLDS
900 RE(J)= RH *V(J) * 1000 / 1.013E-03
910  REM ***  CALCULATE F (FRICTION FACTOR)
920 REM **USE SMOOTH PIPE CORRELATION
930 GOSUB 5000
940 REM ***  ADD 20% TO ACCOUNT FOR SURFACE ROUGHNESS
950 FF=FF*1.2
960 REM *** CALCULATE PRESSURE DROP FOR SECTION
970 REM *** USE FOLLOWING EQUATION
980 REM *** P = L * (4/RH) * 0.5 * DE *V[2 * FF
990 REM ** WHERE DE = DENSITY =1000 KG/M[3
1000 P(J)=L1*(4/RH)*0.5*1E3*V(J)[2*FF
1010 REM ** CALC NEW VELOCITY AT NEXT SECTION
1020 REM V(J+1)=V(J) - F*L1*B/A
1030 V(J+1)=V(J)-F*L1*B/A
1040 NEXT J
1050 REM CALCULATE TOTAL PRESSURE DROP
1060 PD=0
1070 FOR J=1 TO 10
1080 PD=PD+P(J)
1090 NEXT J
1100 REM **** NOW NEED TO COMPARE PD WITH CAPACITY
1110 REM *** OF PUMP
1120 REM *** GET PUMPING CAPCITY OF PUMP AT
1130 REM *** FLOWRATE Q
1140 GOSUB 6000
1150 REM *** TEST PD AGAINST PP
1160 REM USE FOLLOWING AS CONVERGENCE PROCEDURE
1170 REM *** OUTLET PRESSURE SHOULD BE 15PSI
1180 REM ***THUS PP=PD+15 PSI
1190 REM *** THUS USE NEW Q=(PP/(PD+15PSI)*Q) *0.5  + Q
1200 PR=15*6.89476 E03
1205 IF ABS(PP-(PR+PD)) < 0.01*PP THEN1250
1210 Q=(PP/(PD+PR)*Q)*0.5+Q*0.5
1220 GOTO490
1250 REM *** DATA ARE CONVERGENT
1260 LET C$="AT ENTRANCE"
1270 LET D$="AT EXIT
1275 E$="AVERAGE"
1280 PRINT "VELOCITY",C$;V(1)
1290 PRINT ,D$,V(11)
1300 PRINT "FLOWRATE";Q;"M**3/S",Q/7.57682E-05;"GPM"
1300 PRINT "FLOWRATE";Q;"M**3/S",Q/7.57682E-05;"GPM"
1310 PRINT "PRESSURE DROP";PD/1000;"KPA",PD/6.89476E03;
1320 PRINT "PSI"
1321 GOSUB 8000
1325 GOSUB 9000
1330 REM *** PRESENT MENU TO CHANGE FACTORS
1335 IF F1=40 THEN F1=20:GOTO 470
1340 PRINT
1350 PRINT "WHAT DO YOU WANT TO VARY"
1360 PRINT "CHANNEL GEOMETRY     1"
1370 PRINT "NUMBER OF CHANNELS  2"
1380 PRINT "INITIAL FLOWRATE     3"
1390 PRINT "FILTRATE FLUX        4"
1395 PRINT "TO TERMINATE         5"
1400 PRINT "INPUT APPROPRAITE NUMBER ";
1410 INPUT I
1420 IF I=1 THEN120
1430 IF I=2 THEN 310
1440 IF I=3 THEN 370
1450 IF I=4 THEN 440
1460 IF I > 4 THEN STOP
1470 GOTO1360
2000 REM *** SUBROUTINE TO PRINT OUT GEOMETRIC FACTORS
2010 LPRINT"CHANNEL GEOMETRY"
2020 LPRINTTAB(20)"CHANNEL SHAPE  ";
2030 IF R$="REC" THEN LPRINT "RECTANGULAR"
2040 IF R$="EQ" THEN LPRINT "EQUILATERAL TRIANGLE"
2050 IF R$="T" THEN LPRINT "TRIANGULAR"
2055 IF R$="" THEN PRINT "PANIC":STOP
2060 LPRINT TAB(20)"CHANNEL HEIGHT";H*1000;"MM"
2065 BB=(INT(B*10000+0.0005)/10)
2070 LPRINT TAB(20)"CHANNEL WIDTH ";BB;"MM"
2080 LPRINT
2090 RETURN
3000 REM *** PROGRAM TO PRINT OUT NUMBER OF CHANNELS"
3010 LPRINT "NUMBER OF CHANNELS";N
3020 LPRINT
3030 RETURN
4000 REM *** SUBROUTINE TO PRINT FILTRATE FLUX"
4010 LPRINT "FILTRATE FLUX";F1;"M/H",F;"M/S"
4020 LPRINT
4030 RETURN
5000 REM *** SUBROUTINE TO CLCULATE THE FANNING
5010 REM *** FRICTION FACTOR
5020 REM *** ROUTINE USES EQN PROPOSED BY NIKURADSE
5030 REM ** CHECK IF LAMINAR
5040 REM ***IF LAMINAR THEN FF=16/RE
5050 IF RE(J) >2100 THEN GOTO 5100
5060 FF=16/RE(J)
5070 RETURN
5100 REM ** FOLLOWING FORMULA
5110 REM *** 1/SQR(FF) = 4.0 LOG(RE*SQR(FF))-0.40
5120 REM ** USE A 1% CRITERIA FOR CONVERGENCE
5130 REM ** USE F=0.0005 AS AN INITIAL GUESS
5135 X1=0
5140 FI=0.0005
5145 X1=X1+1
5150 FG=4.0 *LOG(RE(J)*SQR(FI))/LOG(10) -0.40
5160 FF=(1/FG)[2
5170 E=FF-FI
5180 IF ABS(E) < 0.01 * FF THEN GOTO 5250
5190 FI=FF
5200 GOTO 5145
5250 REM ** VALUE OF F NOW RETURNED
5260 RETURN
```

```
6000 REM *** THIS SUBROUTINE CALCULATES THE
6010 REM *** PUMPING CAPACITY AT FLOWRATE Q
6020 REM
6030 REM *** FIRST FIND WHICH OF THE SIX
6040 REM *** SECTIONS IS APPLICABLE
6050 REM ***TEST IF Q < Q (MAX)
6060 IF Q > X(7) THEN GOTO 6300
6070 REM *** SEE WHICH OF 6 SECTIONS Q IS IN
6080 J=INT(Q/X(7)*6 +1)
6090 REM *** CAN NOW DECIDE WHICH VALUES TO APPLY
6100 PP=M(J)*Q+C(J)
6110 RETURN
6300 REM *** PUMP CANNOT DELIVER THIS MUCH FLOW
6310 CLS
6320 PRINT @460,"PUMP CANNOT DELIVER REQD FLOW"
6330 STOP
8000 REM *** SUBROUTINE TO CALCULATE AVERAGE VELOCITY
8010 REM *** AND AVERAGE PRESSURE DIFFERENCE
8020 REM *** USE GEOMETRIC MEANS
8030 VA=V(1)
8035 PE=0
8040 PA=PP
8050 FOR J=2 TO 11
8060 VA=VA*V(J)
8070 I=J-1
8075 PE=PE+P(I)
8080 PA=PA*(PP-PE)/10000
8090 NEXT J
8100 PA=PA[0.1
8105 PA=PA*10000
8110 VA=VA[0.1
8120 RETURN
9000 LPRINT "CALCULATED VALUES"
9010 VV=(INT(V(1)*10+0.5)/10)
9020 LPRINT TAB(20)"VELOCITY AT CHANNEL ENTRANCE";VV;" M/S"
9030 VV=(INT(V(11)*10+0.5)/10)
9040 LPRINT TAB(20)"VELOCITY AT CHANNEL EXIT     ";VV;" M/S"
9042 VA=(INT(VA*10+0.5)/10)
9045 LPRINT TAB(20)"AVERAGE VELOCITY             ";VA;" M/S"
9050 LPRINT
9060 LL=(INT(L*100+0.5)/100)
9070 LPRINT TAB(20)"EFFECTIVE CHANNEL LENGTH REQD"
9080 LPRINT TAB(20)"FOR 20% REJCTION             ";LL
9090 LPRINT
9100 PZ=(INT(PD+500)/1000)
9105 PC=PA/6.89476E03
9110 PY=PD/6.89476E03
9115 PC=(INT(PC*10++0.5)/10)
9120 PY=(INT(PY*10+0.5)/10)
9125 PA=(INT(PA+500)/1000)
9130 LPRINT TAB(20)"PRESSURE DROP ALONG CHANNEL";PZ;"KPA"
9140 LPRINT TAB(20)"                           ";PY;"PSI"
9142 LPRINT TAB(20)"AVERAGE PRESSURE DIFFERENCE";PA;"KPA"
9144 LPRINT TAB(20)"                           ";PC;"PSI"
9150 QQ=Q/7.576E-05
9160 QQ=INT(QQ*10+0.5)/10
9170 LPRINT
9180 LPRINT TAB(20);"PUMP OUTPUT   ";Q;"M**3/SEC"
9190 LPRINT TAB(20);"              ";QQ;"GPM"
9200 LPRINT
9210 LPRINT
9220 LPRINT
9230 RETURN
10000 REM *** SUBROUTINE TO INPUT PUMP DATA
10010 REM *** SLOPES AND INTERCEPTS OF
10020 REM *** PUMP CURVE ARE INPUTED FROM TAPE
10030 REM *** THESE DATA REPRESENT A LINEAR INTERPOLATION
10040 REM *** OF THE CHARACTERISTIC CURVE OF THE PUMP
10045 REM *** THE CROSSOVER FLOWRATES ARE ALSO INPUT
10050 DIM M(6),C(6),X(7)
10060 PRINT "THE PROGRAM NOW WANTS THE PUMP DATA"
10070 PRINT "ADVANCE THE TAPE TO '065' ON"
10080 PRINT "THE TAPE COUNTER AND PRESS PLAY"
10090 PRINT "             PRESS ANY KEY TO CONTINUE"
10100 REM
10110 IF INKEY$="" THEN 10100
10120 FOR J=1 TO 6
10130 INPUT £-1,M(J),C(J)
10140 NEXT J
10150 INPUT £-1,X(1),X(2),X(3),X(4),X(5),X(6),X(7)
10160 RETURN
```

APPENDIX 5

PROTOTYPE DRAWINGS.

This appendix presents the detailed drawings of the prototype filter, as developed in the Prototype Design (Section 11).

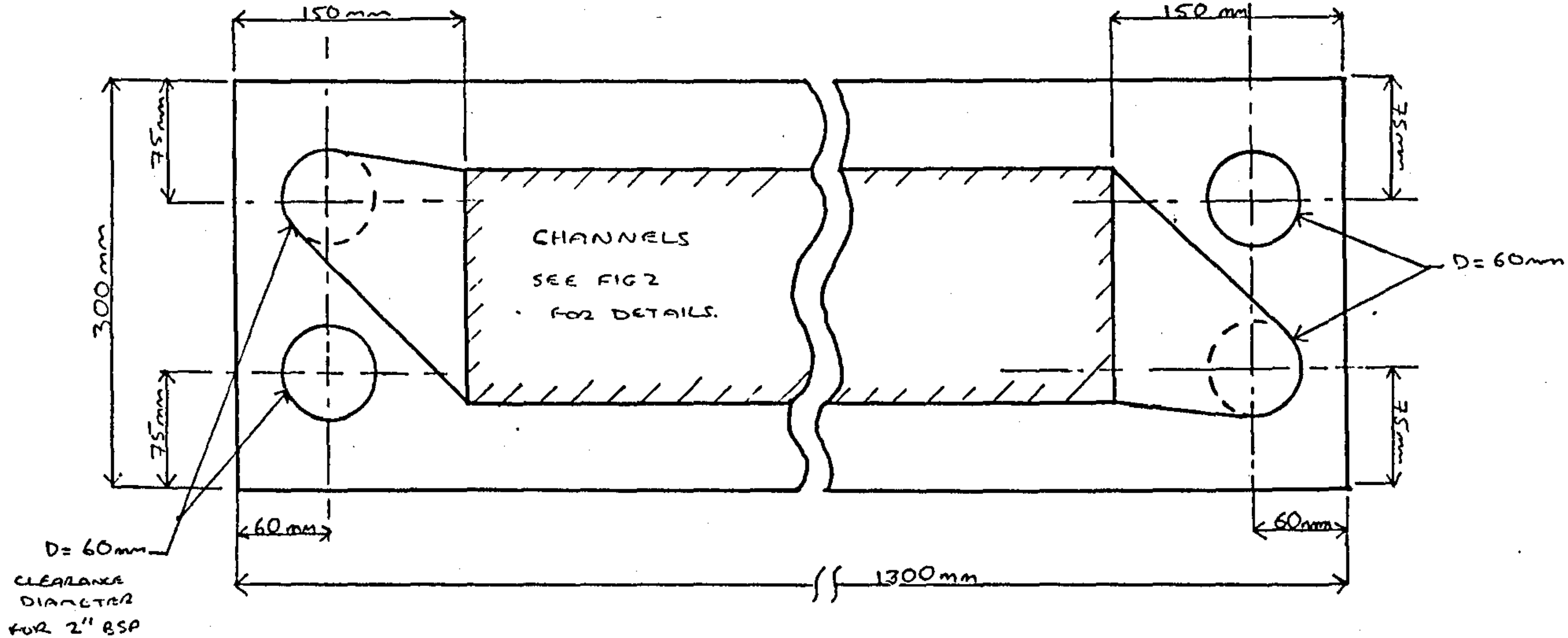
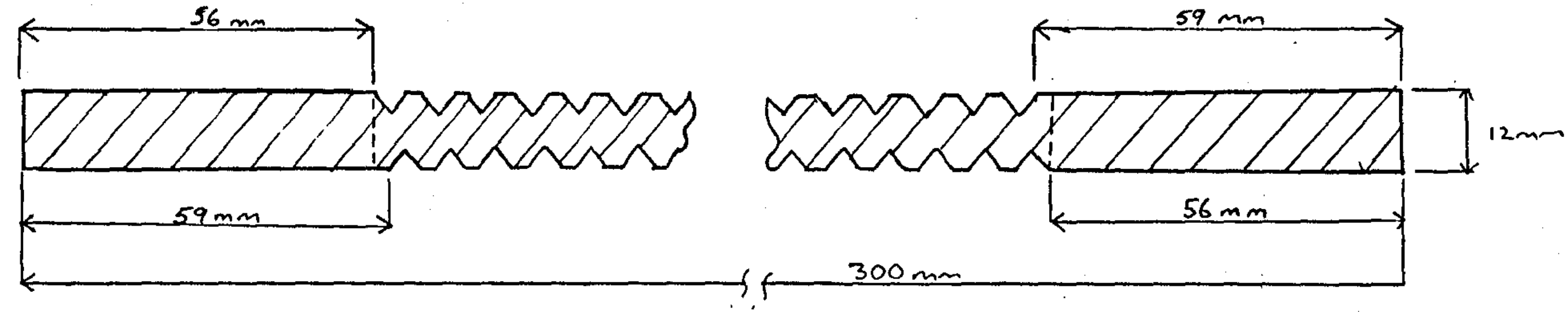
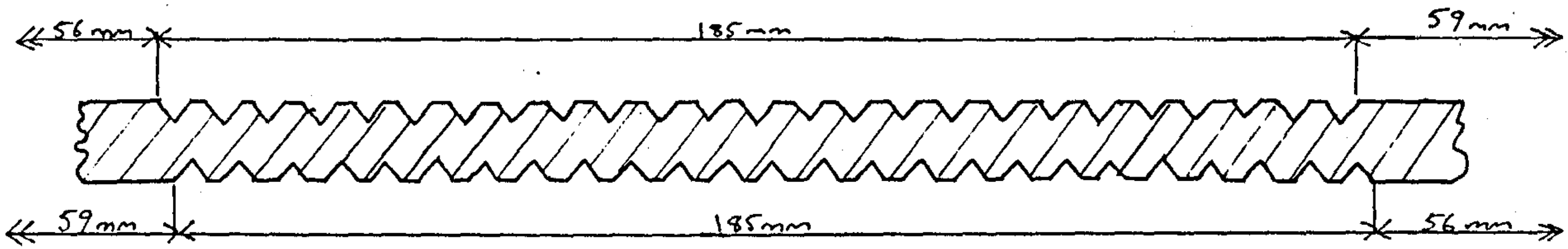
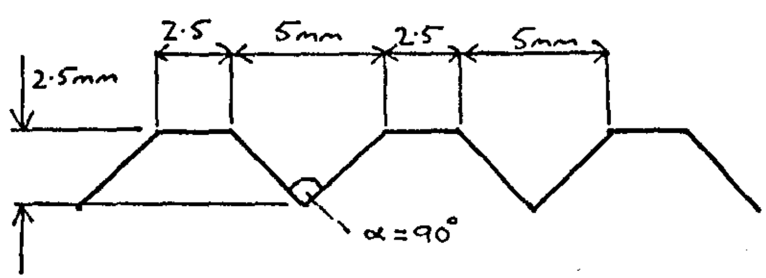


FIG 1. CROSSFLOW PLATE
SCALE 1:4
No Reqd: 4



SECTION AA
SCALE 1:1





DETAIL OF
CHANNELS
SCALE 4:1

FIG 2.
DETAILS OF
CROSSFLOW PLATE

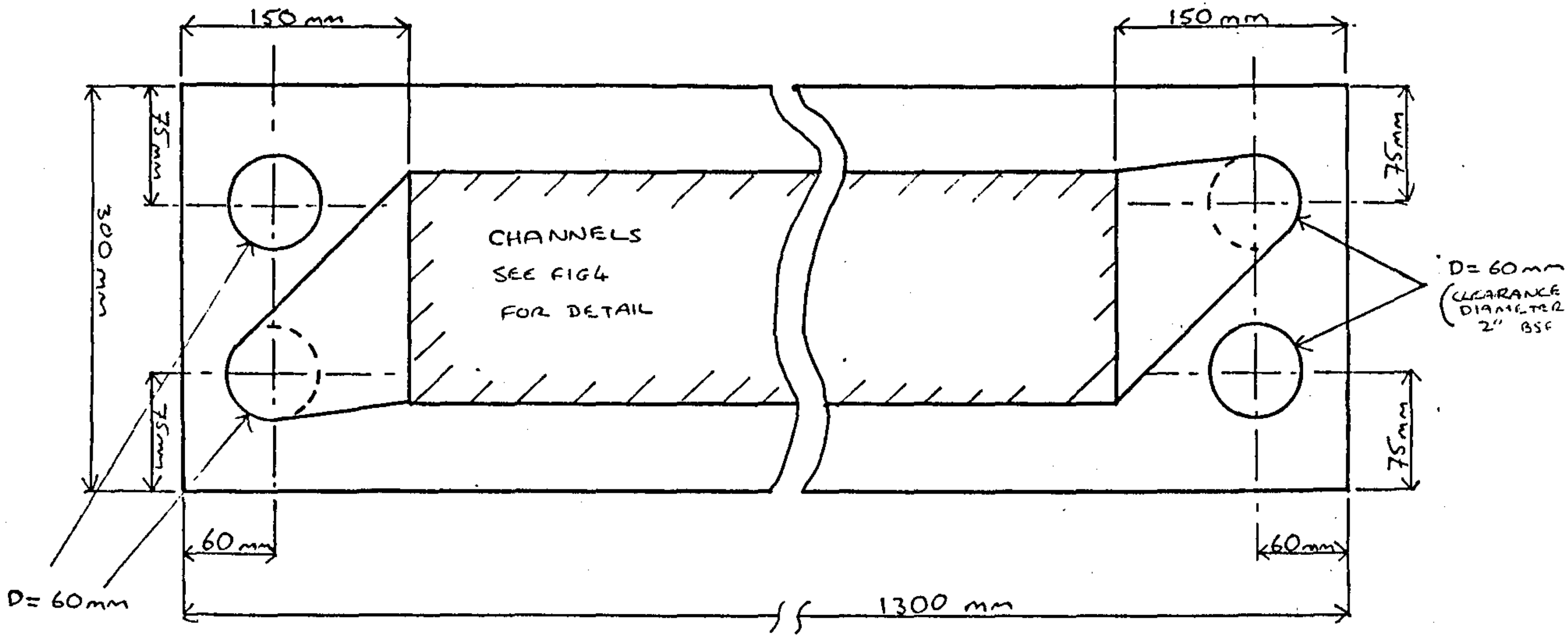
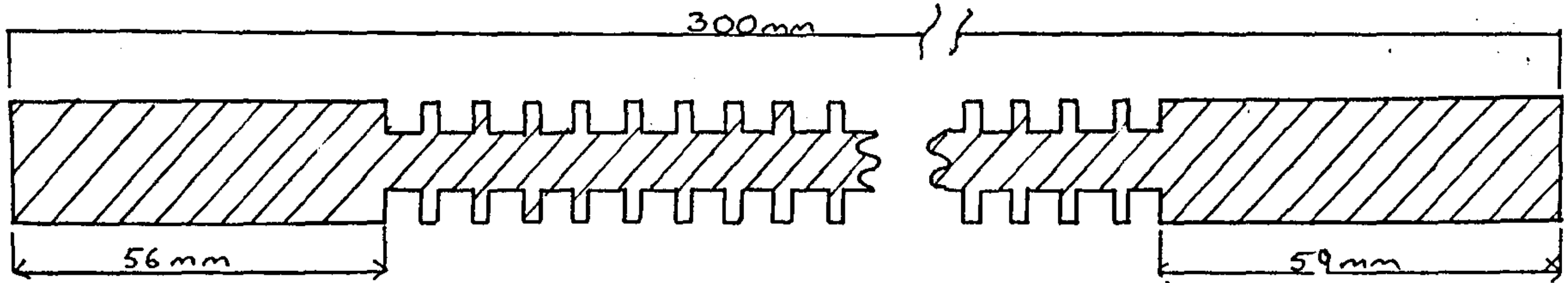


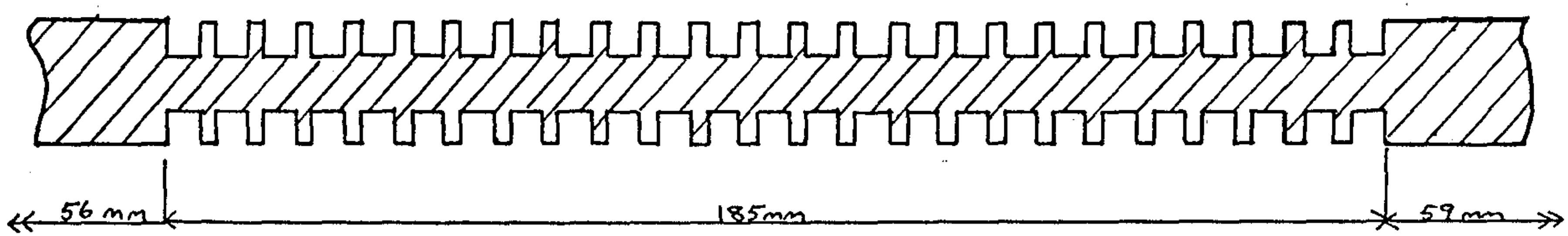
FIG 3: FILTRATE PLATE

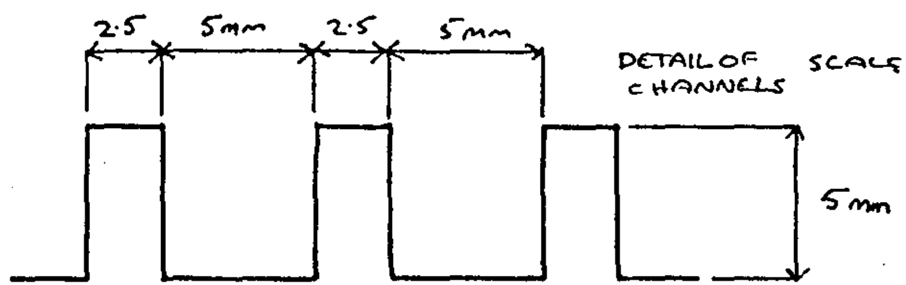
SCALE 1:4

No. Req'd: 3



SECTION THRU
FILTRATE PLATE
SCALE 1:1.





DETAIL OF CHANNELS SCALE 4:1

FIG.4
DETAILS OF FILTRATE PLATE

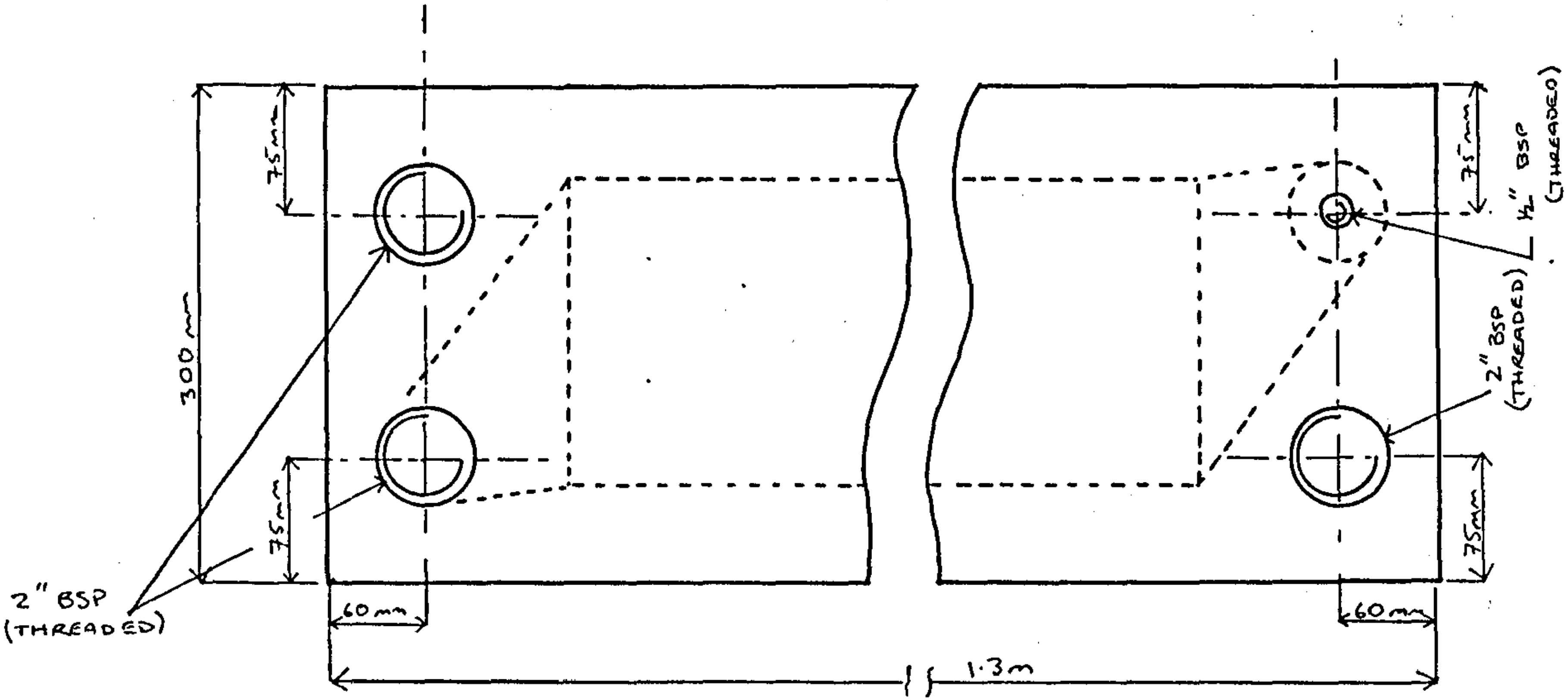


FIG 5A. TOP PLATE OF FILTER AS SEEN FROM ABOVE

SCALE 1:4.

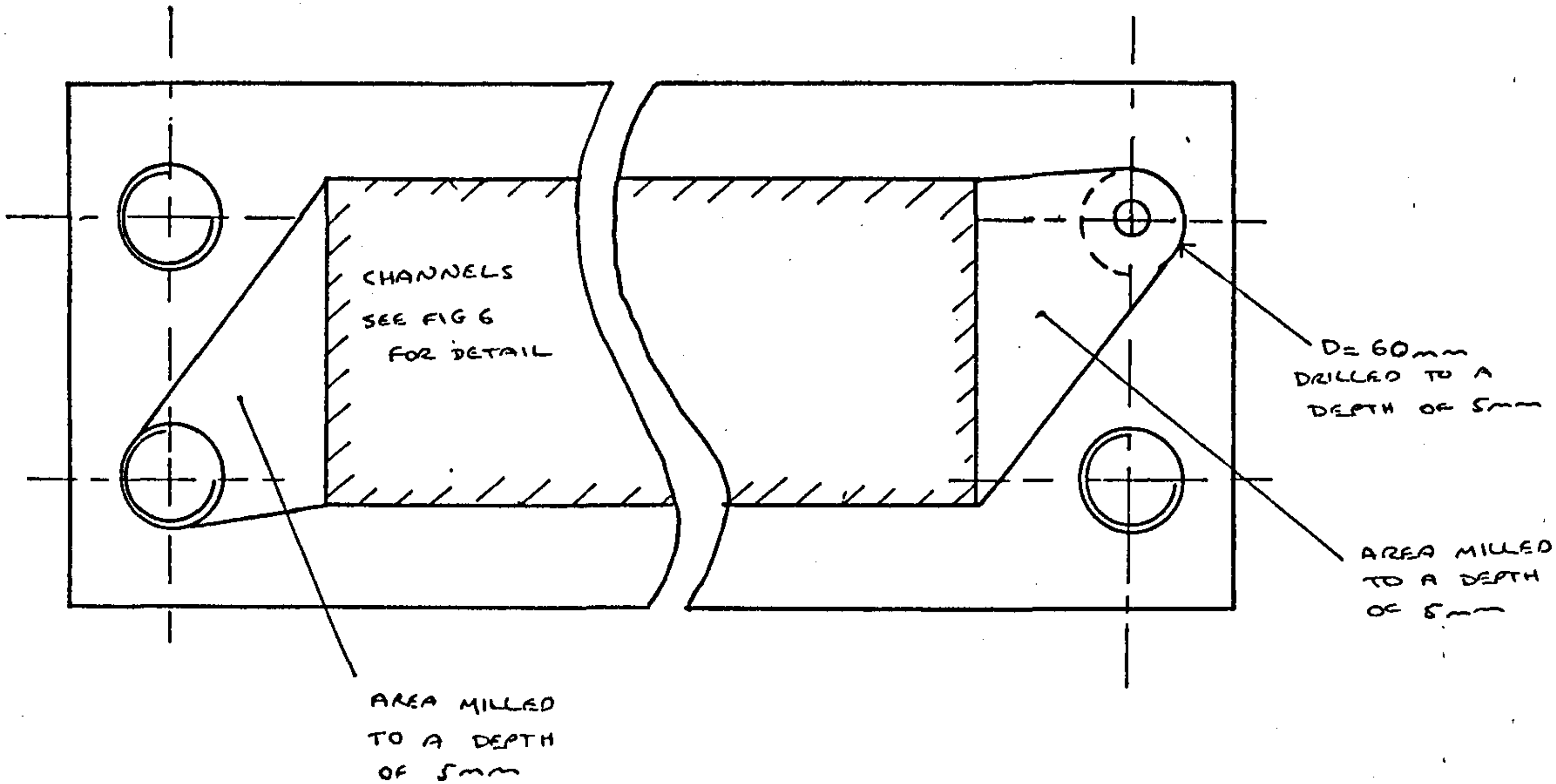
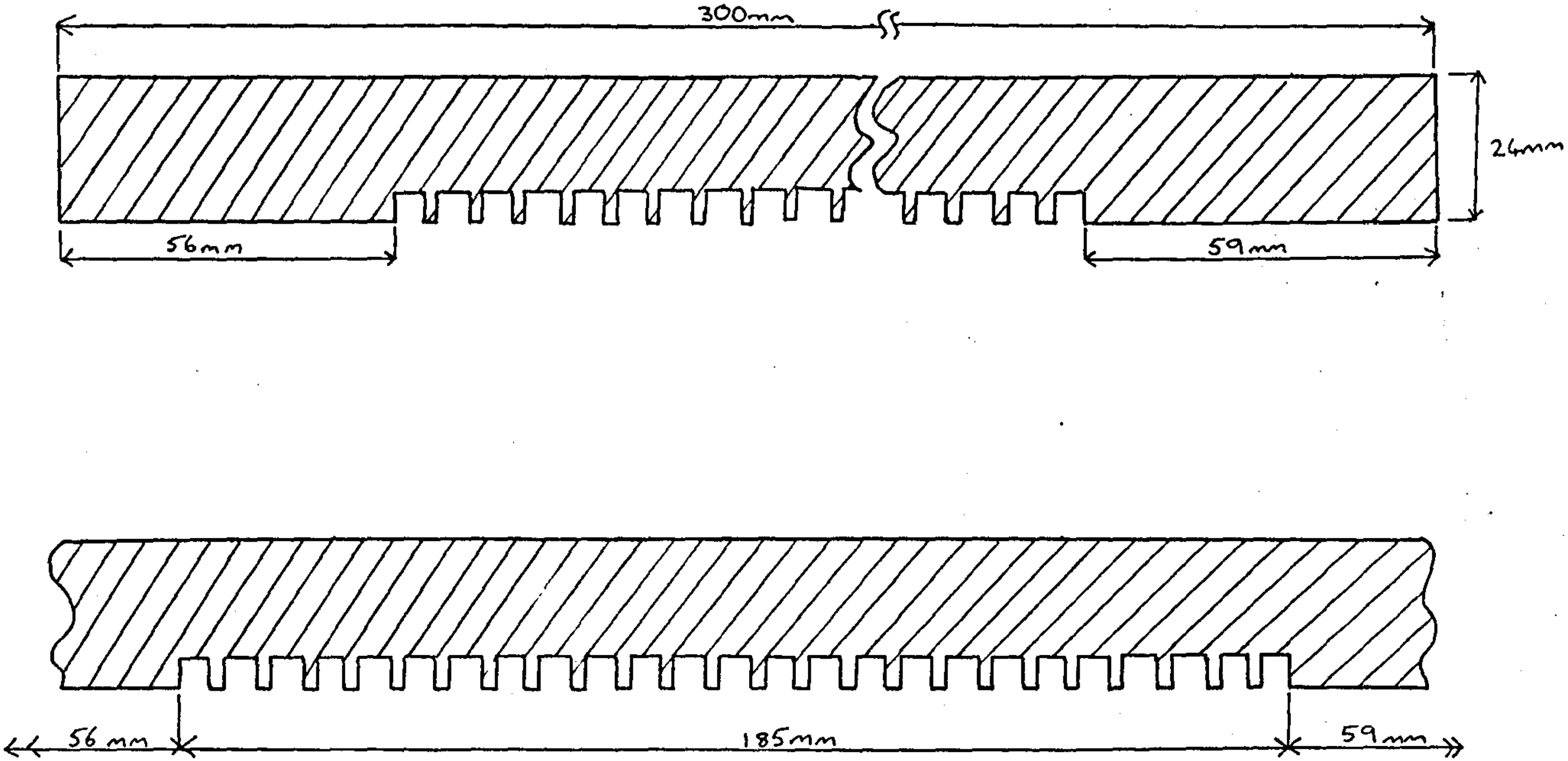


FIG 5B. TOP PLATE AS
VIEWED FROM BELOW.

REFER 5A FOR OTHER DIMENSIONS



SCALE 1:1

REFER FIG 4 FOR DETAIL OF CHANNELS.

FIG 6. CROSS-SECTION OF TOP PLATE.

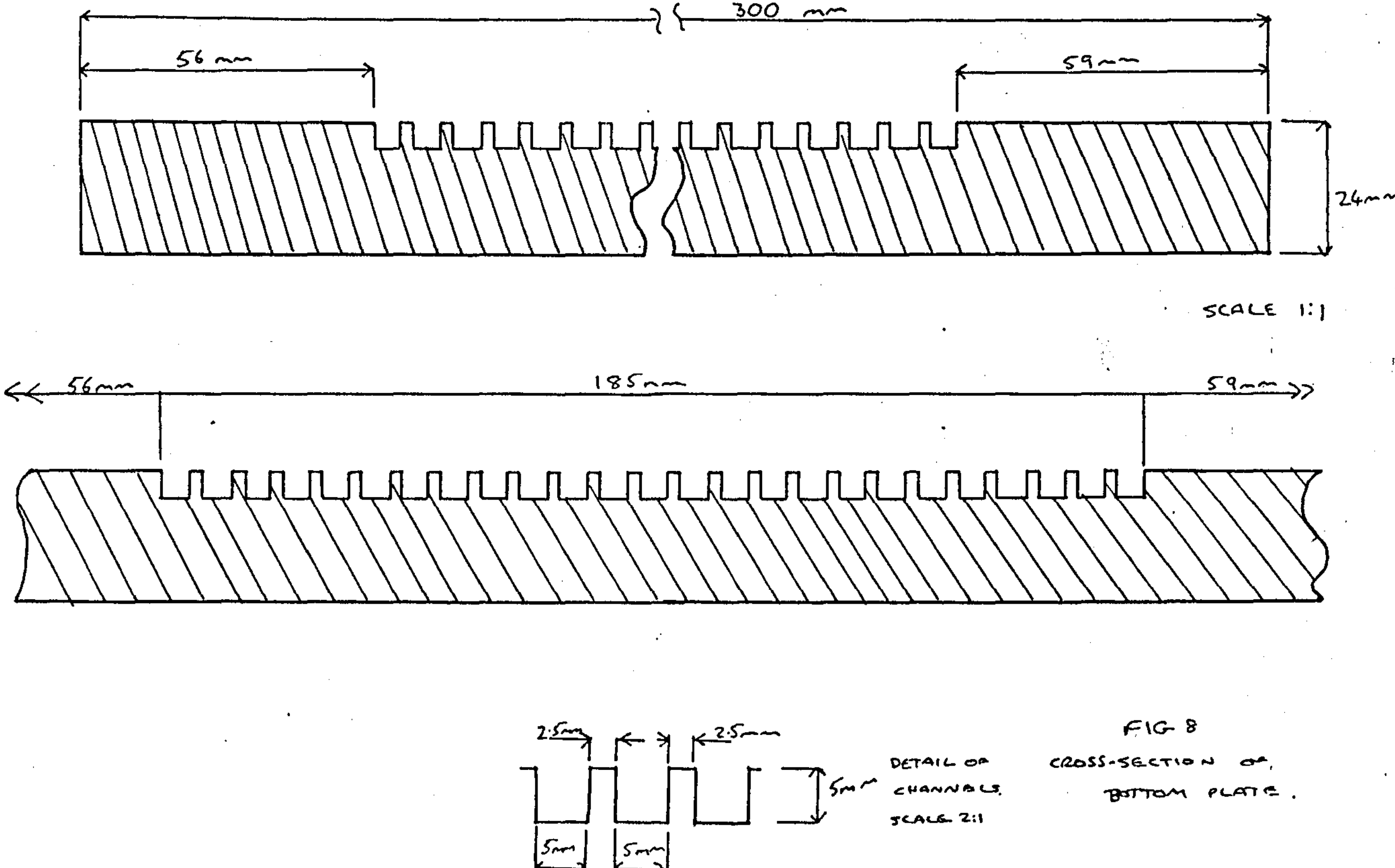


FIG 8
CROSS-SECTION OF BOTTOM PLATE.