CRANFIELD UNIVERSITY

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CHEMICAL CLEANING OF POTABLE MICROFILTRATION AND ULTRAFILTRATION MEMBRANES

SCHOOL OF APPLIED SCIENCES
DEPARTMENT OF SUSTAINABLE SYSTEMS
CENTRE FOR WATER SCIENCE

EngD THESIS

2009

Supervisor: Prof. Simon J. Judd
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ABSTRACT

Concerns over possible waterborne disease forced drinking water supply companies in England and Wales to adopt microfiltration and ultrafiltration technologies rapidly. MF and UF membrane plants are designed to produce water of a consistent quality regardless of throughput and fluctuations in the feedwater quality. To operate well they need to maintain flux and balance the rate of fouling, and chemical cleaning performance is critical to this. Giant steps have been taken into characterising the foulants scientifically in the last few years while cleaning is reactive and ad hoc. This thesis explores the basis for a corresponding cleaning science for the technology to develop quantitively.

Cleaning performance was defined in terms of a response to combinations of explanatory variables in a materials limited cleaning envelope. The study focussed on applying variations of cleanant concentration, applied temperature and soak times to a variety of membranes fouled with different waters and regimes. An experimental design was developed and applied consistently to a number of different sampled sites; allowing an optimised recovery from the polynomial expressions for each treatment, through factorial analysis of the data.

The size and variety of the data set analysed allowed comparison and quantification of the different deviations from optimal cleaning response. This effect was seen to vary temporally and with operating regime and the methods usefulness as a practical tool in the membrane plant lifecycle was considered.

Cost evaluation of the variation in cleaning response showed that sub-optimal cleaning costs and energy use may be significant and the thesis also illustrated how module geometry affects initial cake deposition and thus cleanability. By demonstrating the potential for cleaning factor analysis, the potential for a combined heuristic and predictive cleaning control science is possible, but will need new strategies to manage technology change.

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ABBREVIATIONS

AFM	Atomic force microscopy
Allochthonous	Originating from outside the water body
Autochthonous	Originating from within the water body
AMP	Asset Management Plan: A five year capital spending plan set by the UK water Regulator OFWAT to maintain and improve water
ANOVA	Analysis of variance
AOC	Assimilable organic carbon
AOM	Autochthonous/algal organic matter with extra/intra-cellular macromolecules
AOP	Advanced Oxidative Processs
ASTM	Technical standards organisation. Formerly the American Society for Testing and Materials. Now known as “ASTM
AWWA	American water works association
AwwARF	American water works association research foundation
BBD	Box Behnken Methodology: a response surface method suited to experimental design about a central matrix point.
CEB	Chemically enhanced backflush
CIP	Clean in Place
Conductivity	Ohms, Surrogate measure for dissolved solids content or chemical purity of water (MΩ
CPT’	Cleaning conditions within a specified experimental design envelope for a treatment.
CT	Capillary Tube: membrane where feed is through lumen.
DEFRA	UK Government Department of Environment Food and Rural Affairs
DI	Demineralised water- Water taken from an ultrapure source usually RO quality. See conductivity/TDS
DOC	Dissolved organic carbon
DWI	UK government’s drinking water inspectorate
EA	UK environment agency

EDTA	Ethylene diamine tetra-acetic acid
EDX	Energy dispersive x-ray analysis detection system
EPS	Extracellular polymeric substances
EU	European union
H_2	Hydrogen gas
H_2O_2	Hydrogen peroxide
$H_2\ SO_4$	Sulphuric Acid
H_3PO_4	Phosphoric acid
HCl	Hydrochloric acid
HCl	Hydrochloric acid
HF	Hollow fibre
HPSEC	High performance size exclusion chromatography
IC	Inorganic carbon
LRV	Log Removal Value- the log of the ratio of the inlet to outlet concentrations
MBR	Membrane bio reactor
MF	Microfiltration
MIT	Membrane integrity test
MLD	Megalitres per day (thousands of cubic metres per day)
MLSS	Mixed liquor suspended solids
molar	Unit of concentration. Amount of solute per unit volume of solution, $M = 6.022 \times 10^{23}$ mol per litre
MW	Molecular weight
MWCO	Molecular weight cut off
NaOCl	Sodium hypochlorite, $mg.L^{-1}$, %w/v, ppm, $mol.L^{-1}$

NaOH	Sodium hydroxide, $mg.L^{-1}$, %w/v, ppm, $mol.L^{-1}$
NH_4HF_2	Ammonium biflouride, $mg.L^{-1}$, %w/v, ppm, $mol.L^{-1}$
NOM	Natural organic matter
NPOC	Non-purgeable organic carbon
NTU	Turbidity, nephelometric turbidity units
OFWAT	Office of Water Regulator UK
PAC	Powdered activated carbon
PDT	Pressure decay test
PE	Polyethylene
PES	Polyethersulphone
Polydadmac	Polydiallyldimethylammonium chloride
PP	Polypropylene
PS	Polysulphone
PTFE	Polytetrafluoroethylene
PVDF	Polyvinyldiethylene
PVDF	Polyvinylidene difluoride
PZC	Point of zero charge
RO	Reverse osmosis
RS232	Hardware and software protocol for communications between peripheral apparatus and a personal computer.
RSM	Response Surface Methodology: An experimental design technique to explore the several explanatory variables.
SDWA	Safe drinking water act
SEM	Scanning electron microscope
SMBR	Submerged membrane bio reactor

SUVA	Specific ultraviolet absorbance
TMP	Transmembrane pressure
TOC	Total organic carbon
UF	Ultrafiltration
UKWIR	UK Water industry Research
UV	Ultraviolet
V	Volume
WaSCs	Water and sewage companies
WoCos	Water only companies
WQZ	Water quality zone

NOTATION

$\%R_{abs}$	Percentage absolute permeability recovery	(%)
$\%R_{rel}$	Percentage relative permeability recovery	(%)
$\%R_v$	Percentage permeability recovery from virgin	(%)
μ	Viscosity	$(N.s^{-1}.m^{-2})$
A_m	Membrane area,	(m^2)
C’	Concentration of cleanant	$(mg.L^{-1}$, %w/v, ppm, $mol.L^{-1})$
d_{10}, d_{50}, d_{90}	Particle diameter below which lies a % of the total volume of all particles in the sample	(µm)
Di	Effective internal diameter	(m)
dP/dt	Rate of TMP increase	$(mbar.min^{-1})$
g	Gravity constant	$(m.s^{-2})$
hrs.	Hours	(hrs)
J	Flux	$(L.m^{-2}.h^{-1})$
J	Design flux,	$(L.m^{-2}.h^{-1})$
J_b /Q_b,	Backwash flux/rate	$(L.m^{-2}.h^{-1})$
Jc	Critical flux	$(L.m^{-2}.h^{-1})$
J_w	Clean water flux measured at normalised temperature and pressure with DI water	$(L.m^{-2}.h^{-1})$
J_w	Clean water flux normalised to 20°C	$(L.m^{-2}.h^{-1})$
K	Permeability	$(L.m^{-2}.h^{-1}.bar^{-1})$
kDa	Kilodaltons	
K_f	Final permeability after cleaning	$(L.m^{-2}.h^{-1}.bar^{-1})$
kHz	Frequency	(s^{-1})

K_i	Initial pre-cleaning permeability	$(L.m^{-2}.h^{-1}.bar^{-1})$
K_v	Virgin clean water permeability of membrane	$(L.m^{-2}.h^{-1}.bar^{-1})$
min.	Minutes	(minutes)
P	Power	(Watts, $J.s^{-1}$)
P’	Soak time of clean	(minutes)
P_{ave}	Average pressure	(mbar)
PDR	Pressure decay rate	(kPa min^{-1})
pH	Log scale of hydrogen ion concentration (acidity)	
Q_m	Design throughput,	$(L.s^{-1})$
R	Resistance	(m^{-1})
Rc	Hydraulic resistance of the cake layer	(m^{-1})
Rf	Hydraulic resistance of the fouling layer	(m^{-1})
Rm	Hydraulic resistance of the membrane	(m^{-1})
R_m	Clean water resistance,	$(L.m^{-1})$
$Rres$	Residual resistance	(m^{-1})
r_v	Cleanant Ratio-CIP volume per membrane area	$(m^3.m^{-2})$
T’	Temperature of cleaning solution	(°C)
t_b	Backwash interval	(minutes)
t_{bb}	Backwash duration	(seconds)
t_c	CIP Cleaning interval	(hours)
t_{cc}	CIP Cleaning duration	(minutes)
TDS	Total dissolved solids	$(mg.L^{-1})$
temp.	Temperature	(°C)

T_f	Average feed temperature	(°C)
TMP or ΔP	Trans membrane pressure	(m H_2O, bar.g or kPa)
vol.	Volume	(m^3)
X_a	Fouling rate of specific cake resistance removable by backwashing	($m.min^{-1}$)
X_b	Fouling rate of specific cake resistance removable by chemical cleaning process	($m.min^{-1}$)
X_c	Fouling rate of specific cake resistance not removed- irreversible build up.	($m.min^{-1}$)
α	Specific cake resistance	($m.kg^{-1}$)
η	Power supply efficiency conversion ratio	
ηm	Nanometres	m
ρ	Density	($kg.m^{-3}$)

CHAPTER 1 INTRODUCTION

1.1 BACKGROUND

Waterborne zoonotic diseases such as from *Cryptosporidium Parvum.* oocysts are a serious threat to public health; epidemiologists are ever conscious of the risks from municipal potable water treatment plants, especially through unpredicted source water from climatic change (Gajadhar and Allen , 2004). Prior to the 1999 Water Regulations (DWI, 1999), 20% of plants in England and Wales (UK) were deemed by the UK Government's Drinking Water Inspectorate (DWI) to be at risk from *cryptosporidium* contamination. Three options were provided for the management of such plants: (1) supply shut down, (2) installation of an approved treatment process capable of continuous removal of particles greater than 1 μm in diameter or (3) where risk is manageable, implementation of a continuous particle monitoring scheme (Morris and Whitehead, 2006). From 1999 to 2004 forty two membrane treatment plants were proposed for Option 2, since the DWI criteria at this time favoured barrier processes rather than inactivation of viable oocysts using ultraviolet irradiation. It was stipulated that these plants continuously demonstrated a 4-5 log removal value (LRV, the log of the ratio of the inlet to outlet concentrations) of oocysts-sized particles. This stipulation by the DWI was critical in establishing membrane processes in the UK and the use of microfiltration and ultrafiltration increased in this period (see Figure 1.1.).

Microfiltration (MF) and ultrafiltration (UF) are pressure-driven membrane processes that separate sub-micrometre particles (>0.1 μm diameter for MF and >1000 Daltons molecular weight cut off for UF) from a feed stream via size exclusion, or sieving (AWWA, 2000). A number of suitable hollow fibre-based membrane technologies are approved by the UK DWI, the materials concerned being polypropylene (PP), polyethersulphone (PES) and polyvinyldiethylene (PVDF). The hollow fibres (HF) are fabricated into modules of 20-30 thousand square metres of membrane filtration area. The treatment plants are specified based on the retention properties of the fibres, which retain and absorb particulates and colloidal materials (collectively referred to as foulants), during filtering operations. Routine physical and chemical cleaning then needs to be carried out to maintain the membrane permeability, i.e. the flow through the membrane per unit membrane area per unit applied pressure.

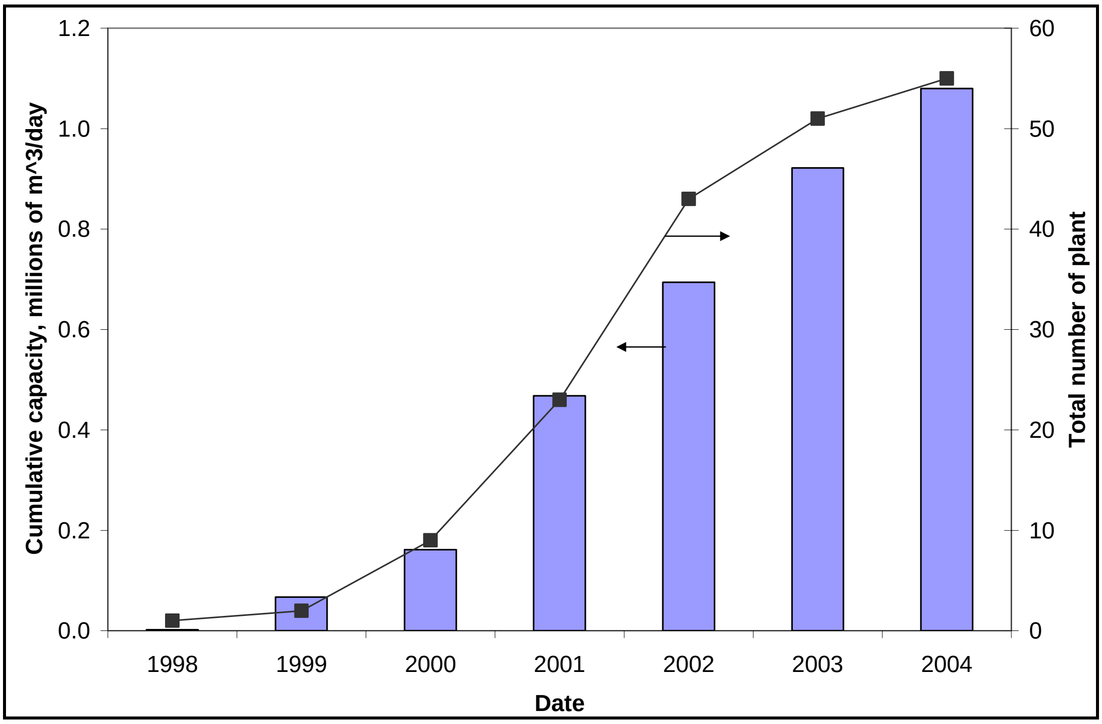

Figure 1.1 Cumulative capacity and number of UK MF/UF Potable Membrane Plants in the UK 1998 to June 2004 (from Judd, 2004)

In response to the difficulties experienced by municipal water companies in England and Wales in maintaining a sustainable operational flux and acceptable LRVs on recently commissioned MF/UF potable plants, a report was commissioned by UKWIR (UK Water industry Research) and the AwwARF (American Water Works Association Research Foundation) to identify trends and best practice (Judd, 2004, see also Appendix A.1). There were evident differences in specified cleaning and operating regimes between the UK and the US where, for the latter plants, pretreatment was more common, design flux lower and cleaning longer in duration, more intensive and at elevated temperatures. The selection of cleaning protocols was as prescribed by the membrane supplier, generalised from pilot test data and/or adjusted on an ad hoc basis rather than optimised through any rigorous process to account for changing feedwater and operating conditions (Liu et al., 2001; Presdee et al., 2006). Research needs were advanced by AwwARF (Escobar et al., 2005) to minimise these "trial-and-error" approaches to cleaning, advocating the identification of specific chemical and physical conditions that influence the cleaning efficiency and efficacy as well as quantification of the effects of cleaning factor variables..

1.2 AIMS AND OBJECTIVES

The work reported concerns the study and quantification of the impact of membrane cleaning regimes on permeability recovery. Whilst the original programme was envisaged to include membrane integrity, it became apparent from a literature study that integrity and cleaning demanded separate research relating to membrane materials, operating pressure and cleaning cycling effects which could not be encompassed in the current study. Similarly, early experiments verified work by Masselin (2001) demonstrating that ultrasonic cleaning, another aspect originally to be incorporated into the research programme, caused damage to the polymer membranes. The following specific research objectives were thus subsequently identified:

1. To assess the status of membrane cleaning
2. To determine the impact of flow and fouling distribution variation with membrane geometry for dead end hollow fibre filtration.
3. To devise a protocol for measuring the efficacy of cleaning across the envelope of cleaning factor parameters.
4. To determine the impact of membrane configuration, material, plant design and operation on the optimum permeability recovery.
5. To determine the cost benefit of cleaning.
6. To assess the influence of economic regulation influences on the receptivity of the organisation with respect to inward technology transfer.

1.3 THESIS STRUCTURE

The Engineering Doctorate addresses a research topic of direct relevance to an industrial sponsor; a non-technical component, essentially commercial or businessoriented, of the study provides up to one fifth of the thesis content. To this end, the subject area of management of change was included as the final objective, and the study include in the thesis.

A programme of site visits was conducted, collecting operational data and sample fouled modules, and a test protocol with experimental methodology was developed. The sites were selected so as to capture the variety of applications and operating regimes for MF/UF plants in the UK. Data from the survey is provided in Appendix A B and C.

The thesis is presented in paper format. All papers were written by the first author, who also conducted all experiments and provided the intellectual lead in analysing and interpreting the data. The assistance of the industrial sponsors in providing site information is gratefully acknowledged, as is the practical assistance of Jose Poyotas of the University of Granada, Dr Riyadh Al Zubaidy from Baghdad University and Radouan Najhi, a Cranfield University MSc placement student.

A review of the available literature on MF and UF cleaning is presented in Chapter 2. This confirms the paucity of publications in the science of cleaning as relating to potable water membrane filtration. The paper includes an analysis of general cleaning models from the literature and reviews major cleaning mechanisms, and at the time of publication of this thesis is in editing for publication by Separation and Purification Technology, Porcelli, N. and Judd, S.J.: *Chemical Cleaning Of Potable Water Membranes: A Review.*

In Chapter 3 results are presented of a case study carried out to optimize chemical cleaning of modules taken from heavily fouled hollow fibre (HF) MF PVDF membranes from an operating plant. The paper includes a description of the apparatus and cleaning optimization protocol, and has been published in Environmental Technology*,* **30**(6)*, 629-639:* Porcelli, N., Hillis, P. and Judd, S. (2009) Microfiltration Membrane Plant Start Up: A Case Study with Autopsy and Permeability Recovery Analysis.

Chapter 4 was the result of a study undertaken to gather permeability recovery for open pore MBR membranes while examining how the profile for recovery varied depending on the position from where the fibres were removed in the reactor tank. The work examined the impact of cleaning with heterogeneous fouling effects and how homogeneous flow distribution influences cleaning recoverability. (Published in Proceedings of Aachen Water and Membrane Conference, 31st October 2007: Porcelli, N., Poyotas, J. and Judd, S. (2007), *Effects of Variable Flow and Fouling on MBR Cleaning*).

In Chapter 5 a statistical analysis of the results of all the permeability recovery tests carried out throughout the duration of the project is presented. The effect of factor variation on the optimum recovery in terms of cleaning chemical concentration, cleaning temperature and soak time is quantified for all the plant and reagent combinations. The paper is accepted to be published in the Journal of the American Water Works Association: Porcelli, N., and Judd, S. (2009) *Impact of cleaning protocol on membrane permeability recovery: a sensitivity analysis*.

Chapter 6 follows the evident variability in cleaning parameters from Chapter 5 with a study of their effect on operational costs. A cost model based on transient trans membrane pressure (TMP= ΔP m, the headloss developed as resistance (R) builds through foulant accumulation) examined the cost deviation from optimal recovery for a number of cleaning factor envelopes and operational strategies based on Clean in Place (CIP) and backwash sequencing to control. The paper has been submitted (awaiting editorial comment) to Water Resources Journal : Porcelli, N., and Judd, S., *Chemical cleaning of potable water membranes: the cost benefit of optimisation*.

Chapter 7 concerns a mathematical model describing how axial flow and headloss profile changes with fibre geometry. An MF and UF membrane are compared with a surrogate foulant to verify the model, and the paper is in preparation for submission to Water Research: Al, Zubaidy, R., Porcelli, N., Judd, S., *Modelling hollow fibre flow and headloss distribution effects*.

Chapter 8 uses factor analysis to compare regulatory econometric measures with a survey of the implementation of membrane plants between 1999 and 2004. The paper is relevant to ongoing efforts in the UK water industry to improve the rate of inward transfer of new technologies, and has been submitted to Water and Environment Journal (CIWEM): Porcelli, N. and Judd, S., *Innovation in the UK water industry and membrane plant implementation in AMP 3*).

The thesis concludes with a summary of the conclusions and includes some recommendations for future work, particularly the study of integrity and cleaning effects and the development of uniform standards for cleaning protocol evaluation. A summary of the workplan is given below. (Figure 1.2)

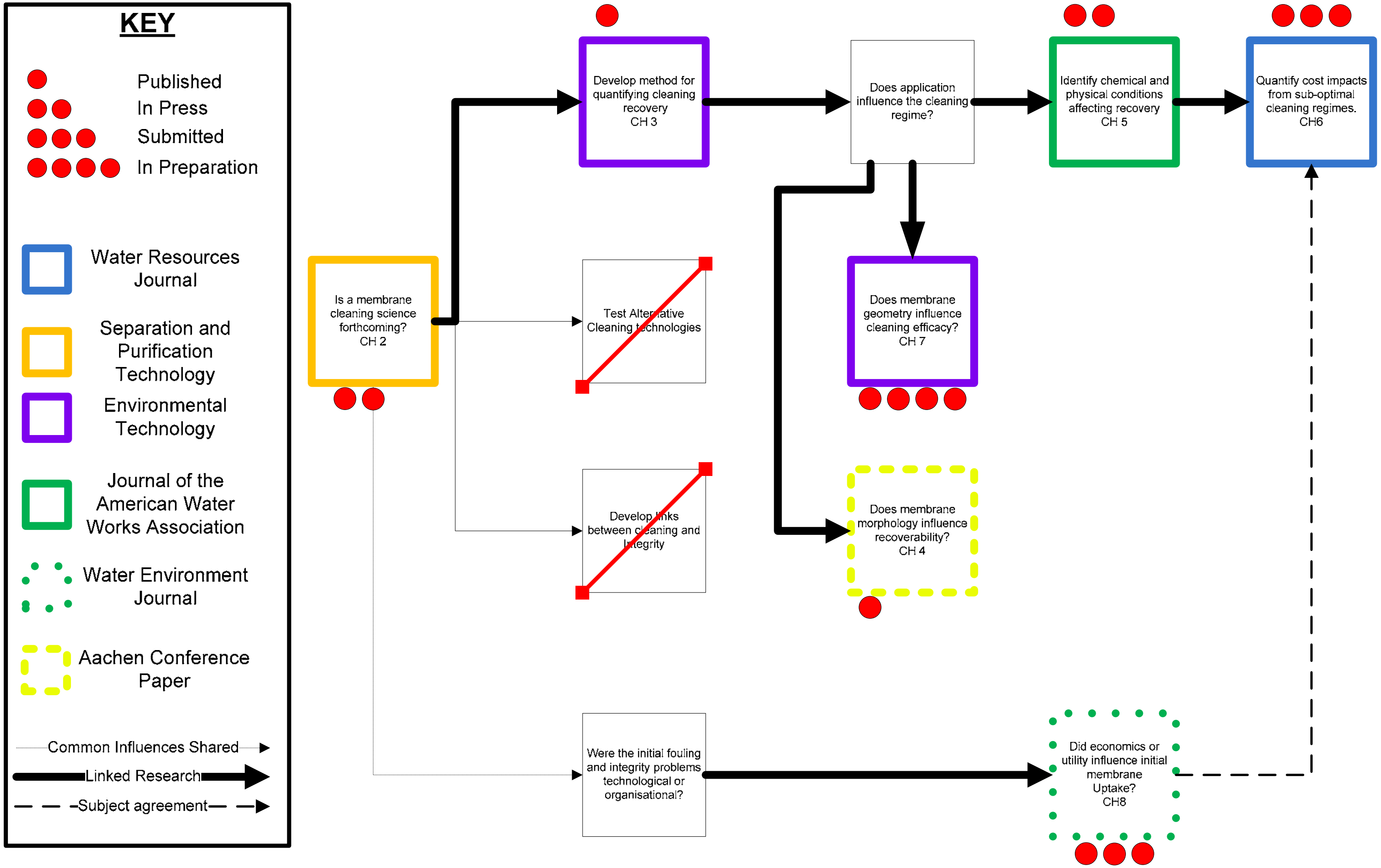

Figure 1.2 **Thesis Workplan**

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CHAPTER 2 CHEMICAL CLEANING OF POTABLE WATER MEMBRANES: A REVIEW

Porcelli, N. and Judd, S.J., Chemical Cleaning Of Potable Water Membranes: A Review Submitted to *Separation and Purification Technology*

2.1 INTRODUCTION: MEMBRANE FOULING AND CLEANING THEORY: HISTORICAL PERSPECTIVE

There is considerably less literature dedicated to membrane cleaning than to fouling (Koltuniewicz & Drioli, 2008; Judd, 2004), particularly regarding theoretical development. The earliest particle fouling modelling studies were based on colloidal particle trajectory analysis with Darcian relationships and Hagen Poiseulle flow (Green and Belfort, 1980). These models assumed homogeneous fouling and pore size and distribution characteristics and did not relate well to field observations. Fane and Fell (1987) discussed the properties of ultrafiltration (UF) and microfiltration (MF) membranes and related these to flux decline behaviour, but the discussion of fouling control mechanisms was more strategic rather than operational (Fane and Fell, 1987). With uptake of water treatment plants increasing, researchers began to relate permeation to the characteristics of natural organic matter (NOM), a key component of most source waters, with many feed water characterisation studies aimed at identifying and understanding the behaviour of major NOM foulants (Krasner et al., 1996; Cho et al., 1998; Fan et al., 2001). This work has extended through prediction of fouling Yuan & Zydney, 1999 and 2000), amelioration through pretreatment (Carroll et al., 2000), and surface modification (Bacchin et al., 1995). Early investigations revealed inconsistencies between fouling from natural raw source waters and that predicted from model or analogue studies (Logan and Jiang, 1990). It has since become apparent that not only NOM but also extracellular polymeric substances, or EPS (Hodgson et al., 1993; Baker and Dudley 1998) are possibly the key foulants where the complexity of organism behaviour and refouling tendencies makes cleaning science development critical to operation.

With increasing membrane installations, and commensurately more available information, modelling based on fouling mechanisms has become more sophisticated, incorporating observed impacts of membrane morphology on heterogeneity of the foulant deposition layers (Ho and Zydney, 2000). Such models have generally been based on laboratory work using ideal surrogate foulants, dedicated to either a fundamental understanding of fouling and/or the development of fouling resistant membranes (Susanto and Ulbricht, 2008), or through mitigation of the adsorption properties of known problem foulants (Maartens et al., 1996; Jones and O'Melia, 2000). From the turn of the millennium there has been increased study of patterns of foulants with variability in cleaning and backwashing. Combinations of operational and material conditions, such as the impact of surface modification of membrane materials (Ma et al., 2000), have been investigated aimed at predicting fouling behaviour.

It is generally the case, across the entire range of UF and MF research, that studies of cleaning have been secondary to developing an understanding of fouling, and have been largely limited to physical cleaning (i.e. backflushing, or reversing the flow back through the membrane, and relaxation, or temporary cessation of permeate flow). Notwithstanding this greatly improved understanding of fouling, including fouling mechanisms and morphology (Lee et al., 2004), there remains a challenge both to develop the techniques and regimes for removing these foulants effectively and efficiently and to obtain the same level of understanding for cleaning as exists for fouling and integrate them into a cleaning science.

2.2 FOULING AND CLEANING LITERATURE SEARCH

A review of the scientific and technical literature (using *Scopus*) based on the search terms “membrane” and “microfiltration” or “ultrafiltration” listed 15522 publications with 3769 publications based on fouling compared with 1302 for cleaning and this reflects that the research agenda has been driven historically by fouling studies. 1020 papers were identified by adding the term “chemical” and, on reviewing these by publication frequency 448 were relevant to cleaning, and the total number of papers studying membrane filtration systems for drinking water production was 125. Of these, 32 had cleaning as the main topic of research (Table 2.1), reflecting the limited attention paid to membrane cleaning by the academic community in the municipal water area in particular. Moreover, many of these studies are ostensibly based on fouling, with the cleaning methods linked to the fouling characteristics. It can be concluded that the science of cleaning, and chemical cleaning in particular, has received basic attention, notwithstanding its pivotal role in the development of cleaning optimisation strategies.

Table 2.1 Literature study: cleaning papers by synopsis.

Topic and synopsis	No.	Papers
Cleaning review-Industrial applications	6	Fane and Fell 1987; Kim et al., 1993;Vaisanen et al., 2002; Maartens et al., 1996; Tragardh, 1989; Tran-Ha and Wiley, 1998
Optimisation and control (cleaning scheduling)	5	Zondervan and Roffel, 2007; Chen et al., 2003; Zondervan et al., 2007; Pontie et al., 1998; Nystrom and Zhu, 1997
Cleanant review - single plant	5	Kimura et al., 2006; Maartens et al., 1999; Lee et al., 2001; Liang et al., 2008; Strugholtz et al., 2005
Cleaning Control Factor Model Development	3	Bartlett et al., 1995; Bird and Bartlett 1995; Huang et al., 2008
Cleaning and Foulant Interactions	3	Heijman et al., 2007; Huang et al., 2009; Nuengjamnong et al., 2006
Cleaning and Membranes -NF	3	Al-Amoudi and Lovitt, 2007; Liikanen et al., 2002; Hong and Elimelech 1997
Cleaning and Membranes -RO	2	Lee and Elimelech, 2007 ; Ang et al., 2006
Cleaning - Ultrasound	2	Masselin et al., 2001; Muthukumaran et al., 2004
Cleaning and Membrane Damage	1	Arkhangelsky et al., 2007
Cleaning and Pre-oxidation	1	Song et al., 2004
Membrane Cleaning	1	Liu et al., 2001

2.3 CLEANING STUDIES

2.3.1 Membrane cleaning.

The simplest definition of cleaning is a procedure applied to a material to relieve it of all non integral substances which are generally termed “foulants” (Tragardh, 1989). Fouling is therefore the general term for the decline in flux, or more accurately permeability, with time. The foulants may be reversible or irreversible, the latter causing permanent changes in the membrane performance (Zeman and Zydney, 1996), and cleaning aims to remove all foulants either by physical, chemical or biochemical means (Cheryan, 2000). This objective is reached.

2.3.2 Cleaning and fouling theory

Membrane cleaning can be envisaged as a 6-step process (Figure 2.1:1) bulk reaction of cleaning materials as the CIP is introduced; 2) cleanant is transported to membrane surface, 3) cleanant transits through foulant layers to membrane surface; 4) cleaning reactions solubilise and detach foulants; 5) waste cleanant with suspended foulants transported to interface and finally 6) transport of wastes to the bulk solution and out of retentate side of membrane fibre (Plett,1985; Shorrock and Bird, 1998). The *de facto* accepted conceptual model for cleaning and fouling (Figure 2.2) has been presented by Liu et al., (2001), who linked the major fouling and cleaning forces. The foulant and membrane interact through hydrophobic attraction; the mechanism for cleaning is electrostatic repulsion. An understanding of the foulant chemical properties expedites the selection of cleaning chemicals, since chemical effects are largely understood: fouling by potable waters is increased at higher foulant molecular weights and charge ratios, the presence of divalent cations in the water and membrane hydrophobicity. An increase in the cleaning medium electrostatic potential through charge density, polarity or pH thus suppresses the attraction forces and increases cleaning efficacy.

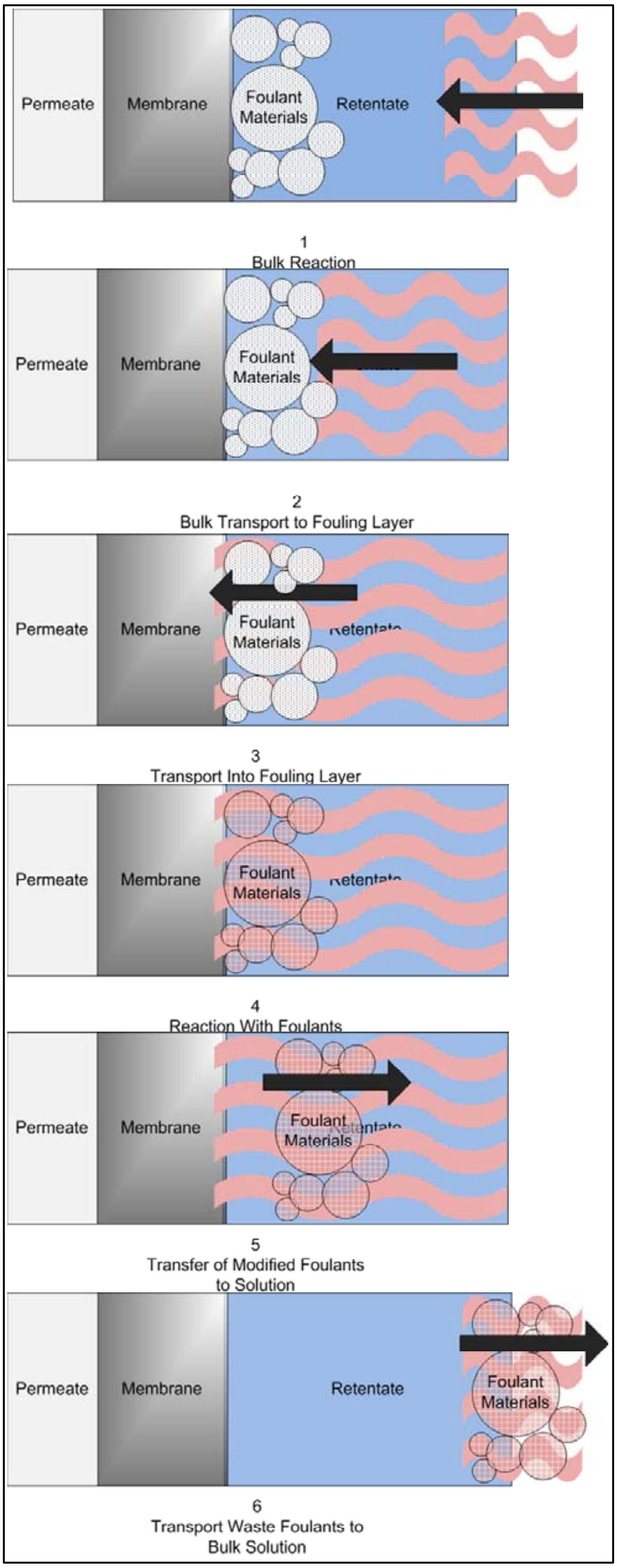

Figure 2.1 **Model for cleaning solution transport, (Plett, 1985)**

2.3.3 Cleaning: Plant Operation and Design.

Cleaning is applied routinely via a chemically enhanced backflush (CEB) or a clean in place (CIP). A chemical reagent is introduced into the retentate side of the membrane; the lumens for in-out capillary tube (CT) or the module casing channel for out-in hollow fibre (HF) membranes (AWWA, 2002).

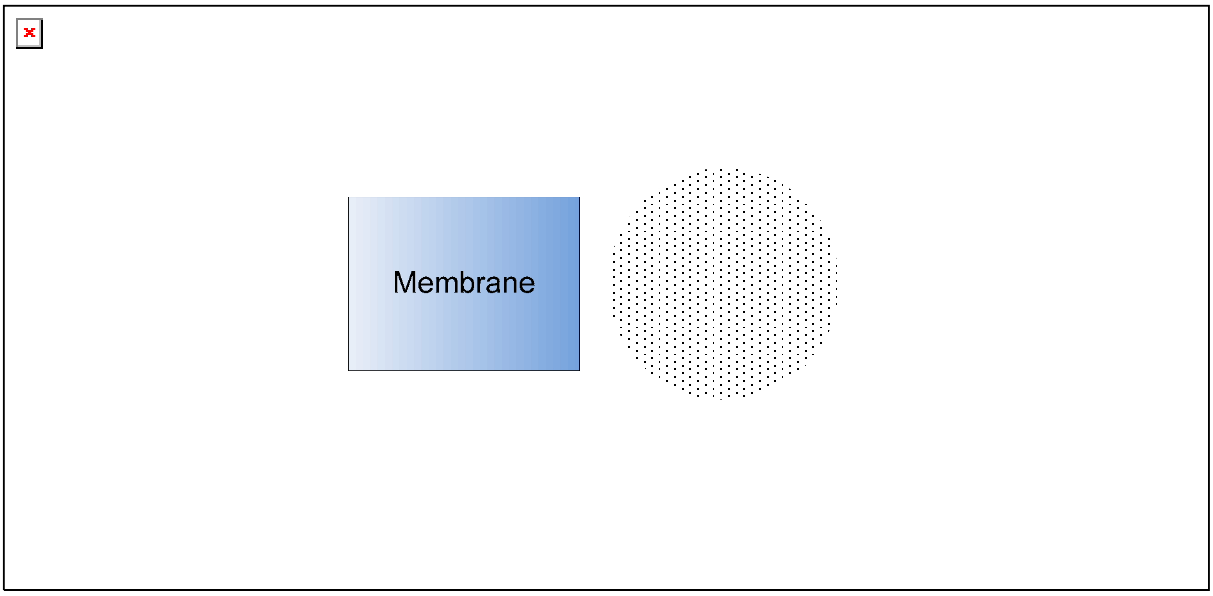

Figure 2.2 Conceptual electrostatic equilibrium model for membrane cleaning (Liu et al. 2001)

For a CEB, the chemicals are introduced with the routine backflush. CEB is used for stabilising/maintaining the permeability (and is often referred to as “maintenance” cleaning) by preventing foulant build up, and is automatically scheduled daily to weekly with no manual intervention. The reagent concentration is relatively low, soak times relatively short and application temperatures normally ambient, as compared to a CIP where the objective is to recovery the flux. For this the cleanants are used for longer soak periods, higher concentrations, and often elevated temperatures. They are normally on an ad hoc basis in response to excessive permeability decline, and can involve a sequence of cleans using different reagents (normally basic and acidic).

Studies of cleaning efficacy across 21 existing full-scale potable water MF/UF plants (Judd, 2004; Rachwal and Judd, 2006) revealed differing sustainable fluxes according to the level of pretreatment, backflush cycle time, chemical cleaning cycle time, cleaning temperature and, in particular, design flux. It is now generally recognised that the capital cost advantage offered by higher-flux operation is more than offset by the greater diligence demanded with cleaning, the reduced cleaning cycle time, and performance and life of the membrane (AWWA, 2005). In a more recent and extensive study, Adham et al., (2005), surveying 87 plants, revealed very diverse cleaning practices. Backwash frequencies varied from 5 to 96 per day with a median of 32 and lasted from 10 seconds to 10 minutes with a median of 77 seconds. Less than half the plants surveyed carried out maintenance cleaning on average once per week (as described above). The CIP cleaning frequency ranged from 0.2 to 50 per year with a median of 4 per year. Results were skewed by the under sizing of some plants, confirming the findings of Judd (2004) regarding flux sustainability and possibly confirming the critical flux concept (Field et al., 1995).

Table 2.2 Generic chemical cleanants with usual reactions

Cleanant	*Chemical*	*Reactions*
Base	Caustic Soda, (NaOH)	Hydrolysis and solubilisation, saponification.
Oxidant	Hypochlorite, (HOCl), Hydrogen Peroxide (H_2O_2)	Oxidation and disinfection
Acid	Hydrochloric (HCl), Sulphuric Acid (H_2SO_4), Nitric Acid (HNO_3)	Solubilisation
Acid Chelate	Citric	Chelation
Alkaline Chelate	EDTA	Chelation
Surfactants	Proprietary	Emulsifying, dispersion and surface conditioning

2.3.4 Cleaning reagents

As mentioned above, in the electrostatic equilibrium model forces retaining the foulant at the membrane are reduced during cleaning allowing their physical removal (Liu, et al., 2001). It follows that the most appropriate chemical will depend upon the nature of the foulant (organic/inorganic, acidic/basic, charged/uncharged), and the most commonly used are summarised in Table 2.2.

2.3.4.1 Caustics

Caustic soda (sodium hydroxide, NaOH) solutions are applied at high pH levels (pH 11-12, or possibly less if membrane integrity is an issue). Hydroxide encourages dissolution of weakly acidic organic matter, generally with carboxylic and phenolic functional groups, and promotes break down and cleavage of polysaccharides and proteins into smaller sugars and amides (Faust and Aly, 1998; Lee, et al. 2001; Al-Amoudi & Lovitt, 2007). Hydroxide also expands NOM molecules, allowing mass transfer of the cleanant solution to the membrane surface (Hong and Elimelech, 1997). Alkali agents are also effective for removing inorganic colloids and silicates where charge effects and ionic strength increase the solubility and electrostatic repulsion (Kim et al., 1993).

Increased concentrations of NaOH have been seen to increase permeability recoverability up to a threshold which varies for different foulants and membrane materials (Bartlett, et al. 1995; Kim et al. 1993; Liu, et al. 2001). NaOH is poorly buffered so for heavily fouled membranes it is necessary to maximise the concentration (Zeman & Zydney, 1996). Studies looking at optimal recovery using NaOH for MF and UF membranes fouled during industrial and dairy applications cleaned with NaOH have revealed a threshold concentration however these do not relate it further to material or process factors (Vaisanen et al., 2002). This threshold limiting the ability to apply enough NaOH affects the “adsorption-desorption” interaction of the cleanant with the membrane and may explain why the threshold varies with membrane, foulant chemistry, surface charge and fouling type (Al-Amoudi and Lovitt, 2007). The primary barrier to effective MF/UF chemical cleaning is thought to be mass transfer, where the chemical is prevented from reaching the foulant at high enough concentrations to overcome the attraction forces (Liu et al., 2001).

2.3.4.2 Oxidants

Oxidants include hydrogen peroxide (H_2O_2) and sodium hypochlorite, (NaOCl), of which the latter is by far the most common, (although some polypropylene (PP) membranes are not chlorine tolerant (Cheryan, 2000)). Oxidation degrades the NOM functional groups to more soluble carboxyl, ketonic and aldehyde groups (Thurman, 1985), which makes them more susceptible to hydrolysis at high pH levels. The combination of oxidant with alkaline cleanants has been shown empirically to be more effective than oxidant cleanants alone, especially where organic foulants dominate. (Strugholtz et al., 2005; Lin, et al. 2001; Hong and Elimelech, 1997; Zondervan and Roffel, 2007).

In a comparison study between alkaline H_2O_2 and NaOCl cleaning of NOM-fouled PES UF membranes the NaOCl gave better cleaning regarding the concentration of organics removed (Strugholtz et al., 2005). NaOCl was found to have oxidised the aromatic humic substances when the pH was raised, but this was not repeated for peroxide. The halogenation of the aromatic rings is known to be a common mechanism for the breakdown of humic materials (Carey and Sundberg, 2000), and the Strugholtz et al. study raises questions as to whether the oxidative properties of the hypochlorite are more significant than the hydrolysis catalysing effect of the hydroxide. Liu et al. (2001) suggested that caustic promotes a more open fouling layer to allow the chlorine to reach the membrane surface, which may explain the increased efficacy of hypochlorite at higher pH levels. Huang et al. (2008) indicated that free chlorine affects oxidation of organic colloids while the soak time governs their diffusion from the membrane surface. The variation in results from these studies shows that reducing cleaning studies to the standard reaction properties of the chemicals, the foulant type and the membrane material is not a reliable approach for prediction of cleaning efficacy.

Chlorine is thought to contribute to membrane swelling. Whilst this enhances mass transfer of the cleanant to the foulant at the membrane surface (Zeman and Zydney, 1996), at concentrations above 150 mg.L^{-1} membrane damage may result; particularly at low and neutral pH levels, through chlorination of the C-S bonds in PES (Arkhangelsky et al., 2007). Such chemical modification can alter the membrane charge, increase pore size and ultimately promote flux decline causing permanent loss of recoverability changing its rejection characteristics.

2.3.4.3 Acids

Acid cleaning aims to remove, or break up by solubilisation, those foulants composed of cations, metal salts and metal oxides from membrane surfaces (Zeman and Zydney, 1996; Tragardh, 1989). Mineral acids such as hydrochloric (HCl) and sulphuric (H_2SO_4) are commonly used and are effective for both CEB and CIP down to pH 1.0 for PES and PVDF, below which integrity problems may arise. As well as catalysing acid hydrolysis, acids are mildly oxidative on the NOM; forming soluble aromatic aldehydes and acids at NOM functional groups (Thurman, 1985). However, they are generally used for removal of mineral scaling. Whilst strong mineral acids can solubilise deposits, organic acids, such as citric and oxalic, are more effective for formation and transportation of organo-metallic foulants from the membrane to the bulk solution (Strugholtz et al., 2005).

Citric acid has good buffering and chelating abilities making it effective and easy to use with a lower risk of pH damage from overdose than mineral acids (AWWA, 2005). Along with NaOCl it is reported to be the most used chemical cleaner for MF/UF membranes (Adham et al. 2005). It can also disrupt biofilm formation by removing minerals from foulant layers disrupting enzymes that allow their establishment (Mallevialle and Odendaal 1996; Liikanen et al., 2002). Iron readily forms complexes (Cheryan, 2000) but their limited solubility inhibits transport from the membrane surface (Zeman and Zydney 1996). The combination of citric and oxalic acid has proved effective in practice where multiple organometallic complex species can form (Presdee et al., 2006), supporting observations from cleaning effectiveness studies performed by Liikanen (2002) where combinations of cleanants in multi stage applications proved more effective and high flux recoveries were possible despite incomplete foulant removal (Liikanen et al., 2002). These studies show that subtler cleaning methods can be effective and this bodes well for contemporary efforts to reduce chemical usage.

2.3.4.4 Others

Phosphoric acid (H_3PO_4) is another chelating ligand providing reasonable removal of inorganic scale but, in common with most chelates, poor removal of elemental sulphur and colloidal silicates (Mallevialle and Odendaal, 1996). Its buffering capability and sequestering ability make it potentially useful for routine automated cleaning or even CEB where the risk of overdose and damage are lessened. Its use has been reported in few MF/UF cleaning papers and is generally recommended for only for some proprietary industrial membranes (Tragardh, 1989) and, most usually, for some cellulose acetate RO membranes fouled with metal oxides (Al-Amoudi and Lovitt, 2007). It has been shown to be less effective than other cleaning chemicals against NOM (Zondervan and Roffel, 2007).

Cleaning additives such as EDTA (ethylenediamine tetraacetic acid) provide additional chelation capacity for metals such as calcium and disperse of minerals in general (Zeman and Zydney, 1996; Liu et al., 2001; Zondervan and Roffel, 2007). It is often blended into proprietary cleanants (P3 Ultrasil 11) where studies have demonstrated it can reduce biofilm reformation acting as enzyme disruptor (Shorrock & Bird, 1998). Its use is favoured in the USA, where it was used on apparently irreversibly fouled PP membranes. (Recovery was based on sequential soaking (3hrs) at elevated temperatures (40-45°C) with combined EDTA, NaOH and H_2O_2) Judd (2004).

Ammonium bifluoride (NH_4HF_2) has also been used to assist the removal of silica deposits (AWWA, 2005), though silica remains one of the most challenging of foulants as it forms inert silicate colloids with high surface attraction forces (Brant and Childress, 2004). Silica is a problem foulant as it can accumulate in the pores of MF/UF membranes Judd (2004). It is difficult to remove (AWWA, 2005), as in it's colloidal form it resists solubilisation and is resistant to hydrolysis. It is destabilised however by the more reactive fluoride, (for example calcium silicates are replaced by calcium fluorides). Ammonium bi fluoride is typically dosed to break up silica deposits (AWWA, 2005). An alternative is to consider pre-treatment (Faust and Aly 1998).

2.3.5 Cleaning sequence

Cleaning sequence can affect the degree of permeability recovery (Zeman and Zydney, 1996). Studies where membrane cleaning sequences have been tested consistently indicate that an alkali followed by an acid clean is more effective than the reverse for membranes treating surface water (Tragardh, 1989; Kim et al., 1993; Bartlett et al., 1995; Hong and Elimelech 1997; Huang et al., 2007; Liu et al., 2001; Vaisanen et al., 2002; Al-Amoudi and Lovitt, 2007; Liang et al., 2008). This could be explained by effect of residual charge on the membrane surface and foulants following the more negatively charged alkaline clean. The cleaving and expanding of the foulant molecules with caustic solutions would also be expected to enhance cleaning reagent mass transfer for the subsequent step. Despite this, the acid-alkali cleaning sequence is commonly specified for groundwater sources, the acid removing inorganics such as metal oxides and carbonates which may otherwise precipitates at high pH (AWWA, 2005). Strugholtz (2005) noted that there was reduced temperature sensitivity for removal of hydrophobic NOM with the acid-alkali sequence (Strugholtz et al., 2005). Industrial process waters treating proteinaceous waters reoutinely employ the acid-alkali sequence to prevent gel layer formation (Koltuniewicz and Drioli, 2008) or mineral precipitation (Tragardh, 1989).

The choice of cleaning sequence will be governed by the feed water and the type of fouling and it can be seen from the literature that it is by no means clear which is the preferable. The predominance of hydrophobic NOM or inorganic scaling will determine which is applicable. The optimum may change seasonally or with the ageing of the membranes.

2.3.6 Other cleaning factors

Increased temperatures decrease the reliance on turbulence to overcome mass transfer limitations, since diffusive mass transfer is increased, and this is seen as an important qualitative cleaning model factor (Bartlett et al., 1995). Foulant solubility is also increased at higher temperatures, as is reactivity of the NOM functional groups (Carey and Sundberg, 2000), along with swelling and mechanical destabilisation of organic foulant or biofilm layers (Lee and Elimelech, 2007). However, extremes of temperature and pH can also solidify and crystallise biofilm deposits, making them harder to remove (Bartlett et al., 1995). Also, membrane materials can be sensitive to pH and temperature. This has promoted interest in benign alternative cleaners such as enzymes, although these are expensive and better suited for industrial and food waste treatment (Maartens et al., 1996).

Pre treatment with H_2O_2 / UV, an advanced oxidation process (AOP), has been demonstrated to mitigate fouling and improving cleaning efficiency (Song et al., 2004) of nanofiltration membranes, improving permeability recovery with caustic cleaning. It was suggested by Song et al. that pre-oxidation reduced the absorbance of NOM onto the membrane material. This may be less important for MF/UF membranes, where surface charge is more important than concentration polarisation effects, but since the treatment impacts on the hydrophobic NOM it would be expected to influence cake compression affecting resistance and cleanability characteristics.

Ultrasonic cleaning has been used for treating flat sheet membranes in laboratory tests (Nystrom and Zhu, 1997), though this treatment causes membrane damage and incurs a high energy demand (Masselin, 2001). There is evidence that some membranes are more susceptible to integrity failure than others: PES materials were shown to fail after 5 minutes of exposure to the ultrasound (at $f = 47$ kHz). However, more recent studies have demonstrated the efficacy of treatment in crossflow systems for membranes fouled with dairy wastewater (Muthukumaran et al., 2004) and there is still research needed to establish whether ultrasonic has a place in membrane treatment where integrity is not so critical.

2.3.7 Cleaning factors studies

Given the stated variables influencing cleaning efficacy it is clear that rigorous assessment of cleaning efficacy is challenging, especially given the heterogeneous and sometimes quixotic nature of fouling. Comprehensive modeling of membrane cleaning demands a large number of experiments to cover possible ranges of membrane fouling conditions from natural waters.

Bartlett et al. (1995) sought to model the cleaning factors qualitatively for flat sheet membranes fouled with dairy waste. The study examined concentration, temperature, alkali/acid combinations and cross flow effects on different surface concentrations of the same foulant. It was concluded that optimum cleaning could be achieved using a number of different regimes. Chen et al (2003) conducted the first factorial study of membrane operating parameters using analysis of variance (ANOVA) to analyse the results. The experiments employed HCl and NaOH to clean spiral wound RO and UF fouled with municipal wastewater secondary effluent. Factors considered were: NaOH concentration, temperature and recirculation duration; HCl cleaning time and the use of forward backflush and normal backwash after chemical cleaning. The relationships between the resulting recoveries were explored. For this application it was seen that the temperature and concentration of the NaOH and the applications of a backflush yielded optimum permeability recovery.

The factorising of chemical cleaning variables has been taken further by some reseachers (Lee and Elimelech, 2007, Mahdi et al., 2008). There exists possibilities for automating and optimising the cleaning and backwash cycles on UF plants and this was investigated by Smith (2006) .The need for controlling cleaning has led to the proposal of dynamic models for predicting optimum soak times as presented by Zondervan et al., 2007. These models have been developed for given specific conditions, and mainstream cleaning research remains generally restricted to specific treatment problems and empirical and heuristic data (Liang et al., 2008), and cannot be conflated with the progress suggested towards the generic fouling models (Huang et al., 2009).

2.4 CONCLUSION

Foulant studies, particularly NOM fouling, have informed research into cleaning but are “a single frame of a moving picture” (Thurman, 1985). These studies have evolved from mass transport and attachment/detachment forces equilibrium models, where the problem of cleaning is reduced to factors based on the condition of the foulant. The variety of tools available for characterising foulants and the empirical/heuristic approach employed, has led to different interpretations of the relationship between fouling and MF/UF plant design and operation and it has been proposed that unified fouling index be developed (Huang et al., 2009).

Backwashing and its effectiveness in controlling fouling remains entirely empirical; with laboratory, bench and pilot studies to investigate the optimum frequency, rate and duration for given applications (Kim and DiGiano, 2006; Katsoufidou et al., 2008). For example, as the fouling distributions in MF-UF hollow fibre systems is heterogeneous, it is possible to partially backwash the fibre length with systems preferentially backwashing nearest the inlet pots (Van de Ven et al., 2008a).

There are issues with the use of autopsy tools for foulant analysis to develop a better understanding of cleaning. AFM (Atomic Force Microscopy) has been proposed as tool to measure the foulant removal (Song et al., 2004) and the effectiveness of the cleaning techniques (Ang et al.,2006) based on the analysis of the concavities of the surfaces. Zeman and Zydney (1996) see problems with the representativeness of scanning electron microscopy. The question remains whether the foulant is on the pores or surface and laboratory technology cannot address mass transfer modelling with pore structure variation or the reversibility of charge effects from NOM adsorption.

Permeability recovery alone is not enough on its own to characterise changes to cleaning (Evans et al., 2009) and development of dynamic cleaning models takes us more closely to a systems approach to cleaning where input variations, surface modification and ageing effects could be compensated for. A broader understanding of how the cleaning efficacy changes under different loading and fouling conditions is required. Further use of statistical methods such as response surface models (Bas and Boyaci, 2007) could be used to generate polynomials of the cleaning design space, if applied across a variety of membrane plants and types systematically it may help improve understanding of the changes to the cleaning efficacy.

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CHAPTER 3 MICROFILTRATION MEMBRANE PLANT START UP: A CASE STUDY WITH AUTOPSY AND PERMEABILITY RECOVERY ANALYSIS

Porcelli N., Hillis, P., and Judd S. (2009) Microfiltration Membrane Plant Start Up: A Case Study With Autopsy And Permeability Recovery Analysis. *Environmental Technology, 30 (6), 629-639*

Abstract

A potable water treatment plant, supplied from a low NOM (natural organic matter), low turbidity source with pre-coagulation and two-stage pressure sand filtration, had a MF (microfiltration) membrane process added to meet UK Water Regulations. Modules autopsied showed that despite upstream coagulation/filtration with chlorination, a bio-film of EPS (extracellular polymeric substances) and inorganic particulates had developed. Backwashing under laboratory conditions yielded near full recovery. Laboratory-scale modules of fouled fibres from pre-commissioning and post-commissioning were assembled and cleaned. The recovery was modelled and optimised with a response surface experiment using variables of concentration, soak period and temperature. The pre-commissioning fibres were recovered more by longer chemical soak times at higher cleanant concentrations the post commissioning fibres. Comparative tests on post commissioning fibres indicated that full recovery was possible with organic acids. It was concluded start up of new membrane plants may involve fouling conditions not necessarily representative of those under routine operating conditions, such that modification of prescribed cleaning operations may be required.

3.1 INTRODUCTION

A microfiltration (MF) membrane plant was installed in response to the UK *cryptosporidium* regulations. It is in a remote rural upland location in the English Lakes National Park with soft waters; dosing sodium carbonate for pH correction and pre-coagulation and poly aluminium chloride periodically in autumn when high aluminium levels are found (Figure 3.1). There are two stages of pressure filtration and the plant itself comprises five trains of 120 x 23.4 m^2 PVDF (polyvinylidene difluoride) modules, running at 5.1 MLD (75 $L.m^{-2}.h^{-1}$) with five units online in constant flux mode. Raw water is typically low in alkalinity (> 40 $mg.L^{-1}$), has minimal colour (< 30 Hazen) and turbidity (< 2 NTU) and with low organics levels (dissolved organic carbon, DOC < 5 $mg.L^{-1}$) and no reported algal bloom problems. On commissioning the remote location of the plant initially prohibited 24 hrs operation. Operation was instead with varied throughput and cleaning intervals and encountered a higher than expected rate of flux decline. This paper presents the results of membrane autopsy and cleaning optimisation studies carried out on the plant to restore membrane permeability and attain the design throughput.

DOSING
A Sodium carbonate pH Correction to 7.8
B Sodium hypochlorite for Mn and Fe removal, 0.75 mg/l residual
C Poly aluminium chloride for Al removal
D Sodium hypochlorite disinfection, 0.5 – 1.5 mg/l

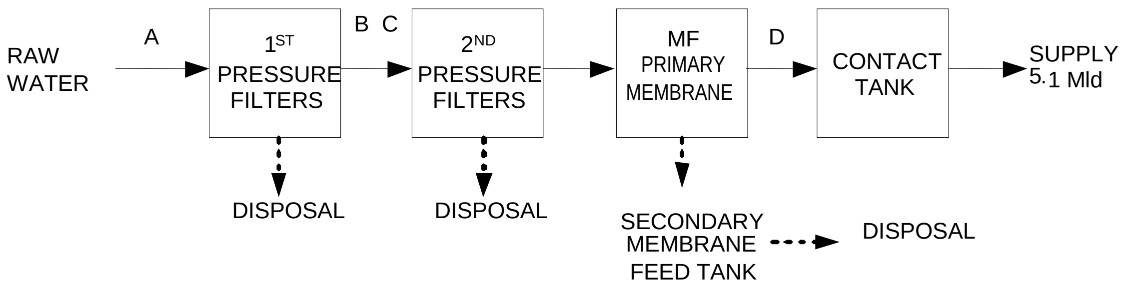

Figure 3.1 **Process Flow Diagram for Plant**

3.2 MATERIALS AND METHODS

Modules were removed when fouled during site visits at start up (< 20 hrs operation - A) and at post commissioning (98 hrs of operation -B) when raw and treated water through the process sampled for characterisation. Rinsed foulant elutes were analysed and permeability recovery experiments tests carried out to optimise citric acid CIP (clean in place) cleaning using variations of concentration (C'), soak period (P') and temperature (T'). A response surface analysis with a two factorial response model was used to test parameter sensitivity to variation. The relative permeability recovery comparing organic and inorganic acid was also determined.

Eluates were analysed using High Performance Size Exclusion Chromatography (HPSEC) using a high performance liquid chromatography device (HPLC) (Shimadzu VP Series, Shimadzu, Milton Keynes, UK) with a UV detector set to 254 ηm. The mobile phase was 0.01 molar (0.01 M) sodium acetate at a flow rate of 1 mL.min^{-1}. The column used was a BIOSEP-SEC-S3000 7.8 mm (ID) × 30 cm downstream of a guard column with a GFC-3000 disc 4.0 mm (ID) × 3.0 mm

(Phenomenex UK, Macclesfield, UK) fitted. The samples were introduced in the machine in Shimadzu LC Silicone/PTFE (polytetrafluoroethylene) wider screw cap 2 mL vials (Fisher Scientific UK Ltd, Loughborough, UK). For each sample a chromatogram of UV254 (absorbance units) against time (minutes) was detected. The HPSEC calibration used was developed at Cranfield University using a raw water sample taken in April 2002 from the Albert WTW (Goslan et al., 2002).

The total and dissolved organic carbon was measured using a Shimadzu TOC 5000-A Analyser (Shimadzu, Milton Keynes, UK). As some of waters and elutes contained increased levels of $CaCO_3$ (inorganic carbon, IC), which can damage the catalyst, the DOC level of the samples was measured using a non-purgeable organic carbon method (NPOC). The samples were pre-treated by adding 2 M HCl solution (10 $\mu L.mL^{-1}$ sample). Before the DOC content of the samples was measured, zero-graded air (80% N_2, 20% O^2, and 0% CO_2) was injected into the sample and the carbonate (IC fraction) purged out. A drawback of the NPOC method is that any volatile organic compounds will also be purged away with the IC. The machine was calibrated using a total carbon (TC) standard, which was prepared by dissolving potassium hydrogen phthalate (2.125 g) in reverse osmosis (RO) permeate water (1 L). The standards produced had a concentration of 1000 mg. L^{-1} DOC and working standards were diluted with RO water to concentrations of 0, 2.5, 5, 10, 25 and 50 $mg.L^{-1}$ DOC.

3.3 AUTOPSY: FOULANT CHARACTERISATION

Fibre tows of surface area of 2.5 x $10^{-3}.m^2$ from module (A) and (B) were extracted and chopped into 10 mm lengths before blending in a 330 Watt bench blender for ten minutes in 500 mL of 18.2 MΩ ultrapure RO water. The initial water temperature was measured and rose from 25°C to 46°C during the blend. A 250 mL aliquot of the solution was then filtered through rinsed 0.45 µm cellulose acetate paper and the eluates kept. A blank of ultrapure water was subjected to the same regime and ran in parallel. Eluates were characterised for particle size profile with HPSEC and total and dissolved organic carbon with Total Organic Carbon (TOC Inorganic Carbon

(IC) Total Carbon (TC) method). Biofilm rinsed off the membranes was examined under an optical microscope.

3.4 PERMEABILITY RECOVERY

Fibres from the fouled modules A and B were prepared for pure water dead end permeability tests. These tests were carried out before and after cleaning with chemicals to assess the recovery for a controlled set of cleaning conditions. Fouled fibres in the modules were rinsed with 500 mL of DI water from a hand spray, so that bio-film and surface layers were removed; simulating efficiently backwashed fibres. The rinsed fibres were stored in darkness at 3°C wetted with source water whilst awaiting assembly into bench test modules. Fifteen modules were prepared for each test.

3.4.1 Bench test module fabrication

Single rinsed extracted fibres were each threaded into end-caps and potted with silicon sealant. Small hand instruments were used so as to avoid crushing. The free end was sealed in rubber tube with epoxy resin. The end-cap and module was assembled forming a bench membrane module assembly to replicate an "out-in" process. Each assembly had a "tow" of ten 400 mm membrane wetted hollow fibre lengths. A 45 mm of pot length was included as shown in Figure 3.2a. Fibres were drawn through the end-cap (Figure 3.2b and 3.2c) and potted with the heaviest fouled at the end-cap end. Each module had a calculated a filtration area of 0.010053 ± 0.000025 m^2.

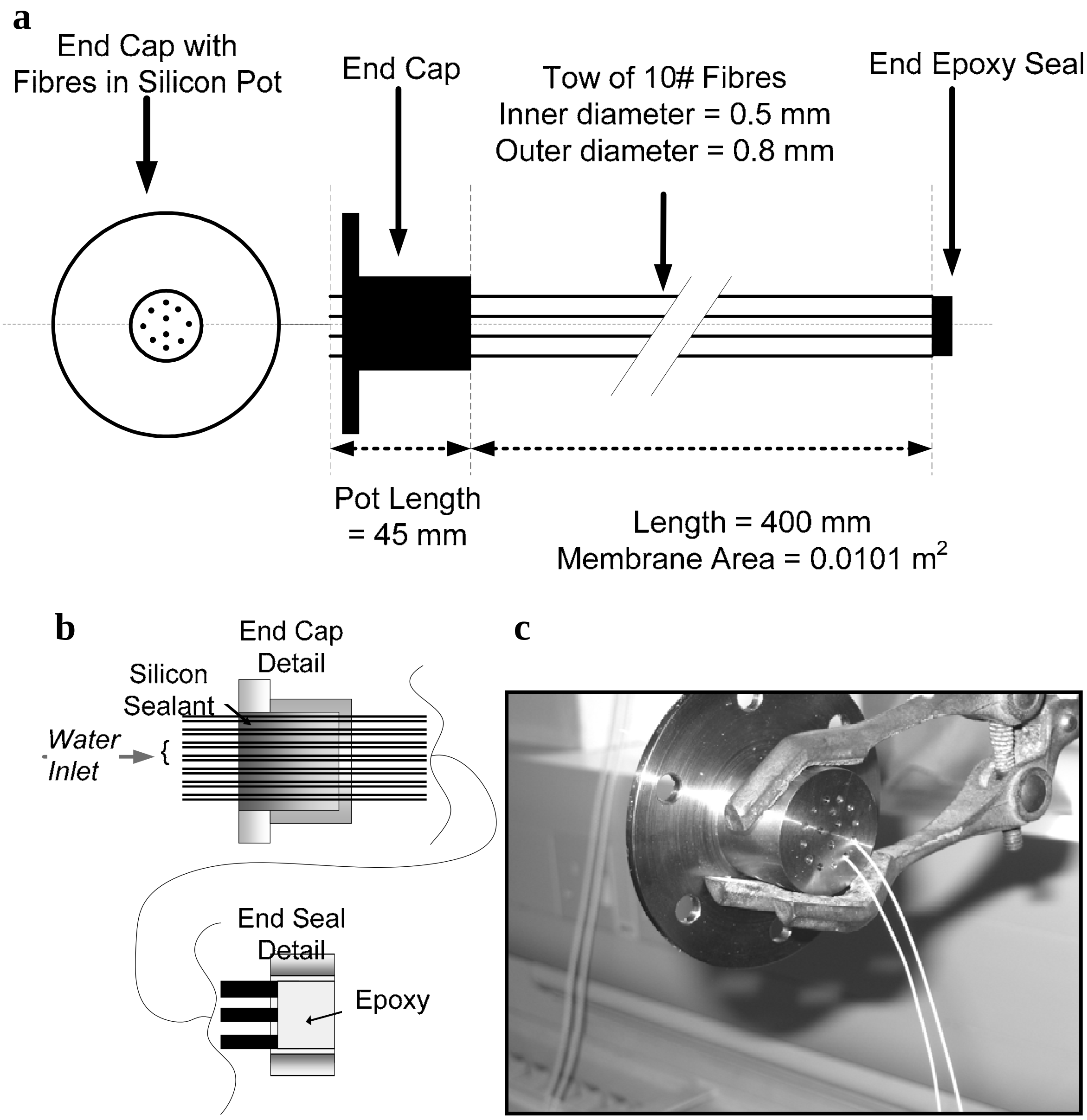

Figure 3.2 **Test module assembly: a) Schematic of bench module, b) end-cap and sealing and c) fibres threading**

3.4.2 Integrity test

Manual handling poses a risk of fibre damage and this will be evident from large permeability deviations when each assembly is tested. To detect this, each assembly was checked for integrity using a Pressure Decay Test (PDT). A rig was designed and constructed (Figure 3.3) and yielded a calculated log reduction value for 0.1 μm particles using a standard water industry procedure (ASTM, 2003). Pressure was measured with a 0 to 2.000 +/- 0.001 bar g pressure sensor/transmitter in a 4-20 mA current loop scaled at 0 – 1.000 bar g. The sensor/transmitter relayed pressure data every 0.5 seconds via an RS232 connection to a computer data-logger. The test used dried oil-free instrument air through a pneumatic filter regulator. A volume of air at a manually set pressure was trapped on the upstream side of membrane bench module and the pressure decline monitored over ten minutes. The air leakage rate was calculated and related to the Log Reduction Value (LRV) using the standard method (AWWA, 1992). This allowed us to check the assembled test modules had similar characteristics and were not damaged in construction before cleaning tests commenced.

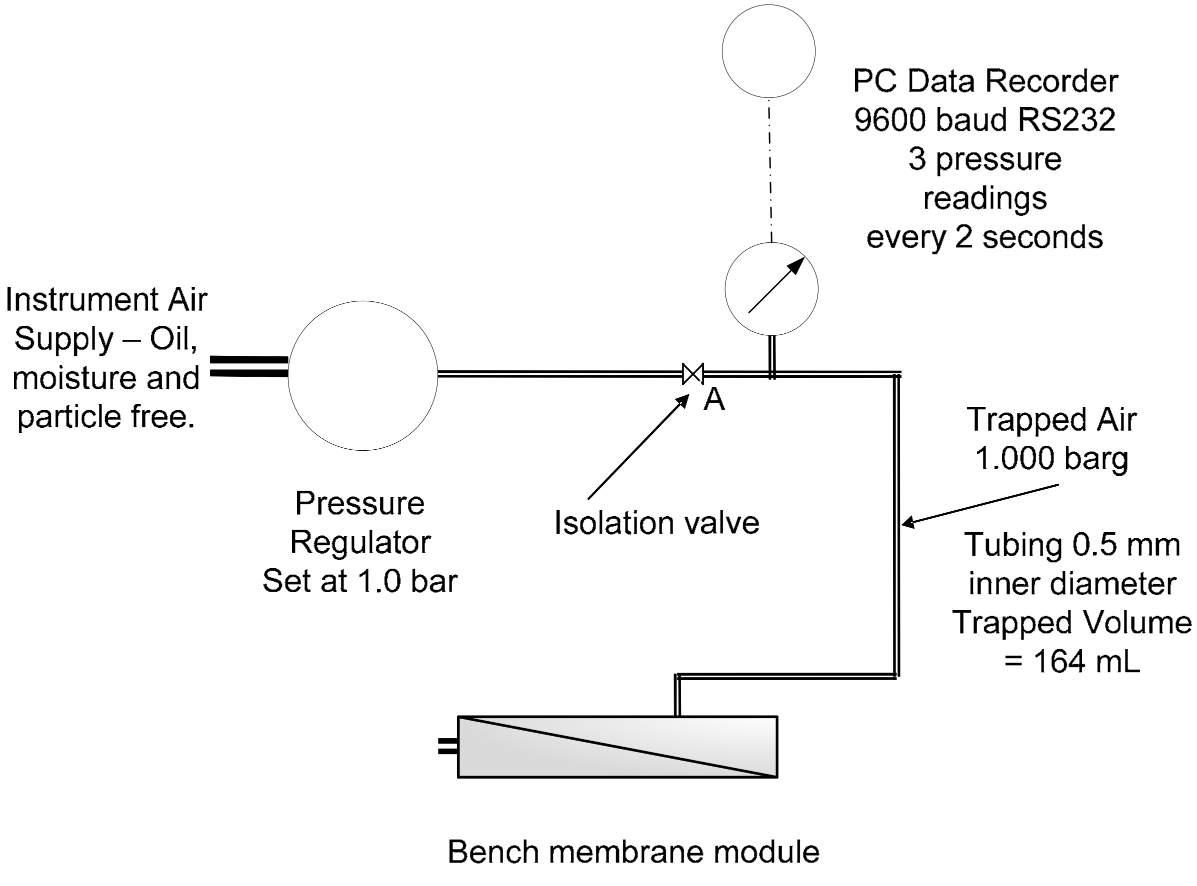

Figure 3.3 **PDT equipment for membrane integrity tests.**

3.4.3 Permeability test

Clean water permeability tests were performed on the modules before and after each cleaning operation. A bespoke rig (Figure 3.4) gave a constant head of distilled feed water to an attached bench module. The feed water is kept circulating through a 0.45 μm membrane to maintain purity and also to maintain a constant meniscus in the inner reservoir, which is fixed at a 2000 mm over the centre line of the bench membrane module in the rig, so that the driving pressure applied across each membrane at was constant at 0.196 bar (ΔP m = trans membrane pressure TMP).

The temperature of the laboratory and the inner reservoir water were measured at the start of each test. Flow was out to in, and permeate was collected on an electronic balance having +/- 0.1 g increments. Readings were relayed at 5 per second via an RS232 to a data-logger over the ten minute test period. The permeability was calculated by converting the mass rate found through this module area into clean water flux (J_w), normalised for 20°C:

$$K = \frac{J_w}{\Delta P_m}$$

Equation 1

Equation 1 is derived from Darcy's equation and as the test medium is the same ultrapure water at the same temperature for each test, for engineering purposes we can ignore the viscosity (*u*) for practical conversion of TMP to p. g. h, where h is the water head in metres and ρ the relative density is assumed to be unity.

Permeability recovery ($L.m^{-2}.h^{-1}.bar^{-1}$) was calculated using a relative measure to the virgin permeability:

$$\%R = \frac{K_f - K_i}{Kv - Ki} \cdot 100$$

Equation 2

where: K_f is final permeability measured after chemical cleaning ($L.m^{-2}.h^{-1}.bar^{-1}$), K_i the initial measured permeability, and K_v the virgin pure water membrane permeability from clean tests determined as 700 $L.m^{-2}.h^{-1}.bar^{-1}$.

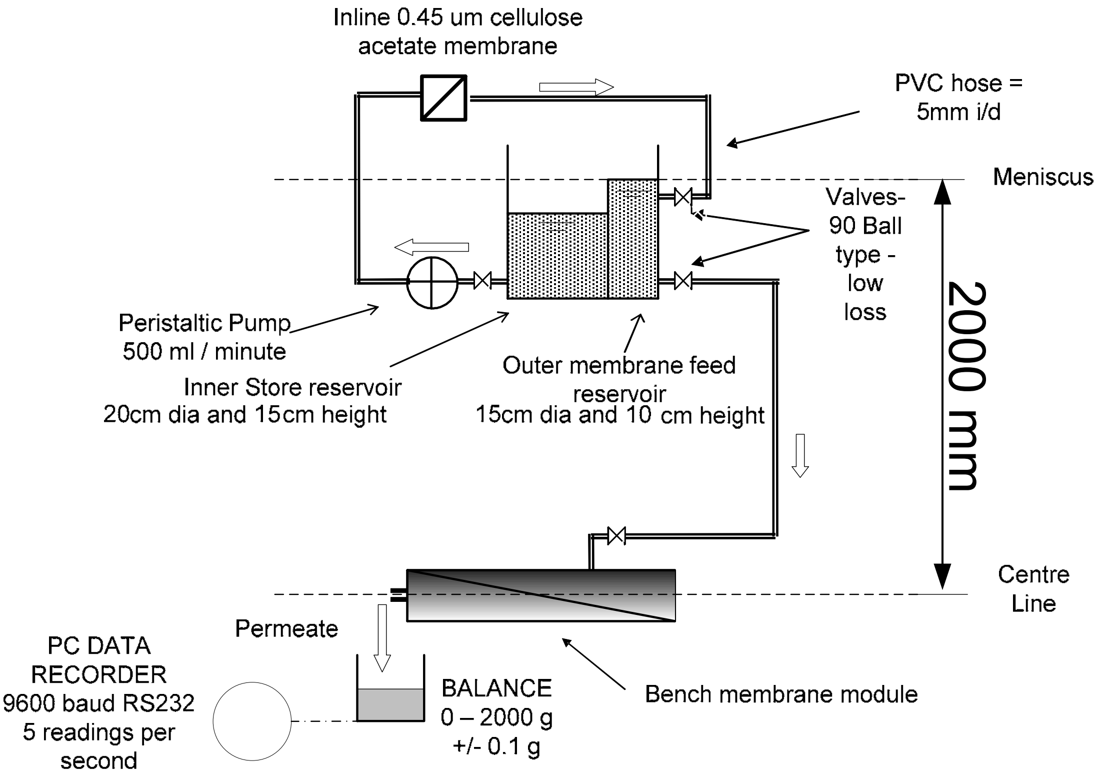

Figure 3.4 **Schematic of pure water permeability test rig.**

3.4.4 Cleaning recovery tests.

Cleaning procedures for both citric acid and hydrochloric acid at a phosphate buffered pH of 2.2 were carried out. Fibres were placed in flasks with the cleaning chemicals drawn through from the fouled side to the permeate side using a vacuum pump until the membrane was saturated with the cleanant. Flasks were immersed in a stirred temperature-controlled water bath for a set test period in each experiment.

Experiments to optimise citric acid performance were carried out. Fifteen tests at varying conditions were performed using a response surface Box Behnken experimental design (Myers et al., 1989). Cleaning parameters are varied in a cube matrix in equal proportions; with the central points of each parameter repeated three times. Some of the benefits of this method over other factorial are given (Box and Draper, 1986):

1. Good lack of fit detection.

2. Internal estimate of error.
3. Constant variance check.
4. Minimum number of treatment combinations.
5. Good graphical analysis through simple data patterns.
6. Good behavior when errors in settings of input variables occur.

The design uses coded parameters:

a = Concentration of citric acid (C'), 0.6 to 1.8 g. L^{-1},
b = Period soaked, (P'), 30 to 90 minutes, and
c = Temperature of cleanant (T'), 5 to 35°C.

Fifteen bench modules were assembled from fibres at the start and completion of commissioning and the relative recoveries from each variation were analysed with Design Expert v7.0 statistical modelling software (Stat-Ease, Inc., Minneapolis, USA) was used to analyse the responses. A two factorial model was used to generate algorithms and least square optimums were established and a variability analysis carried out.

Following the response surface experiments a protocol experiment was designed to compare the mineral (HCl) versus organic (citric acid) acid reagents at the optimised regimes. Eight more modules were assembled from fresh fibres taken from the post commissioning module (B) and a further test (Test C) carried out on post-commissioned fibres to compare inorganic hydrochloric acid alone at pH 2.2 and citric acid (1.2 g.L^{-1}) solution at pH 2.2, 20°C for 60 minutes.

3.4.5 Supplementary tests

Scanning electron microscope (SEM) imaging was carried out using a high-resolution cross-sectional SEM a Phillips XL30 SFEG (Philips Analytical, Munich, Germany) with energy dispersive x-ray analysis detection system (EDX). A focused electron beam was scanned at 30 kV over gold and palladium spluttered specimens such that the interaction between the beam and the specimen excites different radiation forms including secondary electrons and x-rays. The radiation is detected and analysed to reveal information regarding the composition and topography of the specimen at resolutions allowing images of between 0.1 µm to 400 µm. Topographic imaging allowed a quantitative evaluation of the foulant as well as an evaluation of membrane surface and pore definition. In combination with SEM imaging, elemental composition by sample % weight was determined using EDX analysis and processed using INCA^{tm} software, (Oxford Instruments, Abingdon, UK).

Surface charge measurement of the plant's raw water particles was conducted using a laser diffraction instrument (Malvern Mastersizer 2000, Malvern Instruments, and Malvern, UK). A measured volume of sampled raw water from the membrane plant was dosed incrementally with a dilute solution of 20% Polydadmac (Polydiallyldimethylammonium chloride) (6.2 meq.g^{-1}), whilst being maintained at a pH of 7.0 by addition of 0.01N NaOH, until the point of zero charge (PZC) was indicated. The charge per mass of DOC was calculated from the amount added to bring the raw water to the PZC indicated by the instrument.

3.5 RESULTS AND DISCUSSION

3.5.1 Autopsy and Site Investigations

Initial examination of the pre-commissioning fouled fibres showed them to be coated with a gelatinous clear film that was easily rinsed off with de ionised water (DI). Observation under an optical microscope indicated a film comprising a matrix of inorganic minerals and bacteria. This film was more concentrated in the core and the module had flecks of black, initially suspected of being manganese (Figure 3.5a). Foulants rinsed using DI water were examined under optical and electron microscope; a classic biofouling matrix of extracellular polysaccharides (EPS) and mineral detritus >50 µm (Figure 3.5b) with bacteria was observed. These foulants were rinsed from the surface of the fibre and deposited onto 0.45 µm cellulose acetate membrane papers and subjected to SEM/EDX analysis. Elemental analysis of the eluate showed it to be almost identical to the EDX analysis of the fouled membrane (subtracting membrane material analysis), suggesting that the fouling layer was readily removed by rinsing.

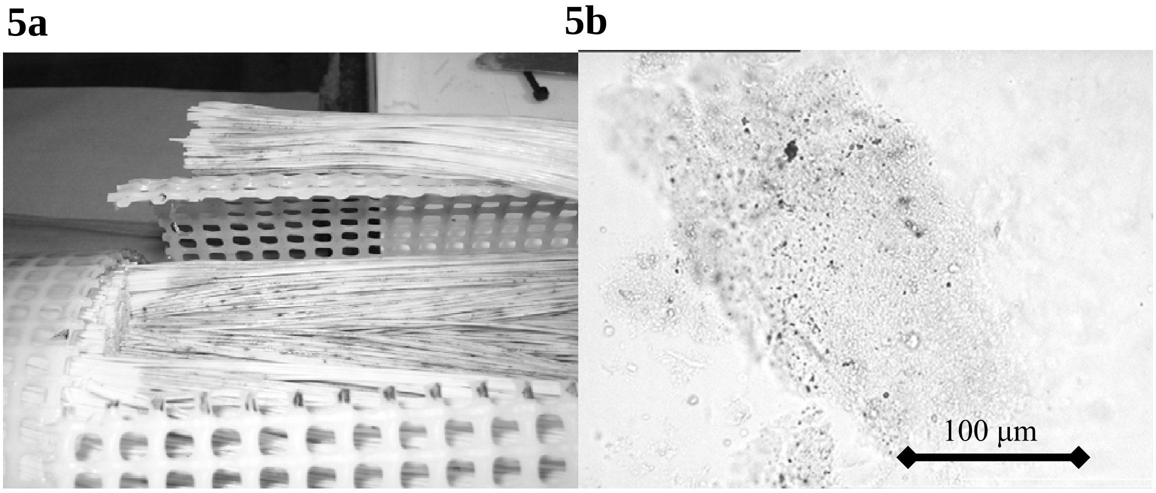

Figure 3.5 Fouled fibres a) and optical microscope image of reinsed surface foulants (b).

HPSEC chromatography (Figure 3.6) showed that whilst the foulants found were of similar molecular weight (MW) ranges, the NOM for samples A was at a higher level, especially in the 3.5 to 2 kiloDalton range (kDa) and had some larger >5 kDa MW material. The membranes were operated at pH 8.0 so the long elution times suggest aromatic and hydrophilic low molecular weight NOM. Molecular sizes for this method were as per the instrument's calibration provided and related to raw water NOM studies studies performed on multiple UK upland surface waters (Goslan et. al., 2002).

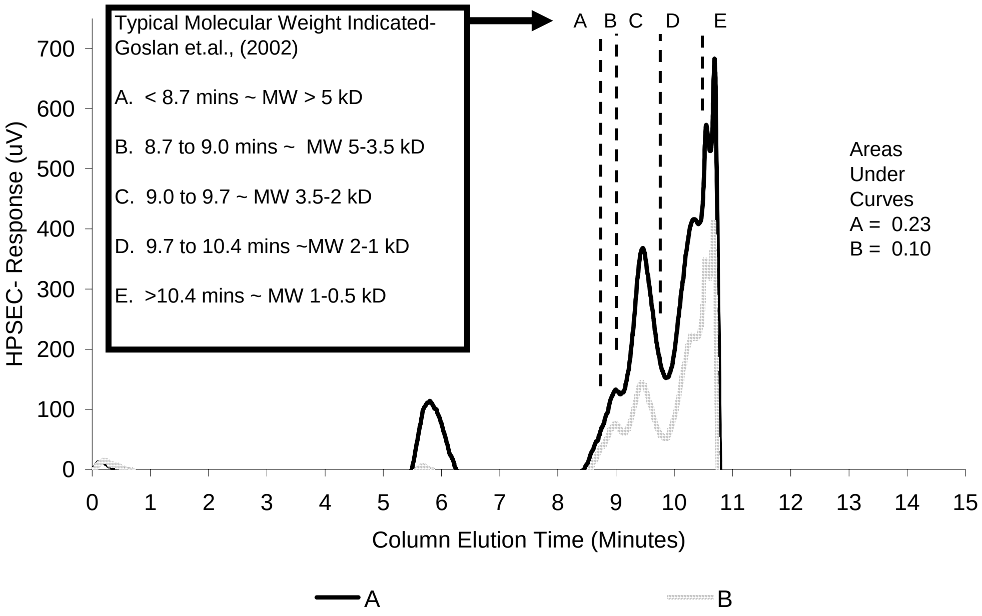

Figure 3.6 **HPSEC chromatogram from pre and post commissioning fibre cleaning elute.**

DOC measurements of the eluate allowed DOC eluted per unit area of membrane to be calculated. Figure 3.7 shows that the A samples were less fouled than B in terms of DOC but the specific ultraviolet absorbance (SUVA) at 254 nm units of per metre divided by the dissolved organic carbon (DOC) concentration (in milligrams per litre) indicated similar non- humic organic matter had fouled both sets of fibres but to a different degree. This suggests differences in operational factors, such as backwashing and cleaning, rather than water character were primarily responsible for the initial fouling.

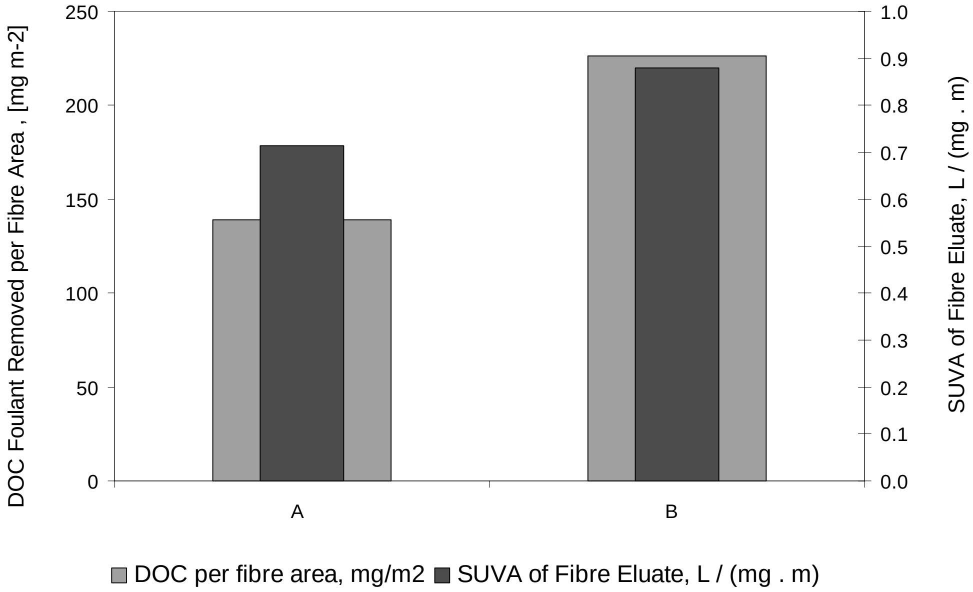

Figure 3.7 DOC and SUVA concentrations for pre (A) and post (B) commissioning elutes.

Raw water turbidity was 1.07 NTU decreasing to 0.6 NTU in the membrane feed and < 0.1 NTU in the membrane outlet permeate. The suspended solids in the membrane feedwater were reported as < 3 mg L^{-1}, with silica as high as 1.5 mg L^{-1} indicating the inorganic nature of the primary and secondary filter carry over being retained on the membranes. NOM was not at high enough concentrations in the source water to be expected to foul the membrane directly. Raw water DOC reduced from 3 mg L^{-1} to 1 mg L^{-1} through upstream processes but was not further reduced by the membranes. The charge density experiment showed DOC in the source water to have a very low charge density of 0.168 meq g^{-1} DOC. At such a low charge density the effectiveness of pre-coagulation would be expected to be compromised, removing only part of the DOC in the NOM; fine particles passing onto the MF membranes would be likely to be colloidal inorganics with hydrophilic and charge-neutral NOM. Studies show that removal of this fraction greatly reduces MF fouling (Fabris et al., 2007), and that biofiltration is similarly effective (Persson et al., 2006).

Three automated Clean in Place (CIP) regimes were established at the time of site commissioning:

- CIP 1 constituted soaking with 10% sodium hypochlorite (NaOCl) at 30°C for 1800 seconds followed by 1800 seconds recirculation.
- CIP 2 employed sulphuric acid (H_2SO_4) and was applied every second CIP 1 for the same duration and temperature.
- CIP 3 used 0.5 wt% citric acid; it was planned to apply this CIP 1 after commissioning every 6th CIP 2, again for the same duration and at 30°C.

At the start of commissioning it was not always possible to achieve these concentrations, temperatures or soak times due to resource constraints and design restrictions on the equipment. During the reported pre-commissioning fouling event, CIP 1 was applied repeatedly and exclusively. CIP 1 and CIP 2 together did not achieve permeability recovery, despite 0.5 wt% sodium hypochlorite being reported from the literature as sufficient to remove EPS from heavily fouled MF- submerged MBR (SMBR) membranes with soak times of 3 hours (Nuengjamnong et al., 2006). It is thought that the presence of inorganic particles in the matrix with elevated NOM may have strengthened the biofilm's resistance to cleaning. Further tests were therfore requested by the plant operator to investigate the fouling and optimise citric acid CIP cleaning and the compare effect of inorganic and organic on the recoverability.

Further field study indicated that manganese was unlikely to be the cause of the black flecks of material observed from the autopsy flushing test (Figures 3.5a and 3.5b). Sodium hypochlorite is dosed for manganese removal on the secondary pressure filters, with 0.9 mg L^{-1} Cl_2 at pH 8.0 reducing raw water manganese levels from < 50 μg L^{-1} to 14 μg L^{-1}. From literature studies of similar waters the levels of manganese and oxidant found in this case are unlikely to be the main cause of the fouling (Takizawa et al., 2001).

3.6 PERMEABILITY RECOVERY TESTS

3.6.1 Box Behnken Test

From the fifteen Box Behnken tests, (as described earlier in section 3.4.4), the responses algorithms were calculated in coded and actual units for pre (A) and post (B) commissioning membrane fibres:

A Cleaning algorithms, pre-commissioning stage

Coded

$R\% = +37.54 - 8.83a - 12.55b - 5.61c + 20.08ab + 23.41ac + 5.76bc$

$SD = 17.04, R^2 = 0.723$, ANOVA $F = 3.5$

Actual units

$\mathbf{R\% = +245.89 - 133.66C - 2.01P - 4.26T + 1.12CP + 2.60CT + 0.01PT}$

B Cleaning algorithms, post-commissioning stage

Coded

$R\% = +48.92 + 7.67a - 4.60b - 11.82c + 7.83ab + 19.12ac + 6.19bc$

$SD = 17.39, R^2 = 0.600$, ANOVA $F = 2.0$

Actual units

$\mathbf{R\% = +125.82 - 55.8C - 0.95P - 2.5T + 0.43CP + 2.12CT + 0.01PT}$

Both algorithms are similar and indicate a positive recovery modified by the citric concentration less in the pre commissioning than in the post commissioning samples. There is a switch in influence from period soaked to temperature as the equation modifier from pre to post commissioning. Optimisation within the ranges used with the criteria of maximizing recovery was carried out and the most desirable recovery chosen for sensitivity analysis. These algorithms yielded the following optimums:

Pre-commissioning C' = 0.6 g L^{-1} P' = 36 min. T' = 15°C R% = 85%

Post-commissioning C' = 1.8 g L^{-1} P' = 90 min. T' = 35°C R% = 97%

Figure 3.8 shows that the initial pre-commissioning membrane is much more sensitive to period soaked and less sensitive to soaking temperature. These suggest that initial fouling of this type is best addressed by increasing the length of soak time to maximise cleaning efficacy. Concentration of cleanant has less influence and temperature of soak solution greater impact as the membrane plant becomes fully operational.

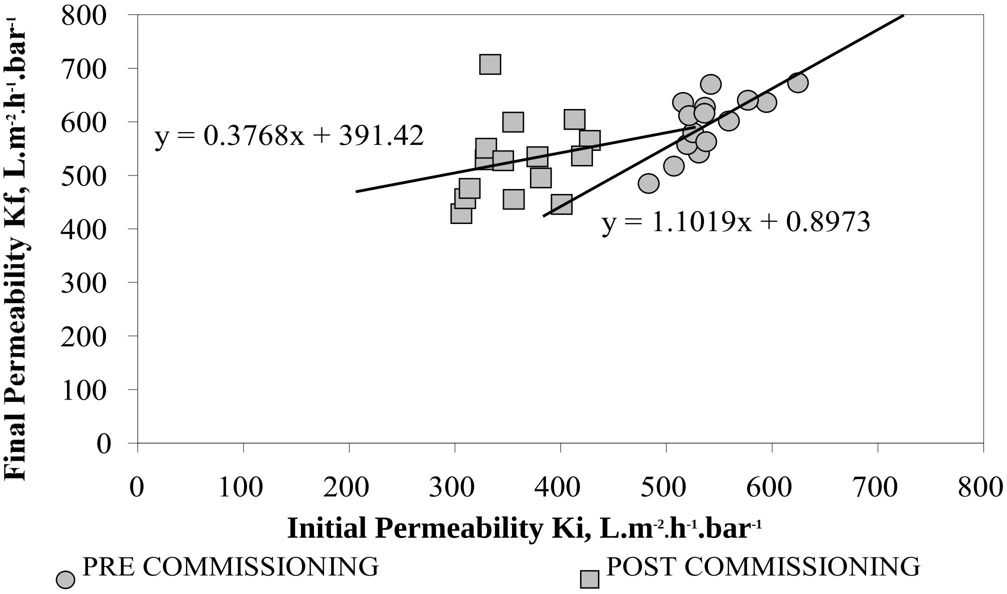

Figure 3.8 **Chemical cleaning response pre and post commissioning**

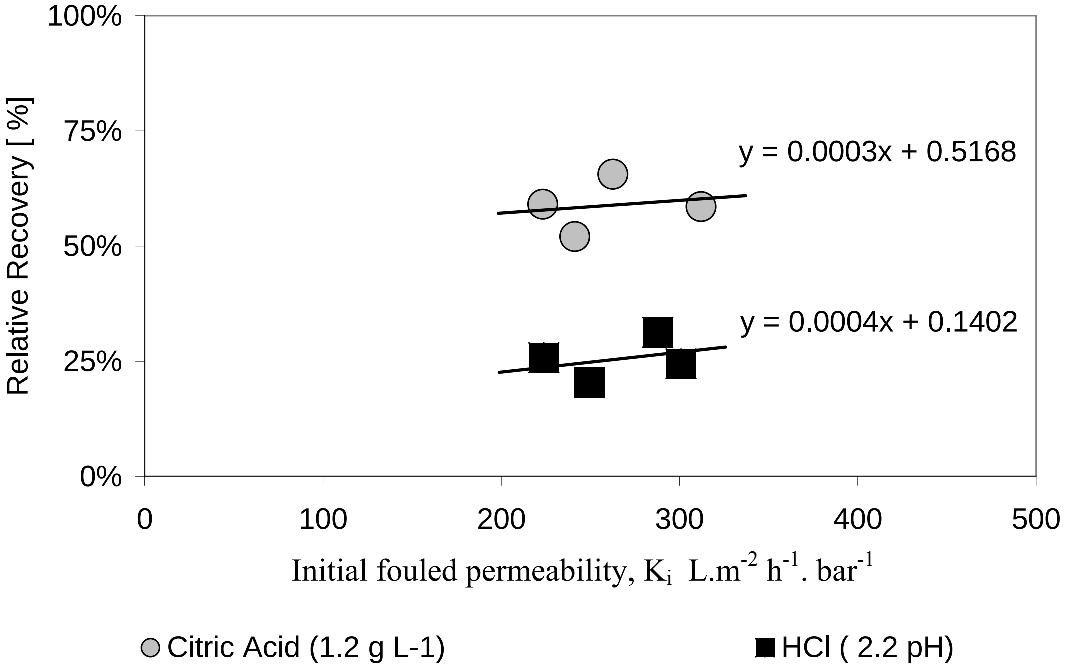

Figure 3.9 **Chemical cleaning recovery Citric and HCl.**

3.6.2 Protocol comparison of cleaning reagents

Figure 3.9 shows normalised recoverability for both cleanants for Test C with Citric (1.2 g.L^{-1}) and hydrochloric acid (2.2 pH) at 60 minutes, 20°C. The organic acid cleaning efficacy is more independent of the degree of fouling than that of the mineral acid. Figure 3.10 shows the initial fouled permeability for each rinsed membrane sample. For the pre commissioned sample A the foulants were readily rinsed off to provide a membrane permeability of 540 L.m^{-2}.h^{-1}.bar^{-1}, whilst for the post-commissioning tests B and C the permeability of the rinsed membrane was lower. For all samples, however, permeability was almost fully recovered with citric acid cleaning to a similar mean value of 520 to 580 L.m^{-2}.h^{-1}.bar^{-1} (Figure 3.11). This indicates that in the case of the pre commissioned fibres, permeability was wholly recovered by rinsing, whereas after commissioning the fouling layer was more tenacious and demanded a citric acid clean for permeability to be restored. In both cases, the on site backwashing, applied every 60 minutes, and daily sodium hypochlorite CIP used was not immediately effective at restoring permeability on the pre commissioning stage fibres.

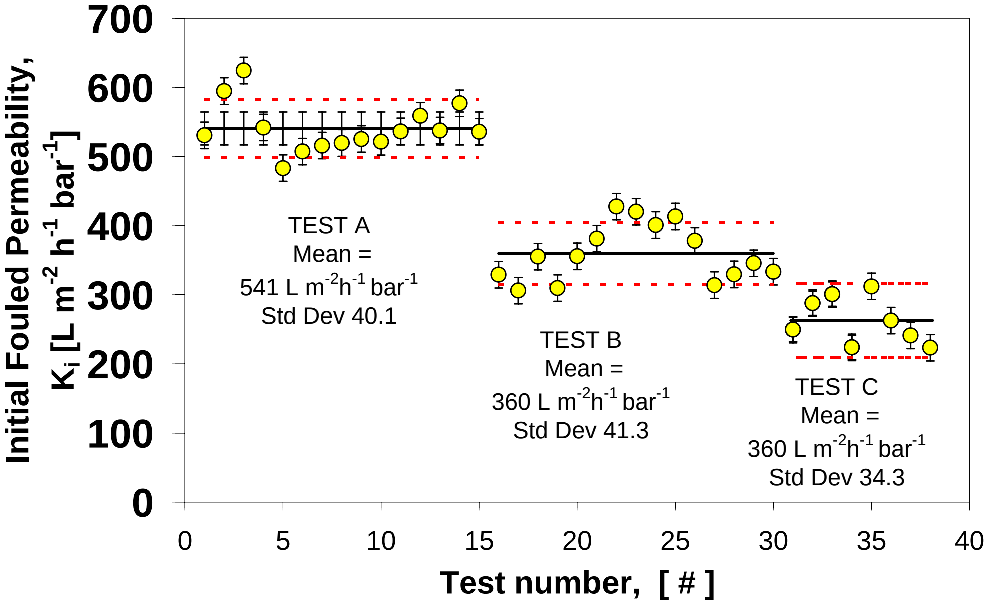

Figure 3.10 **Initial fouled permeability for each sample.**

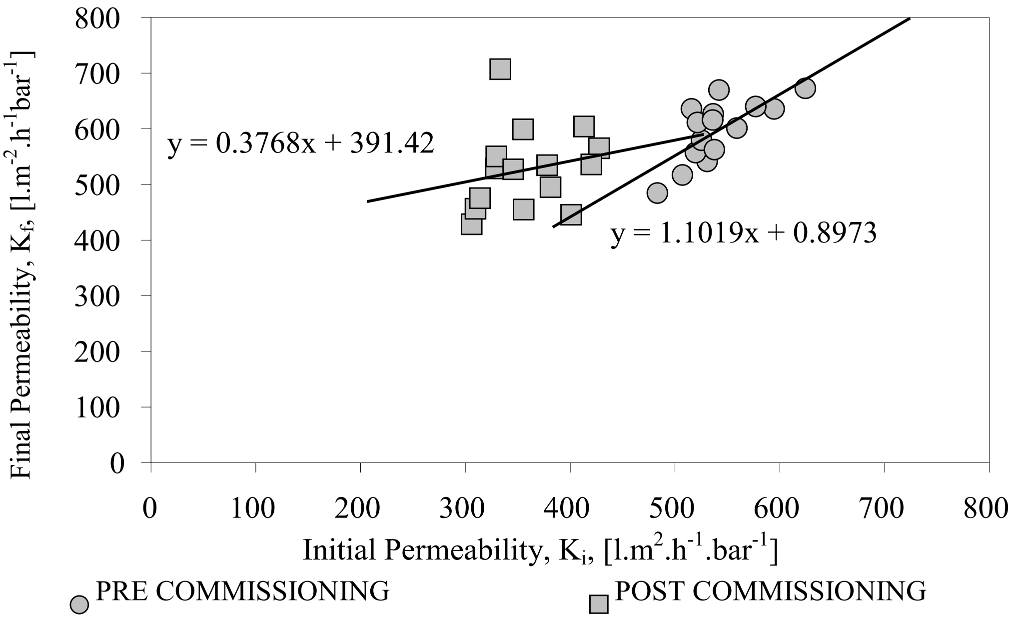

Figure 3.11 **Final recovered permeability for pre and post commissioning fibres.**

3.7 CONCLUSIONS

Autopsy on an MF membrane module challenged with upland surface water at commissioning and experiencing severe flux decline rates, confirmed the presence of an easily rinsed biofilm containing particulate minerals. The biofilm was characterised by organic and colloidal solids, as indicated by the change in DOC and turbidity levels across the plant. A second membrane sampled after commissioning was found to be fouled with a more tenacious foulants than the pre-commissioned fibres and exhibiting a lower surface concentration of DOC but a similar SUVA. It could be concluded that even over a short period cleaning response is variable.

The impact on membrane permeability of chemical cleaning was evaluated using a Box Behnken experiment designed to identify the differing optimum conditions for pre and post-commissioning membrane samples. Acid was found to be more effective than the hypochlorite CIP applied on site, recovering the permeability to 520 to 580 $L.m^{-2}.h^{-1}.bar^{-1}$ regardless of the degree of fouling and suggesting that organometallic chelating influences the biofouling tenacity for this type of membrane material. This method demonstrates cleaning response is quantifiable.

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CHAPTER 4 EFFECTS OF VARIABLE FLOW AND FOULING ON MBR CLEANING

Porcelli, N., Poyotas, J. and Judd, S. (2007). Effects of Variable Flow and Fouling on MBR Cleaning. *Proceedings of Aachen Water and Membrane Conference,Membrantechnik Tagun, Aachen, FDR, 31st October 2007*

Abstract

Fouling in hollow fibre submerged membrane bioreactors (MBR) can be attributed to the effects of mixed liquors on biofilm as well as the effects of colloidal and soluble products. This study looks at the positional effects on fouling in a membrane bioreactor. Fouled membranes were removed from a MBR module processing municipal wastewater. Bench scale permeability recovery studies using a two stage chemical cleaning regime, under conditions of standard chemical strength, temperature and soak time, were undertaken on fibres selected from specific sections of the module. The results showed the permeability recovery to be linearly related to the initial permeability for both chemical cleans performed (alkaline hypochlorite and citric acid), indicating that heterogeneity in flow and flux distribution is likely to be exacerbated by lower permeability recovery at lower local operational fluxes.

4.1 INTRODUCTION

Membrane fouling in a membrane bioreactor (MBR) is due to interaction between components of the biofluid and the membrane. When the membrane is in contact with the biological suspension, a deposition of biosolids occurs with implications of flux drop. The fouling due to biofilm can be partially removed; recovering permeability by physical cleaning with periodical blackwash is typical for hollow fibre membranes (Judd, 2006). When fouling is due to adsorption of matter into the membrane pores this fouling is irrecoverable, necessitating chemical cleaning. With prolonged membrane use the possibility of unrecoverable fouling increases, giving the membrane lesser chance for permeability recovery (Chang et al., 2002). In this context it is reported (Defrance et al., 2000) that fouling is due to mainly the mixed liquor suspended solids and the colloids.

Several studies have shown soluble fraction in the mixed liquor solids (MLSS) to be critical in the membrane fouling process; the colloidal and dissolved fractions have been shown to contribute 75% and the remainder associated with suspended solids (Bouhabila et al., 2001). Other studies have shown the contribution to the membrane fouling of the soluble products to be up to 50% (Wisniewski and Grasmick, 1998);

though fouling is obviously affected by system hydrodynamics (Al-Malack et al., 1996; Tardieu et al., 1998; Defrance and Jaffrin, 1999).

This study looks at the impact of the chemical cleaning protocol on membrane permeability recovery. Hollow fibre membrane fibres were removed from an existing MBR module that had been operated under known conditions, to assess the initial fibre permeability and the potential permeability recovery following chemical cleaning.

4.2 MATERIALS AND METHOD-MBR STUDIES

4.2.1 Membrane bioreactor pilot-plant used at Cranfield

The pilot MBR is shown in Figure 4.1. It consisted of an aerobic bioreactor with a vertically-oriented hollow fibre membrane module. The effective volume of bioreactor was 1.47 m^3 and the total membrane area 20 m^2, with an approximate packing density of 500 m^2 m^{-3}. The membrane module was braid-reinforced 2 mm diameter polysulfone/ polyethersulfone hollow fibre of sub 0.3 µm pore size. Coarse bubble aeration at the base of the module was applied on an alternating 30 s on -off basis. Influent was pumped from the feed tank to the bioreactor to maintain a constant level in the bioreactor. Permeate was extracted by suction pump and collected in the permeate tank. The sludge retention time (SRT) ranged from 26 to 33 days, with the plant operated in constant permeate flux mode. The filtration cycle comprised 10 minutes permeation and 30 seconds backflushing. Operating parameters such as permeate flow, permeation time and backflushing interval were set locally. Permeate flow, air flow, permeate pressure and turbidity were measured in real time, with data simultaneously recorded using a Squirrel-800 data logger. The pilot MBR trials were carried out at the Pilot Plant Hall on the Cranfield University Sewage Treatment Works site from early April to early August, 2006, during which time the MBR was run continuously except for short periods of necessary system maintenance. Municipal wastewater screened influent (<2 mm) was used.

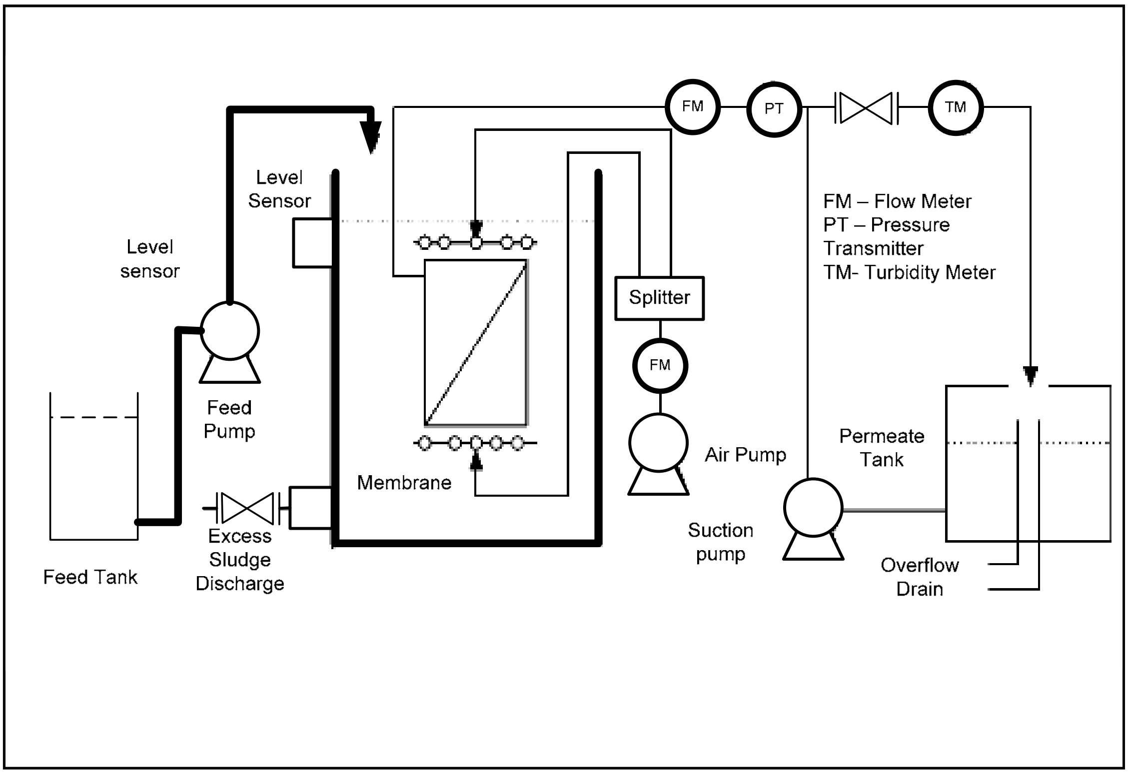

Figure 4.1 **Schematic MBR plant**

After two months operation without chemical cleaning fibres were taken from different locations of the original module, Figure 4.2 shows the locations where sample fibres were taken: the original module was divided into six sections per face, three in the horizontal plane and another three in vertical plane.

Bench test modules were prepared in dead end configuration by taking a sample of fibres from the different zones in the original membrane module. These were assembled in test modules (Figure 4.3) of ten fibres each of 500 mm length with 45 mm of potted fibre length.

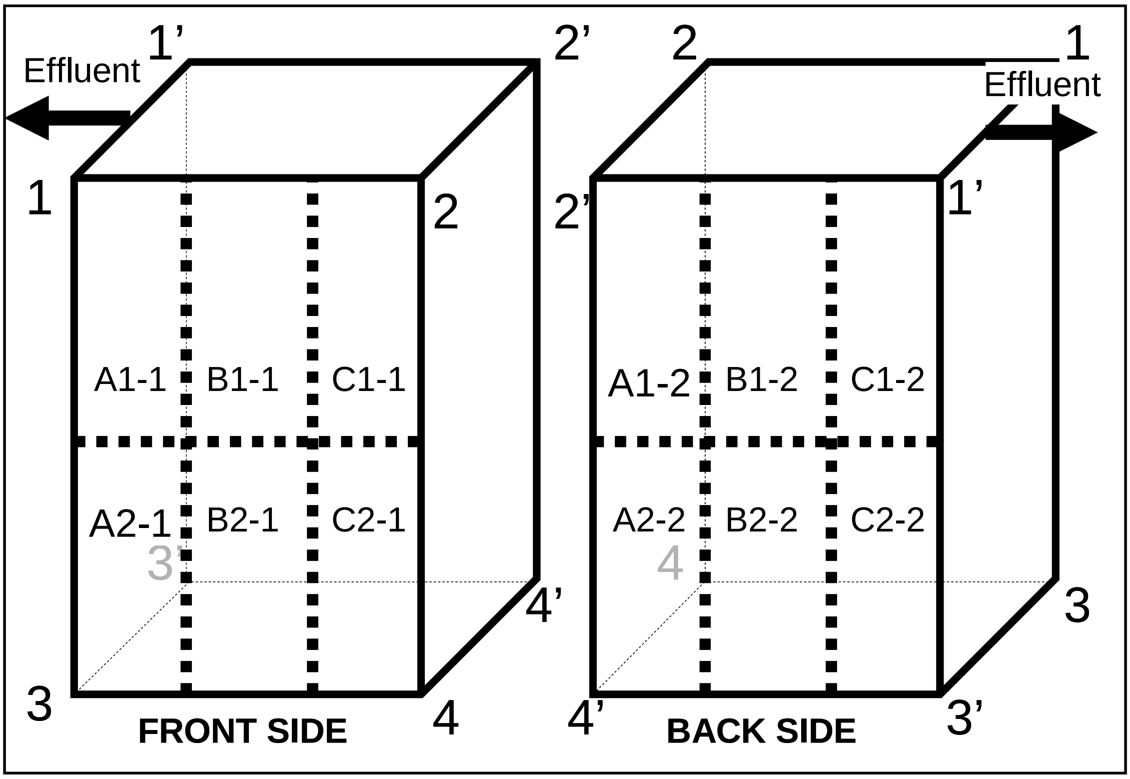

Figure 4.2 **Membrane side view showing subdivisions**

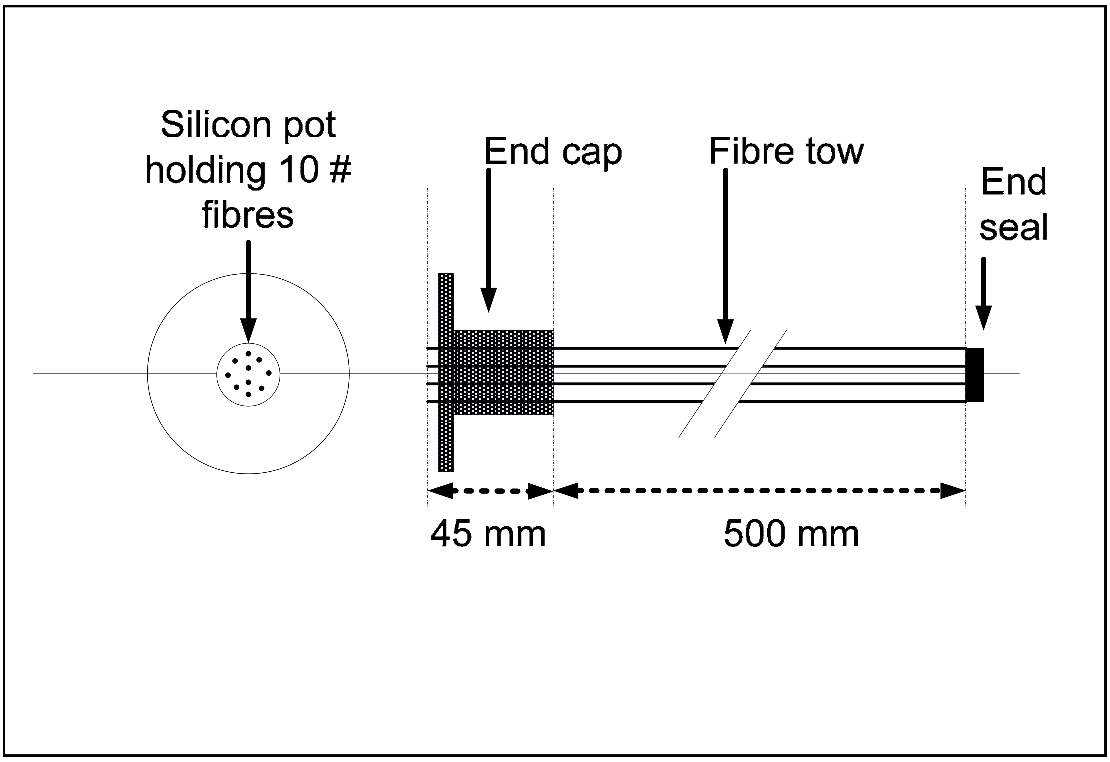

Figure 4.3 **Bench test module assembly detail**

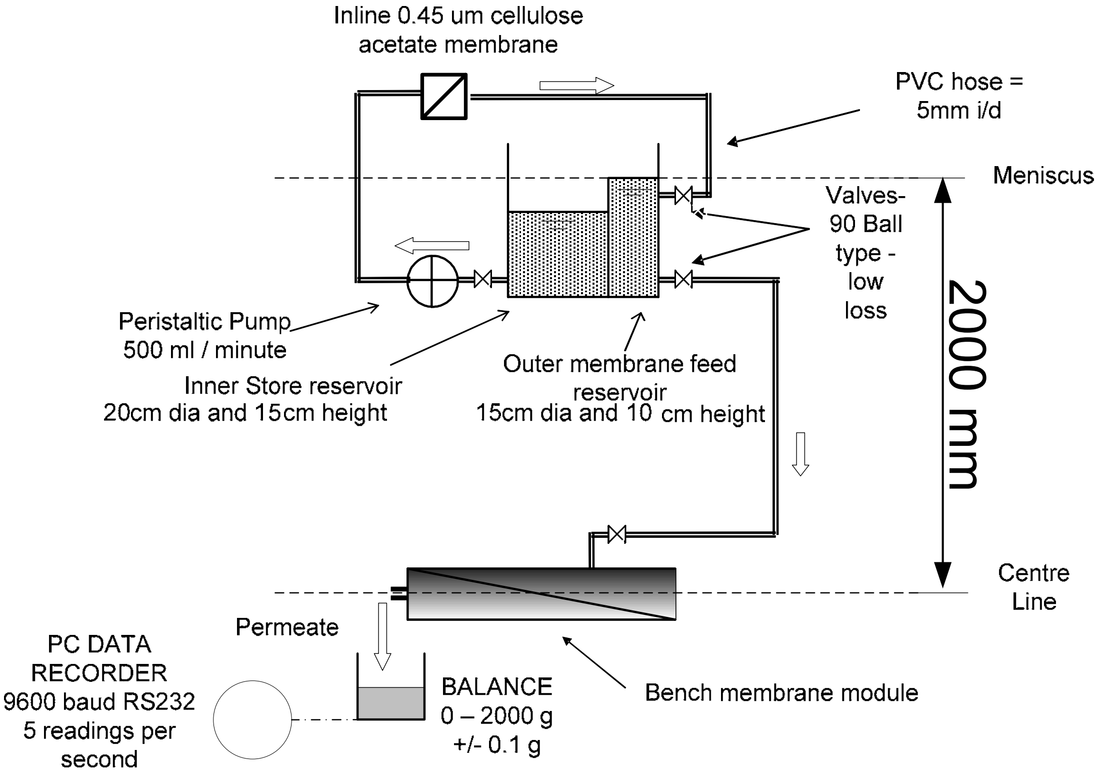

Figure 4.4 **Permeability test equipment schematic**

4.3 SAMPLING AND ANALYTICAL DETERMINATIONS.

Permeability testing was carried out on the assembled modules, each representing a different section of the original module. The experimental rig (Figure 4.4) gives a constant head of distilled feed water. The inner cylinder overflow was 2000 mm from the central axis of the module providing approximately 2 metres of hydrostatic head. As the flow rates relative to the tubing size and rig were small and similar for each test, dynamic headlosses were not considered. The feed was connected to the sealed potted end of the fibres with the permeate discharge to atmosphere.

A standard experimental procedure for the permeability tests on all the sample modules was followed. The membrane was connected to the head. Flow was established and the collected permeate mass recorded over ten minutes. The flux (J) was calculated from the rate of flow at the TMP (ΔP m) over the ten 500 mm long fibres. Permeability (K) is given by:

$$K = \frac{J_w}{\Delta P_m}$$ **Equation 4.1**

where

K = permeability (L m^{-2} h^{-1} bar^{-1})

J_w = flux (m h^{-1})

ΔP_m = Trans Membrane Pressure (ΔP m or TMP) bar (1 bar = kPa 10^5)

Permeability recovery (% R) can be expressed in either relative (% R_{rel}) or absolute (% R_{abs} terms as given in equations 4.2 and 4.3 and for comparative purposes we calculated recovery using both in this experiment:

Relative

$$\%R_{rel} = \frac{K_f - K_i}{K_v} \cdot 100$$ **Equation 4.2**

Absolute

$$\%R_{abs} = \frac{K_f}{K_v} \cdot 100$$ **Equation 4.3**

where:

K_f = Final permeability measured after chemical cleaning (L.m^{-2}.h^{-1}.bar^{-1}),

K_i = Initial measured permeability (L.m^{-2}.h^{-1}.bar^{-1}), and

K_v = Virgin membrane permeability (= 562 L.m^{-2}.h^{-1}.bar^{-1}).

After measuring initial fouled permeability, primary cleaning with 1% NaOH and 0.1% NaOCl solution was carried out. The permeability test was repeated and the experiment continued with another cleaning stage using 1% citric acid solution. Both cleaning stages were for 2 hours at a constant 15°C.

4.4 RESULTS AND DISCUSSION

The permeability from the different zones is shown in Table 4.1. The table shows data for initial permeability (K_i), permeability following alkaline hypochlorite cleaning ($K_{f\,a}$) and that following the citric clean ($K_{f\,b}$). It is evident the data set A1-1 is anomalous, since the cleaned membrane permeability is higher than the virgin. This could indicate a loss of membrane integrity. EDX/SEM analysis suggests damage by silica particles (Figure 4.5). Even ignoring this obvious anomaly, the remaining data shows significant variation in permeability with location, there being a 29% standard deviation around the mean value of 81 $L.m^{-2}.h^{-1}.bar^{-1}$ for the fouled membrane and 39% for the cleaned membrane. Alkaline hypochlorite improves permeability by almost a factor of 3 whilst the secondary citric clean improves permeability by a further 30%.

All data other than the anomalous data set are plotted in Figure 4.6 as relative and absolute permeability recovery against initial permeability respectively. Although somewhat scattered ($R^2 = 0.65$ to 0.80), the data indicate that permeability recovery is linearly related to the initial permeability for both relative and absolute permeability and for both chemical cleaning reagents.

Table 4.1 **Permeability measured across cleaning for different zones of the MBR.**

SAMPLE	**K_i** L.m^{-2} .h^{-1}. bar^{-1}	**K_f a** L.m^{-2} .h^{-1}. bar^{-1}	**K_f b** L.m^{-2} .h^{-1}. bar^{-1}
A1-1	358.43	740.24	873.62
A2-1	115.15	403.49	505.31
B1-1	105.81	319.89	446.01
B2-1	61.66	203.74	240.19
C1-1	114.52	262.31	367.40
C2-1	73.35	224.41	300.21
A1-2	57.49	110.26	153.05
A2-2	72.54	220.42	285.43
B1-2	52.59	102.46	137.00
B2-2	75.62	209.63	258.69
C1-2	79.88	211.36	286.79
C2-2	75.80	193.31	260.86
Average*	81	227	298
Std. Dev.*	23	89	116
Std. Dev., %*	29%	39%	39%

*All data excluding that for Section A1-1

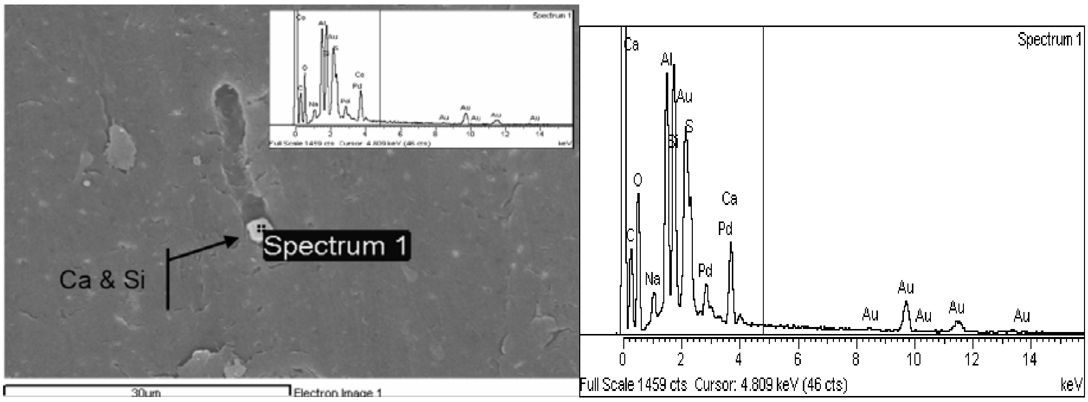

Figure 4.5 **SEM with elemental EDX analysis, SectionA-1-1 of MBR**

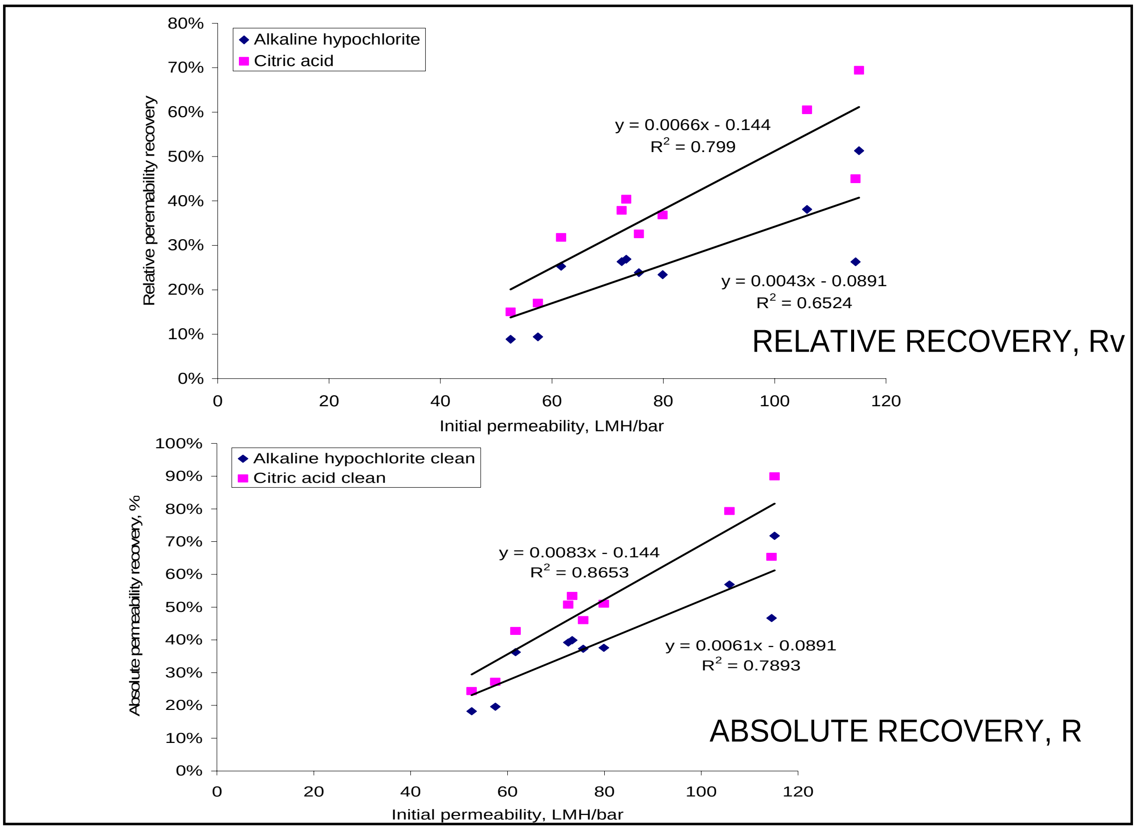

Figure 4.6 **Relative and absolute permeability recovery vs initial permeability**

Section A1-1 was taken from the zone nearest the intake. Fibres taken from the region around both intakes (Sections A1-1 and C1-1) of the module have higher permeability, suggesting these regions provide greater fouling suppression through, presumably, either more effective turbulence promotion or routine physical cleaning (by backflushing). Some sections have permeability values half those of the membranes sampled from the inlet regions. Regional fouling imposes greater hydraulic stress on the less-fouled regions, which may have contributed to the integrity failure in Section A1-1.

4.5 CONCLUSIONS

Work investigating the cleaning efficacy for an operating MBR membrane module indicates that recovery of permeability is roughly proportional to the initial permeability of the fouled membrane. Damage to the membrane in one section may have contributed to the recorded heterogeneous distribution of membrane permeability, exacerbated by the reduced permeability recovery at higher fouling levels. The work performed emphasises the importance of maintaining even distribution of flow throughout the bulk of a membrane module during both the operational and backflush cycle.

The nature of the localised damage to the membranes seen here was evidently due to installation factors. It is possible with good design and equipment selection to ameliorate this. For instance, ingress of refractory inorganic particles large enough to damage the fibres nearest the inlet may be reduced by the inclusion of a microstrainers. Typically a 1 to 2 mm mesh size micro-strainer carefully sized to suit the hydraulic profile, as well as upstream sand/grit traps can help maintain the integrity of the membranes

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CHAPTER 5 IMPACT OF CLEANING PROTOCOL ON MEMBRANE PERMEABILITY RECOVERY: A SENSITIVITY ANALYSIS.

Porcelli, N., and Judd, S. (2009). Impact of Cleaning Protocol on Membrane Permeability Recovery: A Sensitivity Analysis. In press *J.AWWArF:*

Abstract

The results of a 4-year study of chemical cleaning of ultra/microfiltration membranes for inland potable water treatment are presented. The study made use of a bespoke test cell in which sample membrane fibres extracted from modules from full-scale operating membrane plants were examined. The test entailed cleaning, using the reagents employed at each of nine sites, under controlled conditions of reagent concentration (C'), soak period (P') and temperature (T'). Data were analysed using a standard response surface model to identify optimum cleaning protocol paramet er values.

Results revealed the impact and order of importance of the three parameters (C', P' and T' above) to vary both between sites and temporally, making identification of the most appropriate prescriptive cleaning protocols extremely challenging. Optimisation of plant operation and maintenance in practice demands methods for recording and tracking changes in cleaning efficacy, such as the one used in this study, as an adjunct to initial fouling propensity assessment through piloting.

5.1 INTRODUCTION

Ultrafiltration and microfiltration membranes have been applied extensively to potable water treatment globally to achieve combined disinfection and clarification. Most of the work performed on membrane technology for this duty has been based on fouling. The three most extensively reported aspects of membrane fouling have concerned mechanisms (Vaisanen et al., 2002; Allie et al., 2003; Heijman et al., 2005; Yamamura et al., 2007), foulant characterisation (Amy 2008; Lee et al., 2004), and modelling (Bartlett et al., 1995; Zondervan et al., 2008; Chen et al., 2003).

There has been less work on membrane cleaning. Early heuristic studies (Tragardh, 1989) have been followed by more rigorous investigation for industrial membrane applications (Silalahi and Leiknes, 2009; Kazemimoghadam and Mohammadi, 2007), but less attention has been paid to municipal waters. A number of studies have incorporated cleaning based on pilot and full-scale plant operation, designed to achieve process optimisation (Karimi et al., 2002; Panglisch et al., 1997) through enhanced permeability recovery (Table 5.1). Despite this, a review of fouling mitigation (Hilal et al., 2005) has shown cleaning studies at pilot and full scale to be limited in number and scope, as well as being constrained in wider applicability. Reports have included qualitative cleanant responses under controlled fouling conditions (Bartlett et al., 1995). More rigorously controlled and quantitative membrane cleaning studies conducted at bench scale has been limited either by the use of model foulants (Heijman et al., 2005) and/or to a single membrane or membrane material (Strugholtz, 2005).

Recent cleaning work based on the development and testing of a state-based kinetic cleaning model has advanced the science of cleaning. This work has aimed to establish dynamic relationships between various different cleaning agent parameters (Zondervan and Roffel, 2007). These data have arisen from a UF fibre assembly fouled by a single surface water source. Ultimately, such a scientific approach may address the limitation of fouling models, such as the a Unified Membrane Fouling Index (Huang et al., 2009) for low pressure membranes, which are challenged both by the difficulty in predicting fouling from bench scale fouling studies and the lack of information on chemical cleaning. Specifically, standardisation of the membrane permeability recovery from chemical cleaning is thus far beyond the capability of membrane permeation kinetic modelling.

There is evidently a need to address the representation of permeability recovery from chemical cleaning using (a) data from of a wide range of membranes fouled through practical operation at full scale, and (b) a rigorous quantitative assessment of cleaning efficacy using classical statistical modelling. This four-year study aimed to achieve these objectives using a bespoke bench-scale test (Porcelli et al, 2009) to study the influence of cleaning parameters on membranes taken from full-scale plant at various UK locations (Appendix A, B and C). A response surface statistical analysis was designed to optimise and explore the sensitivity of MF and UF permeability recovery to the ranges of typical cleaning regimes.

Table 5.1 MF and UF Literature: Cleaning Studies

REF	***SYSTEM***	***MATERIAL***	***SOURCE, TREATMENT AND FOULANTS***	***PLANT OPERATION***	***CLEANING AGENT AND RECOVERY RATIO R***				
					i	***ii***	***iii***	***iv***	***v***
Strugholtz, et al., 2005 [1]	CT-UF	PES	Pilot plant, lowland reservoir water (TOC = 2.9 mg L^{-1}) plus with precoagulation (PACl 1.5 mg L^{-1})	Constant flux J_{norm} = 50 $L.m^{-2}.h^{-1}$ Permeability K = 343 $L.m^{-2}.h^{-1}.bar^{-1}$	NaOH, pH 12 R = 2.4 @ 20°C, 2.5 @ 40°C	HCl pH 3 R = 1.4 @ 20°C, R =1.5 @ 40°C	250 mg l^{-1} H_2O_2 R = 1.0 @ 20°C	NaOCl 50 mgL^{-1} R = 1.6 @ 20°C, 2.5 @ 40°C	Citric Acid 2000 mgL^{-1} R = 1.5 @ 20°C
Yamamura et al., 2007 [2]	HF-MF	A- PVDF	Pilot Plant MF- 50 days in service UF- 30 days in service River water. Turbidity= 16.2 NTU DOC = 1.4 mg L^{-1}	Constant flux J_{norm} = 27 $L.m^{-2}.h^{-1}$ Permeability K = n/a TMP = 20 to 160 kPa	NaOH pH12 R = 3.0	HCl pH2 R = 2.0	EDTA 20mM R = 1.5	0.5wt% oxalic acid R = 2.2	NaClO 700ppm R = 7.0
	HF-MF	B- PE			R = 1.30	R = 1.3	R = 1.3	R = 1.4	R = 1.25
	CT-UF	C- PAN			R = 1.75	R = 1.25	R = 1.5	R = 1.3	R = 2.65
Zondervan and Roffel, 2007 [3]	CT-UF	PES	Canal surface water: pH 7.5, turbidity= 13.5 NTU, DOC = 8.8 $mg.L^{-1}$	Constant Flux J norm = 100 $L.m^{-2}.h^{-1}$ TMP = 35 to 50 kPa	NaOH $\eta / r_i = 0.24$	HCl $\eta / r_i = 0.44$	H_2O_2 $\eta / r_i = 0.30$	H_2SO_4 $\eta / r_i = 0.26$	Citric Acid $\eta / r_i = 1.00$
Porcelli et al., 2009	HF-MF	PVDF	Upland surface impounded reservoir water with precoagulation (PACl 2.0 mg L^{-1}) pH 5.5, turbidity < 2 NTU, DOC < 2 $mg.L^{-1}$	Constant Flux J_{norm} = 96 L m^{-2} h^{-1} Permeability K = 180 L m^{-2} h^{-1} bar^{-1} TMP = 30 to 60 kPa		HCl: 90 minutes soak at 35°C R = 1.37			Citric acid: 90 minutes soak at 35°C R = 1.90

[1] Alkali then acid (NaOH then HCl) gave best recovery (R = 3.4); [2] All cleaning solutions soaked at 20°C for 24 hrs; [3] ratio of cleaning effectiveness (η) to cleaning rates (r_i), from Figs. 5 and 6 of this paper. ***R = J_1 after clean / J_o before clean.*** η = Cleaning effectiveness; r_i = instant cleaning rate (1/min)
CT capillary tube; HF hollow fibre; MF microfitration; UF ultrafiltration; PAN polyacrylonitrile; PE polyethylene; PES polyethyl sulphone; PVDF polyvinyldene difluoride; DOC dissolved organic carbon; TOC total organic carbon; PACl polyaluminium chloride; J_{norm} = flux normalised to 20°C.

5.2 MATERIALS AND METHODS

5.2.1 Materials: plants sampled

Hollow fibre (HF) and capillary tube (CT) membrane modules were sampled from eleven existing full-scale membrane ultrafiltration (UF) and microfiltration (MF) plants from across the UK treating both groundwater from boreholes and springs and impounded upland surface waters (Table 5.2). The plants were commissioned at different periods between 1999 and 2005 and were sampled sequentially, and so had different operational times in service. Table 5.3 shows the operational flux for each and the transmembrane pressure (TMP) as well as the length of time the eleven sampled modules had been in service with their normal backwash and cleaning regimes.

The HF-MF membranes were of polyvinyldene difluoride (PVDF) and polypropylene (PP), and CT-UF membranes were polyethyl sulphone (PES). The CT-UF membranes were exclusively from two groundwater Plants A and C. Both submerged MF systems (H and I) and pumped (B, E and G) membrane systems were sampled. On Plant F the membranes were tested at pre (F1) and post (F2) commissioning stages. Plant "G" was tested relatively early in its life (G1) and again two years later (G2).

Details of the backwash and clean in place (CIP), chemically enhanced backwash (CEB) used on the plants are included in Table 5.3. These protocols were based largely on the advice of the membrane supplier, modified with operational experience. Generally, the processes were selected following piloting, with pre-treatment and membrane type determined by source water quality deviation.

PVDF membranes were cleaned primarily using mildly basic oxidant hypochlorite, since this material is not alkaline resistant. The PP membranes were generally older installations and used on humic-rich upland surface waters; they were cleaned routinely with heated caustic soda washes coupled with alkaline solutions of strong oxidants (H_2O_2). The frequency of these secondary cleans was determined by the process operators depending on the organic content of the water and the risk of algal blooms from monitoring raw water temperatures. The more chemically-resistant PES CT-UF fibres (Plants A, C and D) had routine chemically-enhanced backwash (CEB) cleans with caustic solutions (NaOH) at similar temperatures (11°C or 15°C) but different soaking times and without recirculation. The groundwater plants (Plants B and D) were followed up with an organic citric acid as a CIP to remove periodic mineral-organic complex fouling.

Table 5.2 Plant feedwater quality, pretreatment and membrane information

Code	*Water Type*	*Membrane*[1]	*Config.*[2]	*Pre-Treatment*
A	Groundwater: single borehole. pH = 7.1, turbidity <5 NTU	UF - PES	CT	Coagulation (PACl 4 mg L^{-1}), clarification, GAC, pre-chlorination
B	Groundwater: multiple boreholes. pH = 7.2-7.5, turbidity = 1.4 NTU	MF - PVDF	HF	Microstrainer (500 µm), pre-chlorination
C	Groundwater: borehole and multiple springs. pH = 7.8, turbidity = 0.17 NTU	UF – PES	CT	Microstrainer (30 µm), pre-chlorination
D	Groundwater: single borehole. pH = 7.1, turbidity < 5 NTU	UF - PES	CT	Ozonation, GAC, pre-chlorination
E	Upland surface-impounded reservoir. pH = 6.9, turbidity = 0.71 NTU	MF - PP	HF	microstrainer (30 µm)
F	Upland surface-impounded reservoir. pH = 5.5 – 5.8, turbidity < 2 NTU	MF - PVDF	HF	pH correction, pre-coagulation (PACl 2 mg.L^{-1}), primary pressure filtration, pre-chlorination, 2nd stage pressure filtration
G	Upland surface-impounded reservoir. pH = 6.8, turbidity = 0.60 NTU	MF - PP	HF	Microstrainer (30 µm)
H	Groundwater: single borehole. pH = 7.1. turbidity < 0.2 NTU	MF - PVDF	sHF	Microstrainer, pre-chlorination
I	Groundwater: single borehole. pH = 7.3, turbidity = 0.1-0.5 NTU	MF - PVDF	sHF	Microstrainer, pre-chlorination

Table 5.3 Plant operating conditions

Plant Code	***Flux J*** *L m^{-2} h^{-1}*	***TMP,*** *kPa ΔP m*	***Days in service***	***Cleaning Regime:*** **Backwash** *(BW) – Water except where stated)* ***Cleaning (#):*** *Agent, Duration, Temperature, Frequency*
A	109	40 – 120, Avg. 85	946	**BW**: 50 s every 55 minutes, **Clean**- 1: NaOH, pH12, 20 mins at 11°C every 18 hrs. **Clean** -2: HCL, pH2, every 18 hrs
B	72	60 – 110, Avg. 75	432	**BW**: 30 s every 60 minutes, **Clean**-1: H_2SO_4, pH2, 100 mins at 15°C every 28 days **Clean**-2: NaOCl, 0.05%, 100 mins at 30°C every 28 days **Clean**-3: Citric Acid, 0.5%, 100 mins at 30°C every 180 days (ad hoc)
C	75	65 – 85, Avg. 71	1703	**BW**: 60 s every 60 minutes, **Clean**-1: NaOH, pH13, 60 min at 15°C every 28 days
D	125	40 – 100, Avg. 60	2010	**BW**: 50 s every 60 minutes, **Clean**- 1: NaOH, pH12, 20 mins at 11°C every 18 -48 hrs. **Clean** -2: HCl, pH2, 20 mins at 11°C every 36–120 hrs. **Clean**-3: Citric Acid, pH2.5, 24 hrs at 11°C every 180 days (ad hoc)
E	68	75 – 140, Avg.105	2585	**BW**: Air, every 60 minutes, **Clean**- 1: NaOH, 2.5%, 30 mins at 27°C every 18 -28 days. **Clean** -2: H_2O_2 5%, in 2.5% NaOH, 20 mins at 30°C as required/every 3days for summer algal suppression. **Clean**-3: HCl, pH2, Not routine- annual as necessary.
F	90	30 to 60, Avg. 45	F_1 = 24 F_2 = 98	**BW**: Air/water, 120 s every 90 minutes, **Clean**- 1: NaOCl, 10%, 30 mins soak then 30 mins recirc. at 30°C every 14 -28 days. **Clean** -2: H_2SO_4 30% pH 2, 20 mins at 30°C every second Clean #1. **Clean**-3: Citric Acid, 0.5%, 30 mins soak then 30 mins recirc. at 30°C every sixth Clean #2
G	41	60 – 120, Avg. 65	G_1 = 245 G_2 = 1003	**BW**: Air, every 60 minutes, **Clean**- 1: NaOH, 1 – 1.2%, 30 mins at 32°C every 14 days. **Clean** -2: H_2O_2 5%, in 1 – 1.2% NaOH, 30 mins at 30°C as required / every 3days for summer algal suppression. **Clean**-3: HCl, pH2, Not routine- annual as necessary.
H	160	n/a, Avg. 60	1553	**BW**: Air every 60 minutes, **Clean**- 1: NaOCl, 300 – 400 ppm, 30 mins Recirc. 60 mins soak at 10°C every 14 to 28 days. **Clean** -2: H_2SO_4 pH 2, 30 mins Recirc. 60 mins soak. at 10°C as required every 14 to 28 days.
I	122	n/a, Avg. 60	1827	**BW**: Air every 60 minutes, **Clean**- 1: NaOCl, 300 to 400 ppm, 30 mins Recirc. 60 mins soak at 10°C every 14 to 28 days. **Clean** -2: H_2SO_4 pH 2, 30 mins Recirc. 60 mins soak. at 10°C as required every 14 to 28 days.

5.3 EXPERIMENTAL PROCEDURES

5.3.1 Experimental analysis

The experiment was designed to explore the relationship between permeability recovery and cleaning parameters based on membranes sampled from the plants detailed in Tables 5.2 and 5.3, the data being subjected to statistical analysis. The experiment used a bespoke bench-scale test apparatus where sampled membrane filaments extracted from the modules were assembled into test modules allowing pure water permeability measurements at a constant head to be determined. Permeability measurements were conducted prior to and following chemical cleaning at a range of values for cleaning reagent concentration (C’), soak period (P’) and temperature (T’). Two plants were sampled more than once; G was sampled in 2005 (G1) and again in 2007 (G2) and F was sampled at the start of wet commissioning (F1) and three months later when the fibres had been in service for a short while (F2). In all cases, sampling of the modules followed a backflush cycle and preceded the routine chemical clean.

The apparatus and protocol has been reported elsewhere (Porcelli et al, 2009). A random sample of 12 to 15 fibres were extracted from throughout the module, both axially and radially, and rinsed with pure water prior to potting in a bespoke module for permeability measurement prior to and following the laboratory chemical cleaning test. Whilst permeability recovery has been reported in different ways (Table 5.1), the definition used for analysis of parameter sensitivity in the current study was the relative percentage permeability recovery R_v, with reference to the virgin membrane permeability K_v:

$$\%R_v = \frac{K_f - K_i}{K_v - K_i} \cdot 100$$

Equation 5.1

where K_i and K_f are the pure water permeability values in $L.m^{-2}.h^{-1}.bar^{-1}$ measured before and after chemical cleaning respectively.

5.3.2 Statistical analysis

Box Behnken determination (BBD) is a response surface methodology (RSM) and provides a statistical framework for generating regression-derived polynomial functions and the optimum recoveries for the three protagonist parameters (Box and Wilson, 1951; Bas and Boyaci 2007; Porcelli et al, 2009). These responses can be subjected to sensitivity analysis. The response function was validated with ANOVA error analysis. From the three independent C’, P’ and T’, twelve different permeability recovery responses (R_v %) were constructed; varied evenly around a central point in a matrix. The cleaning parameter values, corresponding to central point of this cube, represent the optimum values. The cleaning test was repeated three times at these values, such that there were fifteen experiments conducted for each treatment (Table 5.4). Table 5.4 shows details of these fifteen tests, with different cleaning regimes for plants indicated by subscripts following the plant sample identification letter. Plant A was tested with three reagents (A1 A2 and A3), plant G1 with four reagents (G1-1, G1-2, G1-3 and G1-4) and plant G2 was a later fibre sample tested with two reagents (G2-1and G2-2)

The response surfaces built from these recoveries allowed a best fitting factorial relationship to be determined. Each response had six factorial coefficients (a-f) and an intercept (M) as shown in Equation 5.2.

$$Rv\ \% = M + a.(C') + b.(P') + c.(T') + d.(C'P') + e.(C'T') + f.(P'T')$$

Equation 5.2

Table 5.4 Cleaning regimes for constructing RSM matrices.

Test code	***Cleaning agent***	***Strength (C' wt% or pH)***	***Soak duration (P' mins)***	***Soak temperature (T' °C)***
A_1	NaOH, pH 12, then citric acid	1 - 2%	30 - 90	10 - 40
A_2	HCl	2 - 3%	30 - 90	10 - 40
A_3	NaOH	0.2 - 0.8%	30 - 90	10 - 40
B	H_2SO_4	pH 2.0 - 4.0	60 - 120	10 - 40
C	NaOH	0.2 - 0.8%	15 - 45	10 - 40
D	Citric Acid	2.5 - 7.5 g L^{-1}	30 - 90	10 - 40
E	NaOH	0.075% - 0.125%	30 - 90	5 - 35
F_1	Citric Acid	0.6 - 1.8 g L^{-1}	30 - 90	5 - 35
F_2	Citric Acid	0.6 - 1.8 g L^{-1}	30 - 90	5 - 35
G_{1-1}	NaOH	1 - 2%	30 - 90	18 – 32
G_{1-2}	HCl	1 - 3%	30 - 90	18 – 32
G_{1-3}	H_2O_2	0.5 - 1.5%	30 - 90	18 – 32
G_{1-4}	NaOH pH 12 and H_2O_2	0.5 - 1.5%	30 - 90	18 – 32
G_{2-1}	NaOH	0.075% - 0.125%	30 - 90	5 - 35
G_{2-2}	NaOH pH 12 35°C 60 mins then H_2O_2 varied	0.5 - 2.0%	30 - 90	5 - 35
H	NaOCl varied then H_2SO_4 at pH 2.0 15°C 60 mins	250 - 450 mg l^{-1}	30 - 90	5 - 35
I	NaOCl varied then H_2SO_4 at pH 2.0 15°C 60 mins	250 - 450 mg l^{-1}	30 - 90	5 - 35

This algorithm permitted a sensitivity analysis for six parameters: (C', P' and T', and the combined two-factorial parameters C'P', C'T' and P'T') on permeability recovery. From the results a table was constructed to calculate the contribution to permeability recovery made by each of the six (a-f) factorial coefficients. Each factorial was varied over the experimental range whilst the other five parameters were held constant at their central point in order to give a normally distributed set of percentage permeability recovery variation results. The standard deviation of this range was then calculated to provide the variability value, representing the sensitivity of permeability recovery to each cleaning parameter. This was carried out for the C', P' and T' variables on all the algorithms generated.

5.4 RESULTS

Recovery values determined from permeability data (Table 5.5) were in reasonable agreement with literature data (Table 5.1). For Plant A with UF PES membranes, the range of recovery ratios for strong alkalis and mineral acids were similar to those reported by Strugholtz et al (2005); in studies based on a pilot plant fouled by reservoir water. Plants C and D, challenged with groundwater and operated at a relatively high throughput, yielded recoveries lower than those reported by Strugholtz et al (2005) but similar to those of Zondervan (2007), which relate to higher flux values but also higher solids loadings.

The HF–MF plants E and G were operated without pre-coagulation and at low fluxes. These plants showed lower recovery ratios than with similar order of parameter efficacy for the upland sources in the literature, with strong alkalis followed by peroxide the most effective for the older and newer fibres tested.

PVDF MF membranes in both pumped (B and F) and submerged systems (H and I) were susceptible to rapid fouling, as reported by Yamamura et al (2007). The membranes were readily cleaned with mineral acids and are suited to strong oxidant cleaning, (although the high recovery ratios reported in the literature were not replicated in these tests). Plant F, comprising PVDF membranes and fed with upland surface water, was tested at start-up when a temporary bio-fouling took place, and again following commissioning when more tenacious fouling (i.e. permeability decline) was greater but similar in chemical nature (Porcelli et al., 2009). Recovery was optimum when citric acid was used for this case. The more highly fouled PVDF membranes yielded a wider range of initial permeability values, as seen from the standard deviation data. This is indicative of more heterogeneous fouling, possibly linked with their higher operational flux and, in the case of the submerged membranes, higher solids loadings.

The use of modelled factorial responses allows exploration of the impact of these three key cleaning parameters on permeability recovery. Figures 5.1 shows the general plot of percentage ranges of standard deviations for all the tests carried out. The standard deviation values for R_v calculated as a percentage of the range of permeability recovery values for each test (17 in all from A1 through to I for the nine plants A-I).

Table 5.5 Permeability test results and modelled factorial response

	Measured Permeabilities, L.m^{-2}. h^{-1}.bar^{-1}				**Modelled Factorial Response**							
	K_v	K_i	K_f	***Recov. ratio***	R_v % = **M** + a(C) +b(P) + c(T) + d(CP) + e(CT) + f(PT)							
Test Code		***Mean,*** *st.dev.*	***Mean,*** *st.dev.*	J_1/J_0	**M**	*a*	*b*	*c*	*d*	*e*	*f*	**R^2**
A_1	300	**103**, *14*	**118**, 26	**1.15**	**8**	2.6	-0.3	1.8	0.44	-0.69	-0.01	**0.17**
A_2	300	**73**, *31*	**101**, *14*	**1.38**	**24**	-10.2	0.2	1.1	-0.08	-0.27	0.00	**0.71**
A_3	300	**64**, *22*	**141**, *15*	**2.20**	**35**	-65.3	0.0	-0.2	0.38	1.72	0.00	**0.80**
B	200	**149**, *47*	**156**, *46*	**1.04**	**-23**	10.8	0.5	-0.5	-0.19	0.15	0.00	**0.53**
C	400	**57**, *8*	**82**, *75*	**1.43**	**-1**	16.8	0.1	0.0	-0.50	-0.11	0.01	**0.39**
D	400	**275**, *13*	**320**, *27*	**1.16**	**78**	-7.9	-0.7	-1.9	0.07	0.29	0.02	**0.60**
E	300	**100**, *10*	**122**, *7*	**1.22**	**9**	-3.8	0.0	0.2	0.03	0.13	0.00	**0.51**
F_1	650	**541**, *40*	**598**, *55*	**1.10**	**246**	-133.7	-2.0	-4.3	1.12	2.60	0.01	**0.59**
F_2	650	**360**, *41*	**527**, *74*	**1.46**	**126**	-55.8	-1.0	-2.6	0.44	2.12	0.01	**0.59**
G_{1-1}	300	**96**, *7*	**111**, *10*	**1.16**	**27**	-3.8	-0.2	-0.5	0.02	-0.08	0.01	**0.24**
G_{1-2}	300	**111**, *11*	**134**, *16*	**1.21**	**-3**	0.7	0.0	0.4	0.04	-0.02	0.00	**0.62**
G_{1-3}	300	**106**, *6*	**113**, *9*	**1.06**	**-6**	13.0	-0.2	0.3	0.00	-0.57	0.01	**0.49**
G_{1-4}	300	**106**, *7*	**149**, *17*	**1.41**	**10**	-19.8	0.0	0.3	0.14	0.42	0.00	**0.75**
G_{2-1}	300	**135**, *12*	**140**, *15*	**1.04**	**8**	-5.1	-0.1	-0.3	0.08	0.18	0.00	**0.84**
G_{2-2}	300	**153**, *14*	**178**, *14*	**1.16**	**34**	-12.9	-0.2	-0.4	0.20	0.19	0.01	**0.35**
H	400	**183**, *31*	**220**, *42*	**1.20**	**106**	-0.2	-0.1	0.4	0.00	0.00	0.00	**0.19**
I	400	**227**, *8*	**284**, *15*	**1.25**	**24**	0.0	-0.1	-0.8	0.00	0.00	0.01	**0.68**

K_v virgin membrane permeability; K_i, K_f initial, final permeability; J_i, J_0 Initial and final flux, flux;

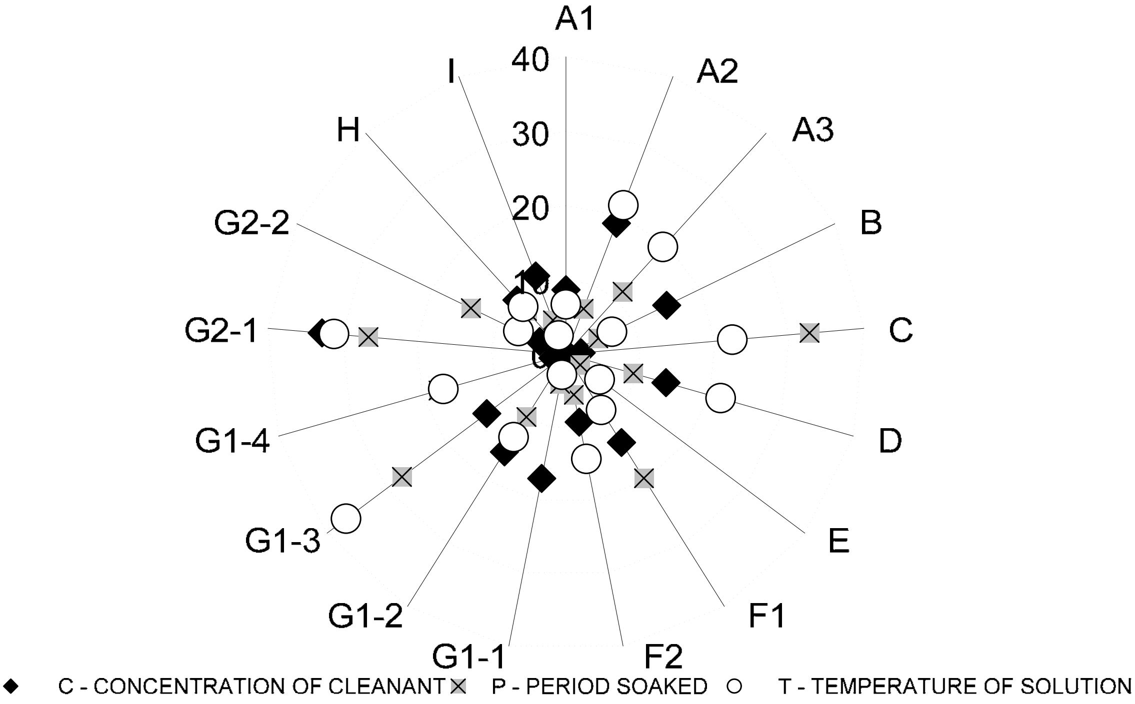

Figure 5.1 **Standard deviation in the modelled percentage recovery values calculated across the ranges of cleaning parameters (C'P'T'). All reported samples.**

Figures 5.2 to 5.6 also show for each case optimum percentage recovery (Rv%) for each of the cleaning experiments. In all these plots, the radial axis value (0 to 40 %) indicates the sensitivity of permeability recovery to each parameter, and thus reflects differences in the relative impacts of C', P' and T' on cleaning efficacy. Optimum recovery, given as a % for each test in Figures 5.2 to 5.6, was determined as the maximum value of R_v attained within the test ranges of the cleaning parameters C', P' and T' as indicated in Table 5.4.

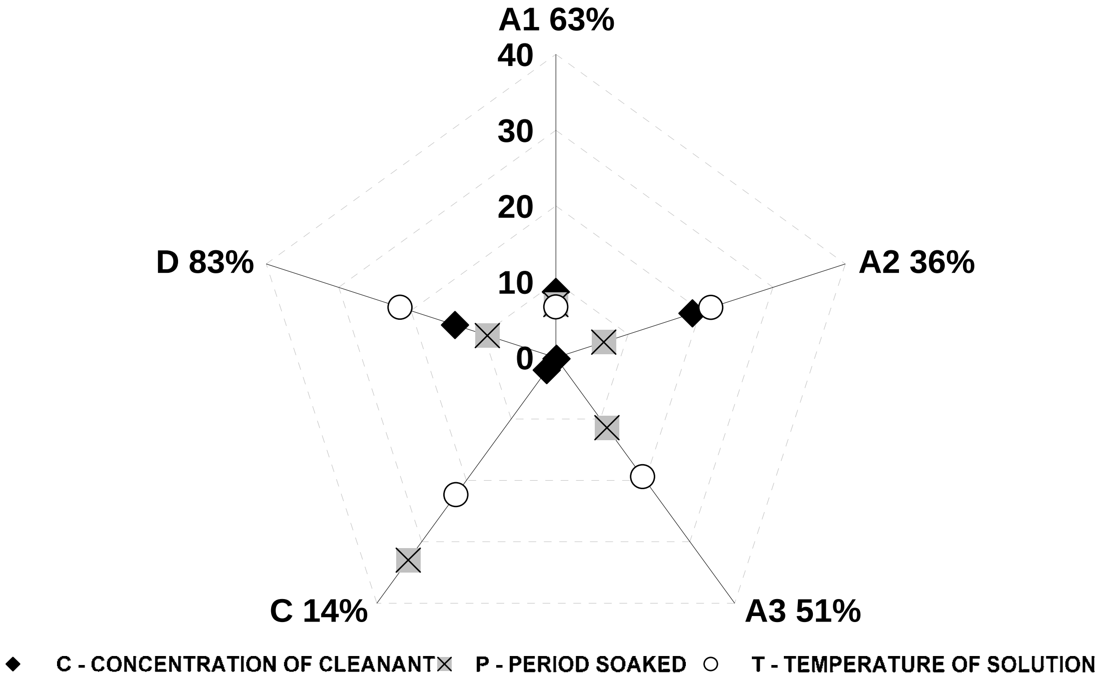

Figure 5.2 **C'T' PES ultrafiltration membranes, from plants A, C and D (with optimal recovery %).**

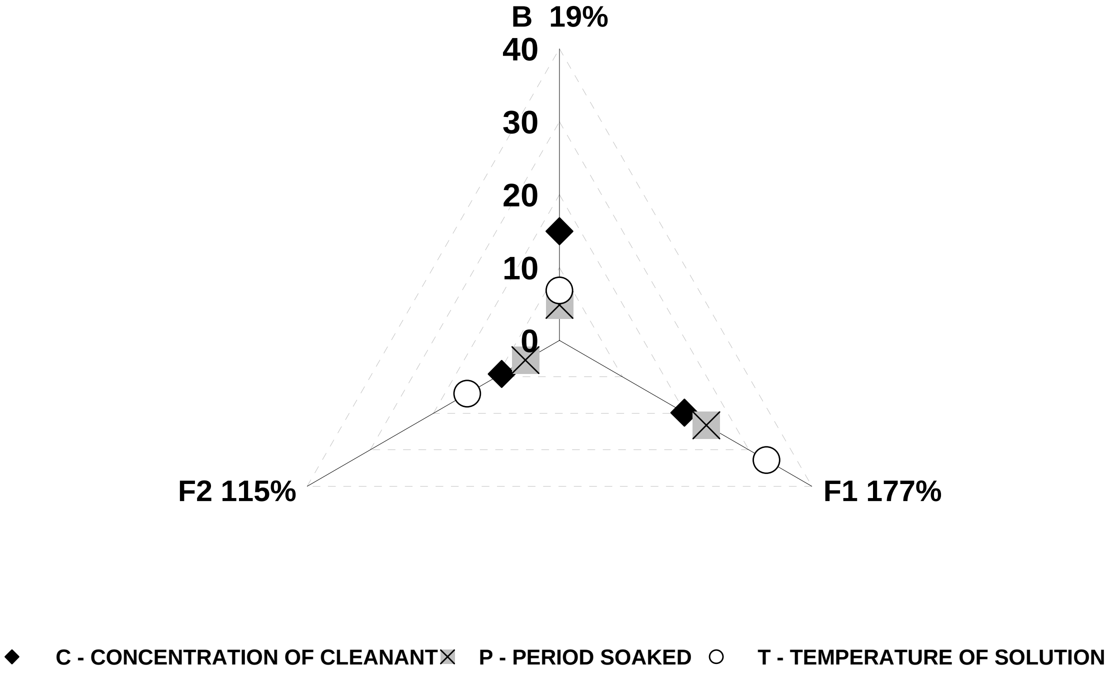

Figure 5.3 HF PVDF MF Plant B and Plant F, -effects of maturation

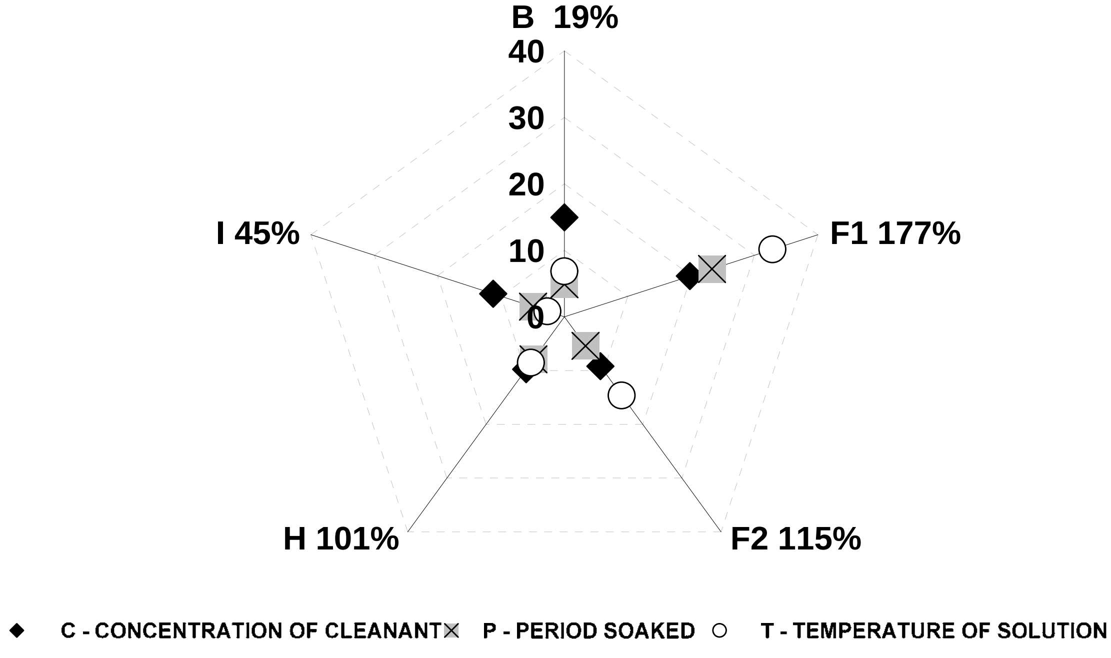

Figure 5.4 HF PVDF MF membranes from pumped system on plants B and F and submerged system on plants H and I (with optimal recovery %).

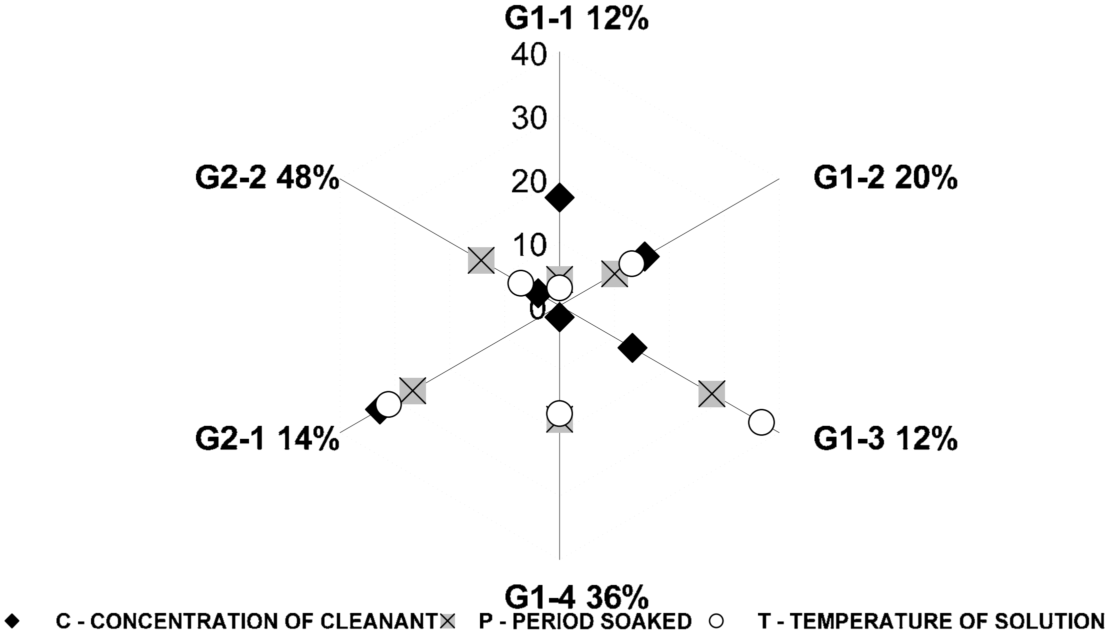

Figure 5.5 Hollow fibre PP microfiltration fibres from plant G at different lengths of time in operation (with optimal recovery %).

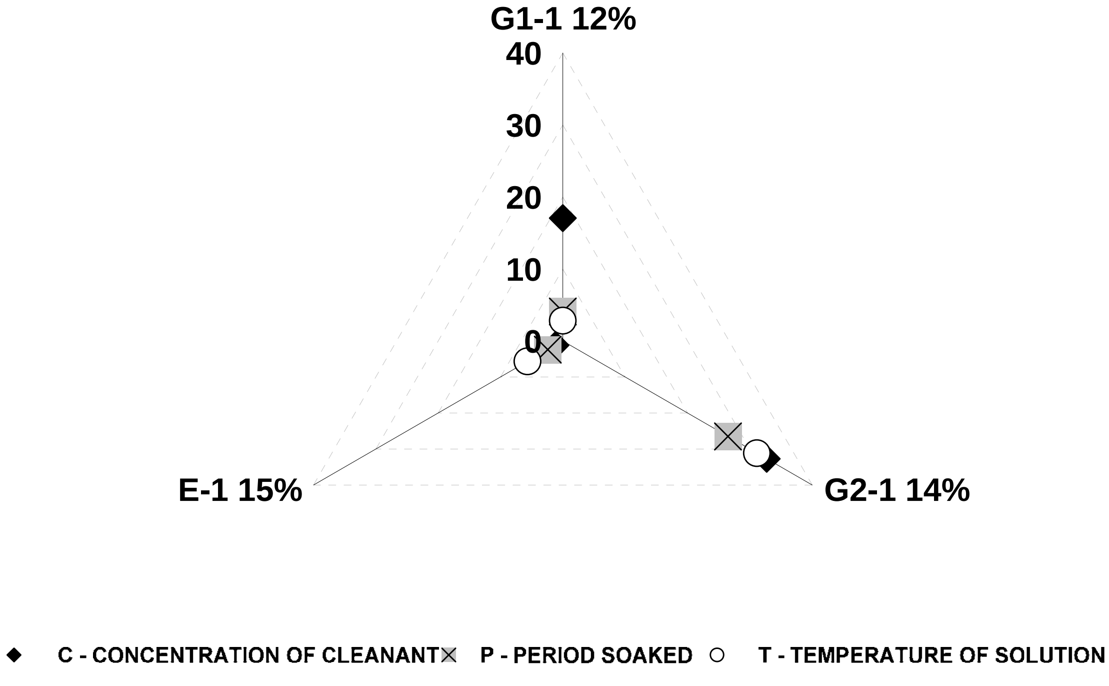

Figure 5.6 Sodium hydroxide cleaning on hollow fibre PP microfiltration membranes from plants G and E (with optimal recovery %).

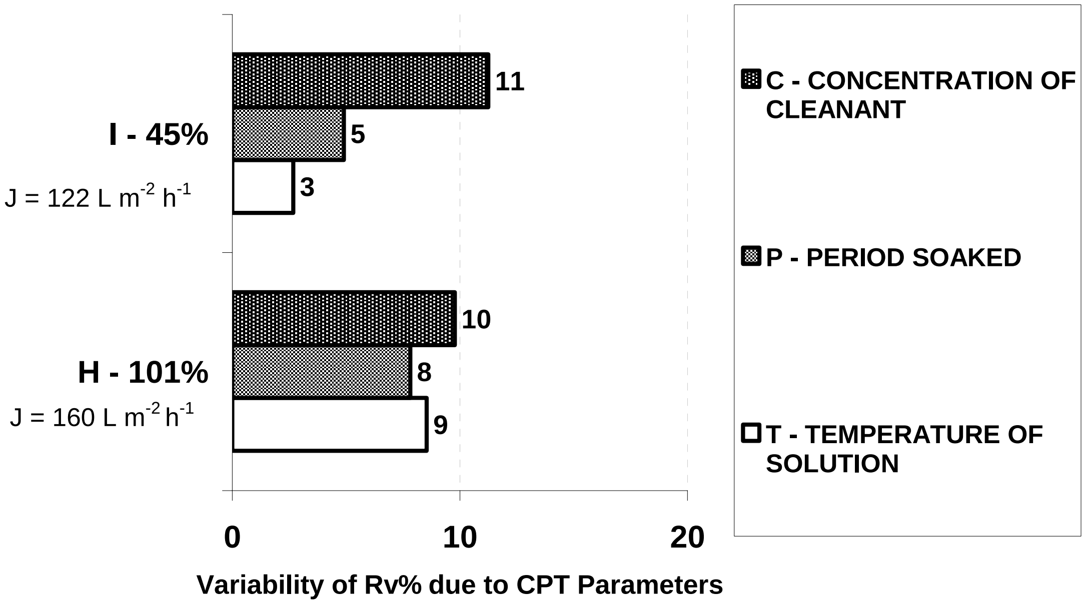

Figure 5.7 **Variability of cleaning parameters for submerged hollow fibre samples.**

5.5 DISCUSSION

The results shown in Figures 5.1 through 5.6 indicate pronounced scatter, such that no single parameter can be assumed to influence recoverability generally. This is in line with practice from cleaning with industrial foulants and fouling theories (Tragardh, 1989); where attempts to characterise cleaning from fouling relationships in dead-end conditions showed that different combinations of temperature, cleanants and regimes produced varying results (Kaiya et al., 2000). This study quantifies it.

The UF membranes from Plants A, C and D exposed to different water sources and pre-treatment all show temperature to be the most significant cleaning parameter (Figure 5.2). Where pre-treatment was absent, as in the case of Plant C, recovery was reduced and soak time became significant. This may reflect a transition in fouling mechanism; even accounting for the lower flux the source water contained untreated organics which can lead to substantial differences in fouling and thus cleaning efficacy (Lee et al., 2001; Amy 2008).

In a previous comparison of membrane materials (Yamamura, 2007) it was noted that MF PVDF fibres yielded greater flux decline rates than other membrane materials, but also that for the same cleaning parameters they recovered well with caustic soda and hypochlorite. The current study has shown, from Figures 5.3 and 5.5 and also from the heterogeneity of fouling manifested as high standard deviation values (Table 5.4), that PVDF fouling propensity can be linked with the NOM characteristic. This was reported by Kimura et al. (2006), where it was found that cleaning efficacy is also dependent on length of time in service. The effectiveness of pre-treatment affects cleaning optimisation and increases the importance of physical cleaning (Babel and Takizawa, 2000). For plant F1 and F2, cleaning with heated citric acid solutions for the later commissioning period membranes was more effective. These findings suggest studies into the changeability of the cleaning parameters have practical implications for membrane operation.

Cleanant combinations of alkaline then oxidant were found most successful for the PP MF Plant G, which is in agreement with suggested treatments for low molecular weight hydrophilic foulant removal (Strugholtz et al., 2005). The current study however shows significantly that ageing increases the importance of cleanant soak time (Figure 5.5). This could be due to longer mass transport at established foulant layers. For caustic cleaning, the efficacy over time remained consistent (12% - 14%) but the impact of concentration factor was diminished and temperature became more influential on the recovery relationship (Figure 5.6).

On a tested submerged HF PES MF plant, oxidant cleaning recovery was improved by temperature (Strugholtz et al., 2005). Results from the current study, for the more hydrophilic submerged PVDF HF MF Plants H and I, showed the contrary suggesting it is not possible to generalise from study to study and that quantification of these effects is more relevant to the development of a cleaning science than trend spotting. (Figure 5.7). For these groundwater plants the key oxidant cleanant parameter was concentration, and from the degree of reliance on concentration is seen to be related to operational flux. This further corroborates the observation that the source water and its pre-treatment affects permeability recovery and that empirical generalisation across applications is not prudent approach to developing a cleaning science.

5.6 CONCLUSIONS AND RECOMMENDATIONS

An extensive analysis of the impact of cleaning parameters on dead end MF and UF potable treatment membrane permeability recovery provided a number of observations across the plants studied. Optimum permeability recovery values from as low as 12% up to 177% were measured over the course of the four-year study, with few recognisable patterns in behaviour.

The impact and order of importance of solution temperature, concentration and cleaning stage soak time change significantly both between plants and with ageing. For example, whilst soak temperature is generally considered to have the most significant impact on cleaning, the study results indicate that ageing effects and pre-treatment can influence its impact. Even with a good knowledge of the fouling characteristics and membrane operating conditions the cleaning parameter impact factors exhibit changeability in practice, which renders the identification of the most appropriate prescriptive cleaning protocols extremely challenging. Fouling in dead end membrane systems is heterogeneous; cleaning efficacy varies not only from plant to plant but from fibre to fibre and along the fibre lengths.

It is recommended that both pilot trials and operation and maintenance of full-scale plant employ quantifiable methods for recording and tracking changes in cleaning efficacy. Cleaning parameter analysis methods, such as the one used in this study, may be used as an adjunct to classical fouling propensity studies based on chemical foulant analysis and flux step studies. As can be seen from the results of the study the impact of heating, soak time and chemical strength varies with time and operational circumstances, therefore this paper recommends such studies are necessary throughout the membranes life. Future research needs to investigate the impact of this cleaning parameter variability on useful life and operational costs.

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CHAPTER 6 CHEMICAL CLEANING OF POTABLE WATER MEMBRANES: A COST BENEFIT ANALYSIS OF CLEANING FACTOR OPTIMISATION

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6.1 INTRODUCTION

Studies into cleaning sequencing and its impact on operating costs are reliant on experimental fouling data to provide head loss information. Whilst abundant fouling data is available, as well studies of the impact and/or optimisation of physical cleaning for fouling amelioration (Lodge and Judd, 2004; Katsoufidou et al., 2005; Smith et al., 2006 van der Ven, 2008), studies of chemical cleaning of membranes in the municipal water sector are less common.

Early studies into optimisation of membrane cleaning have qualitatively modelled the relationship between cleaning regime and recovery for single foulants (Bartlett et al., 1995). These studies were developed from Hermia's blocking model, where foulants form resistance layers (Belfort et al., 1994). Further studies into quantifying the effects of chemical cleanants have been used predominantly in food and industrial applications (Shorrock and Bird, 1998; Blanpain-Avet et al., 2004). Observations of cleaning effects with surrogate foulants in laboratory experiments show differences in cleaning effects and efficiencies for different solutions (Field et al., 2008). Dead end hollow fibre membrane cleaning studies on fibres, from a single field source, showed the impact of different cleaning reagents to be dependent on foulant character (Strugholtz et al., 2005). Recently models have been developed investigating dynamic cleanant performance on membranes fouled with surface waters at high organic loads (Zondervan and Roffel, 2007). Economic simulations suggest that for ultrafiltration (UF) optimising the number of cleaning cycles does not reduce operating costs, and that cleaning should instead be optimised to control fouling (Zondervan and Roffel, 2008). It is disputed from this work that rather than the cleaning of UF and MF membranes is not relevant to operational efficiency and whole life costs, the development of measures for optimising cleaning can be used to optimise a membrane plant's performance.

Factorial analysis using analysis of variance was used to identify and optimise cleaning, with proprietary reagents, of spiral wound UF and RO membranes fouled during wastewater recovery duties (Chen et al., 2003). Recent chemical cleaning optimisation studies based on hollow fibre UF and MF (microfiltration) membranes sampled from full scale potable water treatment plants have quantified optimum permeability recovery from chemical cleaning of hollow fibre (HF) and capillary tube (CT) submerged and pumped membranes (Porcelli et al., 2009ab). The method for these latter studies was based on three factorial analyses using a response surface methodology, Box Behnken Determination (BBD), and has yielded algorithms quantifying the variation in permeability recovery from cleaning as a function of the key cleaning parameters of concentration (C'), temperature (T') and soak period (P'). The method (Porcelli et al., 2009a) has allowed the optimum values of C', P' and T' to be identified across a range of plants and cleaning protocols for hollow fibres operating under various conditions with different feed waters (Porcelli et al., 2009b).

In the following paper the results from a cost model based on the simplest representation of fouling, as resistances in series (Belfort et al., 1994; Zondervan et al., 2008), is presented based on previously published data (Porcelli et al, 2009b). The model has been applied to four full-scale, established MF/UF potable plants, selected to provide a range of membrane material types and configurations, water sources, pre-treatment, fouling conditions and corresponding operation and maintenance conditions, with the latter particularly relating to the chemical cleaning regimes.

6.2 METHODOLOGY

6.2.1 Sampled membrane plants

Cost models for a number of cleaning operational scenarios were built based on cleaning factor relationships to permeability recovery from laboratory cleaning optimisation tests.

Figure 6.1 shows the information flows to a transient TMP model built from site and laboratory data. The factorial algorithms from cleaning experiments allowed four cleaning scenarios to be explored for four potable water treatment sites (A B,C and D, see Table 6.1) with another four scenarios for variation in operational strategy, yielding 64 data in all. Operational costs over a year were calculated from energy and consumable consumption.

The HF membranes sampled from the primary stage of each of the four sites were polyethersulphone (PES) ultrafiltration; polypropylene (PP) pumped system microfiltration and polyvinyldiethylene (PVDF), with one site having a submerged configuration and the other pumped. Modules were extracted in a fouled state, with pre-backwashing and chemical cleaning, transported wet to the laboratory. Storage and autopsy of the modules and permeability testing and chemical cleaning efficacy of the extracted fibres was as described in Porcelli et al., (2009a). The complete procedure is summarised in Figure 6.1.

Table 6.1 Operational and membrane data from Sites A-D

Code	Water Type	Membrane[1]	Config.[2]	Pre-treatment
A	Groundwater: single borehole. pH = 7.1, turbidity <5 NTU	UF - PES	CT	Coagulation (PACl 4 mg L^{-1}), clarification, GAC, pre-chlorination
B	Upland surface-impounded reservoir. pH = 6.9, turbidity = 0.71 NTU	MF - PP	HF	microstrainer (30 μm)
C	Upland surface-impounded reservoir. pH = 5.5 – 5.8, turbidity < 2 NTU	MF - PVDF	HF	pH correction, pre-coagulation (PACl 2 mg L^{-1}), primary pressure filtration, pre-chlorination, 2nd stage pressure filtration
D	Groundwater: single borehole. pH = 7.1. turbidity < 0.2 NTU	MF - PVDF	sHF	Microstrainer, pre-chlorination

[1] PES = polyethyl sulphone; PVDF = polyvinylidene difluoride; PP = polypropylene (PP)
[2] Configuration: (s) HF = (submerged) hollow fibre; CT = capillary tube

6.3 GENERATION OF CLEANING PARAMETER ALGORITHM

A bespoke rig was constructed to measure permeability at constant head of ultrapure water (Porcelli et al, 2009a). Permeability was recorded before and after cleaning to allow calculation of permeability recovery (% R_v) according to:

$$\%R_v = \frac{K_f - K_i}{K_v - K_i} \cdot 100$$

Equation 6.1

where K_i and K_f are the measured initial and final permeability (L m^{-2} h^{-1} bar^{-1}) and K_v is the virgin membrane permeability.

Fouled fibres extracted from modules taken from full-scale potable water membrane plants were rinsed before assembly into bench scale modules for cleaning and recovery measurement (Porcelli et al, 2009ab). Fifteen trials were conducted in total, and %R_v measured for a range of values of cleanant concentration C’, soak period P’ and soak temperature T’ (Table 6.2). The data were used to generate site-specific algorithms from least square optimums based on a two factorial model (Porcelli et al., 2009b).

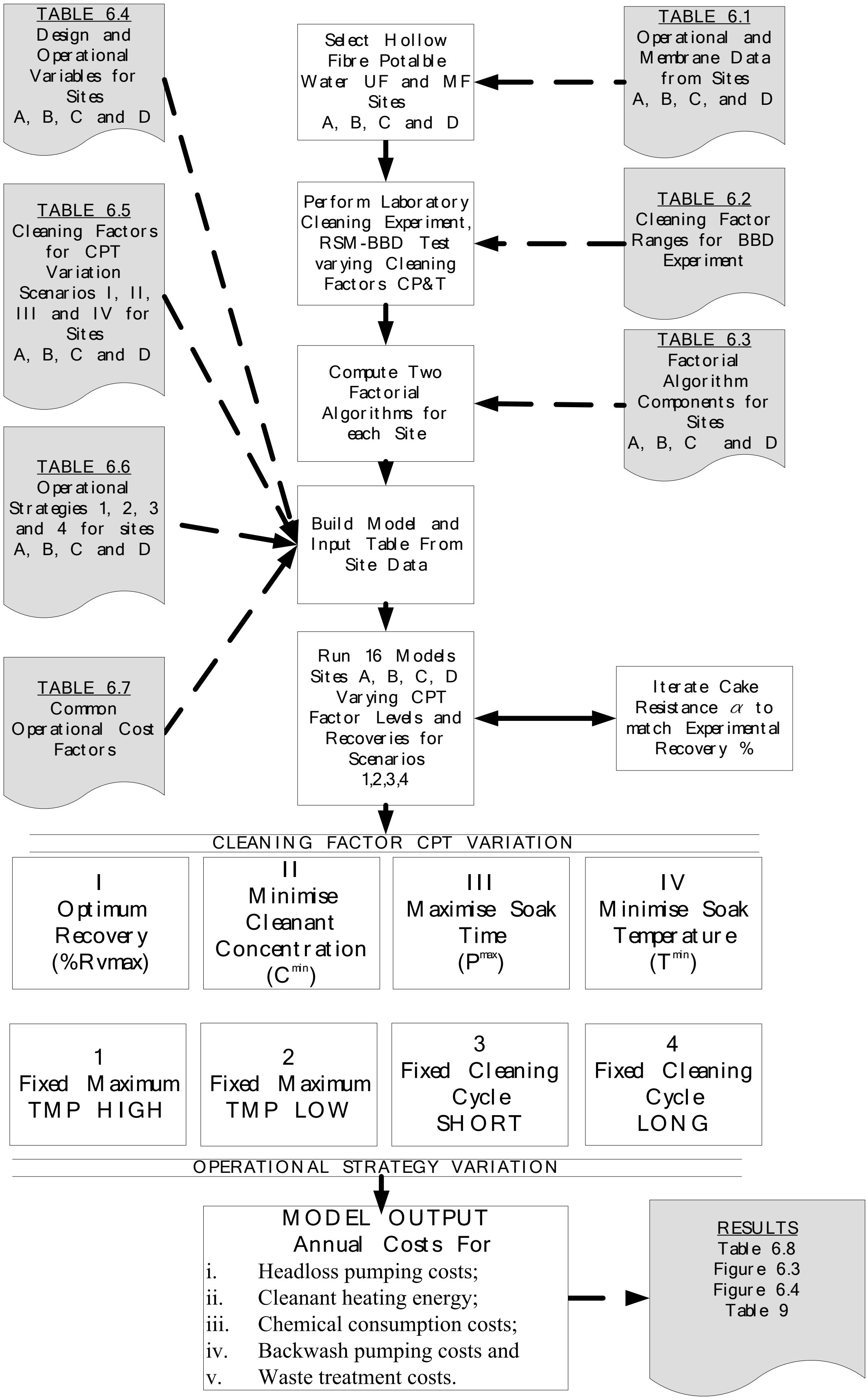

Figure 6.1 **Cleaning variation cost model methodology**

A 2^{3-3} experimental design requiring fifteen tests at varying conditions was performed with a response surface BBD matrix (Myers et al., 1989). Cleaning parameters were varied in the matrix in equal proportions, with the central points of each parameter repeated three times. The factorial multipliers (Table 6.3) from the computed responses from the CPT ranges given in Table 6.2 generated two-factorial expressions specific to each plant.

$$\%R_v = M + aC' + bP' + cT' + d(C'P') + e(C'T') + f(P'T')$$

Equation 6.2

Table 6.2 Cleaning factor ranges for BBD experiment

Site	*Cleaning agent*	*Strength (C, mol.L^{-1})*	*Soak Time (P, mins)*	*Temperature (T, °C)*
A	NaOH	0.050 – 0.175	30 – 90	10 - 40
B	NaOH	0.188 – 0.405	30 – 90	5 - 35
C	Citric Acid	0.003 – 0.009	30 – 90	5 - 35
D	NaOCl varied then H_2SO_4 at pH 2.0, 15°C,60 mins	0.001 – 0.002	30 – 90	5 - 35

Table 6.3 Factorial algorithm components for sites A, B, C and D

	M	*a*	*b*	*c*	*d*	*e*	*f*
Site							
A	34.577	-65.346	0.019	-0.156	0.379	1.716	0.000
B	9.463	-3.830	0.042	0.187	0.029	0.127	-0.005
C	125.822	-55.799	-0.950	-2.587	0.435	2.124	0.014
D	29.755	-0.078	-0.151	0.540	0.001	0.001	-0.014

6.4 COST MODEL BASIS AND OPERATION.

Using the algorithms determined from the cleaning response experiments (Table 6.3) a simple cost model was built to compute the impact of the cleaning factor envelope for each site. The model sequenced backwash and cleaning operations to generate the transmembrane pressure (TMP) transient from the Darcian relationship:

$$R_M + R_f = \frac{\Delta P}{\mu . J}$$

Equation 6.3

where R_m and R_f are the membrane and fouling resistance, ΔP the TMP, μ the viscosity of water at 20°Cand J the flux. The density of the water (ρ) in the model was also assumed to be unity. Foulant deposition was assumed to follow the model (Belfort et al., 1994):

$$R_f = \alpha(X_a + X_b + X_c)$$

Equation 6.4

where X_a, X_b and X_c are the proportions of the resistance from foulants with their a common specific cake resistance α, these being temporary where they are assumed to be completely removed by backwashing (X_a), semi-permanent where they contribute to intermediate pore blocking (X_b), and standard pore blocking where foulant build up is assumed irreversible (X_c) (Huang et al., 2009). The proportion of the intermediate fouling removed by cleaning was iterated such that the mgdel permeability recovery replicated the experimentally determined value represented in the two factorial CPT algorithms summarised in Table 6.3. By converting flux build to average headloss the annual TMP decline data preceding the removal of the membrane modules from the full-scale plants were used to estimate the average rate of foulant build up between backwashes and chemical cleans in place (CIPs) for the MF plants (B, C and D) and chemically enhanced backwashes (CEBs) for the UF plant (A).

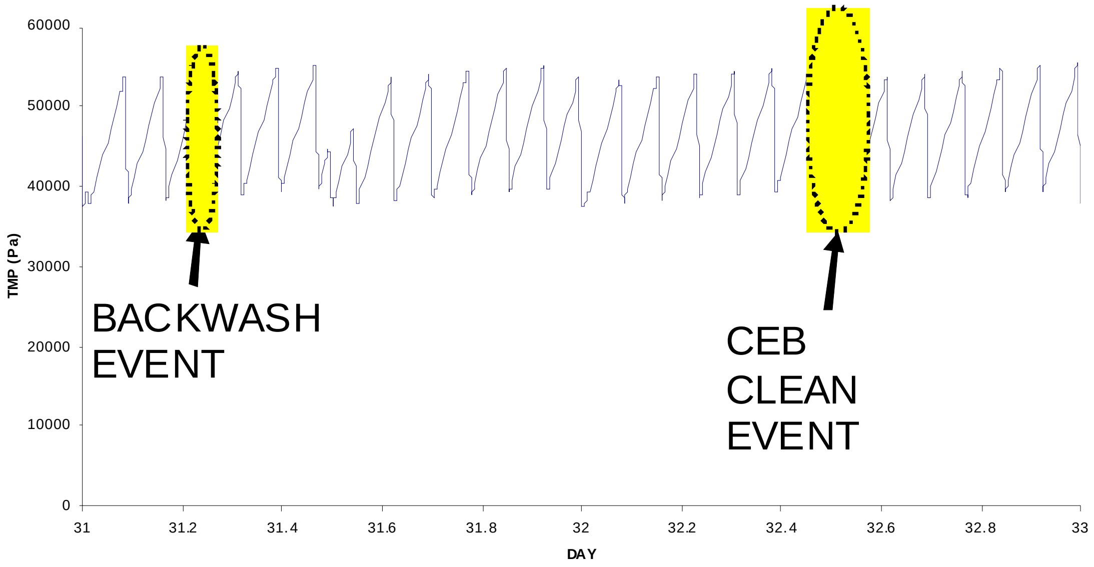

Figure 6.2 **Typical model output TMP transient, Site A, backwash and cleaning events.**

For each of the four sites the model generated the classic TMP "saw-tooth" transient (Figure 6.2 – This profile is for 2 days of operation- day 31 through 33 of 365) through appropriate scheduling of backwashing and cleaning and adjustment of R_f to match reported site conditions. Following every backwash X_a was returned to zero whereas both X_a and X_b were returned to zero following a chemical clean. Chemical cleaning was initiated in the model either at fixed time intervals or on reaching a threshold TMP. The pore fouling Xb component was iterated such that the permeability recovery (%R_v) equalled the experimentally-determined value generated from Equation 6.2 for each site (Table 6.3). The model thus provided a TMP transient over a one-year period as a function of cleaning efficacy which in turn was a function of the C', P' and T' values for which optimum values could be identified from the experimentally-derived cleaning performance algorithms.

6.5 MODEL INPUT DATA

Operational variables used to generate the TMP profiles for each site are given in Table 6.4. The annual design throughput volume (V_m) includes volumes for backwashing (V_b) and cleaning (V_c), such that net permeate production rate is given in equation 6.5:

$$V_p = V_m - (V_b + V_c)$$ **Equation 6.5**

The rate of fouling and the backwash and chemical cleaning intervals, which respectively relate inversely to the number of backwashes (N_b) and cleans (N_c) performed annually, determine the mean TMP. N_b and N_c also determine the volume of water wasted, the total energy demanded for cleaning (primarily for heating) and the chemical demand. Hence, whilst increasing the cleaning frequency between cleaning events maintains a higher average TMP, this is to some extent offset by the decreased production and cleaning energy demand.

For each plant the model parameter values (Table 6.4) were collated from site and plant design information. Fouling rates were based on data provided from plant operation. The specific cake resistance and fouling rates were determined by iteration of the variables separately to replicate the average flux decline rate as reported from site data over an annual cycle.

6.6 SCENARIOS

Four cleaning scenarios based on CPT variation were computed for each site. These represented the optimum permeability recovery, as determined from the bench-scale tests (Porcelli et al., 2009b), plus three other scenarios representing ranges of CPT variation: minimum cleanant concentration (C'_{min}); maximum soak time (Pmax) and minimum temperature (T_{min}). These were used in the algorithm (Equation 6.2) along with the design and operation data (Table 6.4) to generate the TMP transient. Results for the four cleaning scenarios are given in Table 6.5.

Each of the four cleaning scenarios was run for four operational strategy variations based on fixed high-level and low level threshold TMPs and chemical cleaning intervals. The average TMP, and thus the pumping energy demand, was thus dictated by the cleaning frequency and permeability recovery. Table 6.6 gives the operating envelopes for the four operational strategies adopted for modelling each site, for which four the CPT scenarios were applied.

Table 6.4 Design and operational variables, Sites A-D

Model Variable			A	B	C	D
Design throughput,	Q_m	$L.s^{-1}$	417	752	62.5	440
Membrane area,	A_m	m^2	17640	20092	2806	11192
Clean water resistance,	R_m	$L.m^{-1}$	2.5	2.5	1.25	2.5
Design Flux,	J	$L.m^{-2}.h^{-1}$	109	68	80	142
Backwash flux/rate	J_b /Q_b,	$L.m^{-2}.h^{-1}$	250	60	120	60
Backwash interval	t_b	minutes	55	60	60	60
Backwash duration	t_{bb}	seconds	50	20	20	20
Cleaning interval	t_c	days	*Variable*			
Cleaning duration	t_{cc}	minutes	*Variable*			
Cleanant Ratio	r_v	$m^3.m^{-2}$	1.83E-04	1.53E-03	1.53E-03	1.53E-03
Average Feed °C	T_f	°C	10.0	10.4	6.0	6.0
Cake fouling rate, as a function of (Equation 4): d(TMP) /dt,						
Cleanable cake deposits	X_a	$m.min^{-1}$	1.01.E+01	7.40.E-02	2.50.E+00	2.50.E+00
Intermediate deposits	X_b	$m.min^{-1}$	2.01.E-01	6.89.E-06	2.33.E-02	2.33.E-02
Irreversible pore fouling	X_c	$m.min^{-1}$	6.89.E-06	1.15.E-03	5.00.E-09	6.89.E-06
Specific cake resistance	α	$L.m^{-2}$	1.43.E-03	8.66.E+00	1.25.E-03	1.25.E-03

Table 6.5 Cleaning factors for CPT variation, Scenarios I-IV, for Sites A- D

			CPT VARIATION SCENARIO			
SITE A			**I** %Rv max	**II** C min	**III** P max	**IV** T min
Caustic Soda	**C**	*mol.* L^{-1}	0.175	0.050	0.175	0.175
	P	*minutes*	89	89	30	89
	T	*°C*	37	37	37	10
	RECOVERY	$\%R_v$	52.0	36.0	35.8	24.4
SITE B						
Caustic Soda,	**C**	*mol.* L^{-1}	0.405	0.188	0.405	0.405
	P	*minutes*	30	30	90	30
	T	*°C*	35	35	35	10
	RECOVERY	$\%R_v$	14.8	13.5	10.4	8.4
SITE C						
Citric Acid,	**C**	*mol.* L^{-1}	0.009	0.009	0.009	0.003
	P	*minutes*	90	30	90	90
	T	*°C*	35	35	10	35
	RECOVERY	$\%R_v$	97.0	78.0	35.0	28.0
SITE D						
Hypochlorite	**C**	*mol.* L^{-1}	0.002	0.002	0.002	0.001
	P	*minutes*	30	90	30	30
	T	*°C*	35	35	5	35
	RECOVERY	$\%R_v$	26.6	26.6	16.1	5.6

Table 6.6 Operational Strategies 1-4 for Sites A-D

		Operational strategy			
		1	2	3	4
		Fixed maximum TMP - HIGH TMP_{max}, kPa	Fixed maximum TMP - NORMAL TMP_{max}, kPa	Fixed no.cleans - SHORT t_c, days	Fixed no. cleans (t_c) - LONG t_c, days
SITE	A	50	45	0.5	1
	B	150	120	14	28
	C	70	40	14	28
	D	90	60	14	28

6.7 OPERATING COST CALCULATIONS

Determination for operational costs ($£_{op}$) for each of the sixteen scenarios (Figure 6.3) on the four sites were expressed in GBP based on the summation of the individual costs of pumping energy (£p), heating (£h), chemicals (£c) and waste (£w), using the baseline cost factors given in Table 6.7.

$$£op = £p + £h + £c + £w$$ **Equation 6.6**

Table 6.7 Baseline cost factors

Cost Factor	*Value*
Power supply efficiency (η)	0.60[1]
Pumping and heating energy (Cf_e, £/kWh)	£ 0.10[2]
Waste treatment energy (Cf_w, £/kWh)	£ 0.25
Chemical cleanant cost (Cf_c, £/tonne)[3]	Citric acid, = £ 900.13 NaOH, 50%wt = £ 60.13 Chlorine (as Cl) = £ 130

[1](Rishel , 2002) [2](Holden, 2008) [3]((IChemE, 2002), 2002 indexed to 10/2008 from UK national Statistics Office Annual Variation Tables, NSO-UK, 2008)

Pumping energy costs are derived from the throughput Q_m, the average TMP and represents the mechanical and electrical power conversion efficiency η (Table 6.7), with the total water production modified for loss of product through downtime and backflushing as given in equation 6.7:

$$£_{op} = \frac{Cf_e}{\eta}\left[TMP_{ave}\left[\left(Q_m\left(31.5x10^6 - (N_c t_{cc}) - (N_b t_{bb})\right)\right) + \left(Q_b N_b t_{bb}\right)\right]\right]$$

Equation 6.7

Heating costs are the summation of the gross energy required to heat the cleaning solutions (kWh), and is therefore a function of the difference ΔT between the ambient temperature and the reagent temperature:

$$\text{£}_h = \frac{Cf_e}{\eta}\left(\frac{C_v \Delta T (V_c \rho)}{3600}\right) \qquad \textbf{Equation 6.8}$$

where C_v is specific heat capacity (4.2 kJ. kg.K^{-1}), the density ρ is adjusted to the cleaning temperature and the volume of cleanant (V_c) is a function of the membrane area A_m and the number of chemical cleans. The cleanant volume per clean was taken from site chemical usage data from which the ratio of chemical cleanant volume to membrane area ratio was derived (r_v, m^3.m^{-2}, Table 6.4).

Chemical costs were assumed as delivered with no supplementary handling costs. The volumes used in tonnes per annum were calculated based on the volume used per CIP/CEB event over the year:

$$\text{£}_c = C_{fc}.\, C.\, N_{c.}\, V_c \qquad \textbf{equation 6.9}$$

where C_{fc} is the reported UK unit cost in £/tonne (IChemE, 2002) adjusted to 2008 values from government data. (NSO-UK, 2008).

A common notional waste disposal route was assumed based on backwashing and cleaning waste volumes which were converted to energy demand using a common waste energy cost factor C_{fw}:

$$\text{£w} = C_{fw}\,(((N_c\, V_c) + (Q_b\, N_b\, t_{bb}))\, \rho\, g\, \Delta h) \;/\; \eta \qquad \textbf{equation 6.10}$$

The cost variation in contribution from each cost group was compared for the different CPT envelope responses (Rv_{max}, C_{min}, P_{max}, T_{max}). By converting total operating costs to relative costs per cubic metre of water produced it was possible to compare costs across the different sites. The deviation from the optimum cleaning scenario for each of the operational strategy variations scenarios could also be compared.

6.8 RESULTS

Costs for each site were categorised according to headloss pumping, cleanant heating energy, chemical consumption, backwash pumping and waste treatment. These are presented as percentage of total operational costs, indicating the differences that each operational scenario provides, in Table 6.6. The operational cost of each scenario at optimal permeability recovery is shown in Figure 6.3, where the error bars indicate the minimum and maximum operational costs according to the CPT' envelope responses considered (*I* Rv_{max}, II C'_{min}, III P'_{max}, and IV T'_{max}). The % component contribution to the operating cost, based on optimum cleaning recovery, is given in (Table 6.8).

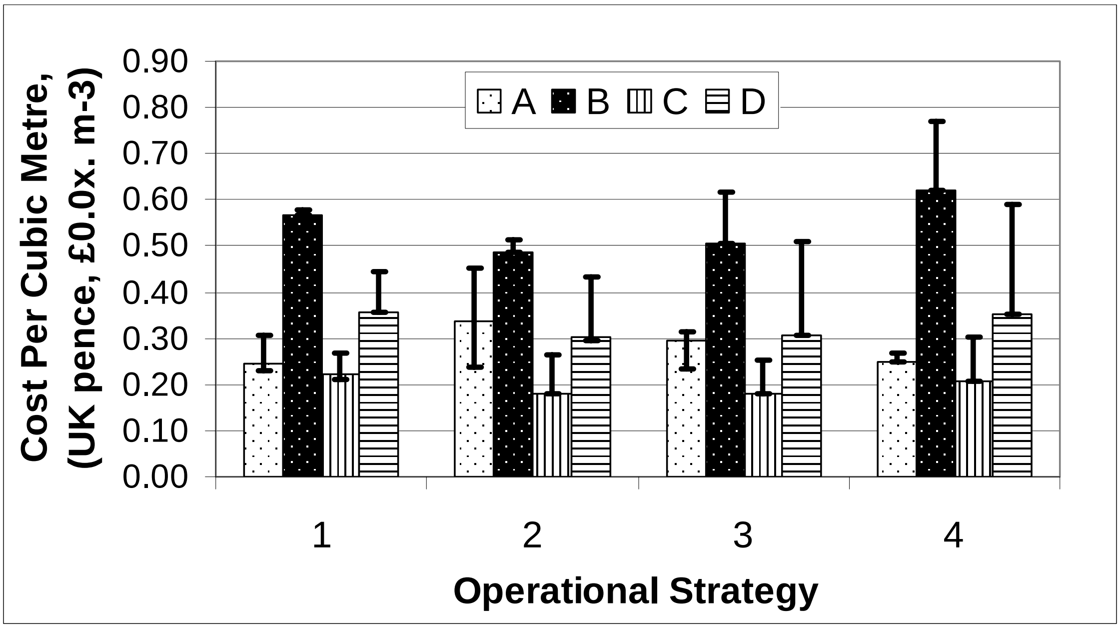

Figure 6.3 Optimal cleaning recovery and relative of water with strategy variation.

It can be seen that the cost associated with the TMP headloss represents the primary component in all cases and that, despite optimisation of cleaning, there is significant variation in calculated operational costs with chosen operational strategies.

The influence of the cleaning factor envelope variations on the unit cost is shown in Figure 6.4, which shows the relative cost variation across the ranges of recovery attained from the four cleaning scenarios applied to the four operational strategies chosen for the four sites. For all the models the range of variation from treatment factors was found to have cost implications that varied with operational strategy. The MF membrane sites indicate increasing costs under optimal cleaning conditions as the operational strategy proceeds from a high fixed TMP (1) to a long fixed interval cleaning cycle (4). For Site A, the UF plant, the wider range of costs for a fixed low TMP strategy indicates that frequency of cleaning operations together with the optimisation of the cleaning factors has the largest impact on operational costs. Calculation of the variation in the plant life operating costs for each of the scenarios (Table 6.8) indicate cost savings from cleaning optimisation to vary with the plant operational strategy. Table 6.9 includes data for cash flow using a discount rate of 10% and a 15 year plant life.

Table 6.8 Contribution to OPEX costs from optimum cleaning protocol (I)

SCENARIO	***SITE***			
	A	B	C	D
Scenario 1 Fixed Maximum TMP- HIGH				
Headloss Pumping Costs	75.2	97.2	89.5	96.4
Cleanant Heating Energy	18.2	1.9	6.1	2.6
Chemical Consumption Costs	0.3	0.1	1.8	0.1
Backwash Pumping Costs	1.8	0.2	0.7	0.2
Waste Treatment Energy Costs	4.5	0.6	1.9	0.6
Scenario 2 - Fixed Maximum TMP – LOW				
Headloss Pumping Costs	49.7	95.1	72.9	91.6
Cleanant Heating Energy	44.9	3.9	19.4	7.2
Chemical Consumption Costs	0.7	0.2	5.6	0.4
Backwash Pumping Costs	1.3	0.2	0.6	0.2
Waste Treatment Energy Costs	3.4	0.6	1.5	0.6
Scenario 3 Fixed Number of Cleans- SHORT (t_c)				
Headloss Pumping Costs	57.7	95.8	74.7	92.9
Cleanant Heating Energy	36.5	3.2	17.9	6.0
Chemical Consumption Costs	0.6	0.1	5.2	0.3
Backwash Pumping Costs	1.5	0.2	0.6	0.2
Waste Treatment Energy Costs	3.7	0.6	1.6	0.6
Scenario 4 Fixed Number of Cleans - LONG (t_c)				
Headloss Pumping Costs	73.2	97.8	87.5	96.5
Cleanant Heating Energy	20.3	1.3	7.7	2.6
Chemical Consumption Costs	0.3	0.1	2.2	0.1
Backwash Pumping Costs	1.8	0.2	0.7	0.2
Waste Treatment Energy Costs	4.4	0.6	1.8	0.6

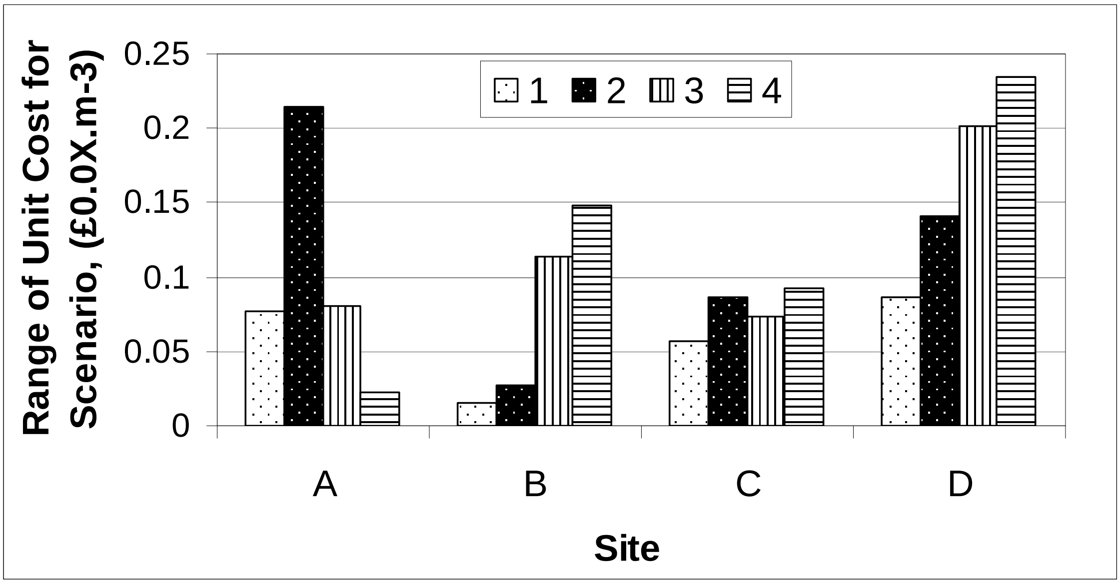

Figure 6.4 **Relative operational costs with cleaning factor envelopes for operational strategy on Sites A-D.**

Table 6.9 **Difference between best and worst case operational costs.**

	A		B		C		D	
	i	*ii*	*i*	*ii*	*i*	*ii*	*i*	*ii*
1	7,532	*55,487*	3,531	*26,015*	1,103	*8,123*	11,865	*87,409*
2	23,386	*172,274*	6,347	*46,754*	1,664	*12,256*	19,402	*142,931*
3	9,254	*68,170*	26,838	*197,710*	1,412	*10,403*	27,796	*204,766*
4	2,690	*19,817*	34,897	*257,075*	1,833	*13,503*	32,423	*238,850*

i Saving in £ per annum between worst and best case operational cost with cleaning factor CPT variation.
ii Annual operational savings as capital amortised for 15 yrs at 10%

6.9 DISCUSSION

Evidence provided from the analysis (Figures. 6.3 and 6.4, Table 6.9) indicate that there is a significant impact of both the plant operating protocol (i.e. the basis chosen for applying the chemical clean) and the degree of optimisation of the chemical clean (i.e. the attainable permeability recovery and energy/consumables expenditure) on overall operating costs. For example, for Plant C the spread of the costs arising from operation across the range of the cleaning envelope (i.e. range of CPT' values) can be as high as 0.2 pm^{-3} above an optimum of 0.3 pm^{-3} for a strategy of less frequent CIPs (Figure 6.3). If the operating strategy is changed to more frequent CIPs the operating costs decrease by around 10% and its cost variation for operation across the CPT' envelope is reduced by a third. As the scenario is changed, the impact of the cleaning envelope is also changed, with the largest impact - from 0.08 to 0.23 pm^{-3} - measured for Plant D (a submerged MF plant) on changing from operation at fixed high threshold TMP to operation at less frequent CIPs (Figure 6.4). Moreover, there is a distinct difference in the pattern of behaviour between the UF plant (Plant A) and the microfiltration plants (Plants B to D), according to Figure 6.4.

It has previously been observed (Zondervan and Roffel, 2008), that optimisation of cleaning cycles has little impact on the lifetime cost (based on models built for a UF plant scenario with varied cycle time for a particular set of parameter data). This study, conducted across a range of plants, shows that the cleaning response is variable. Annual operational costs relating primarily to energy and consumables can vary from below £20k to over £170k for an ultrafiltration plant (Plant A) challenged with pre-treated groundwater (seen in Table 6.4).

This has implications regarding existing and proposed control strategies for backflush frequency and cleaning cycle control, such as those based on neural networks based on cleaning efficacy attained on site (Veerapaneni et al., 2004, Oh et al., 2004). Such approaches using heuristic data for feedback control are necessarily constrained by the operating envelope used on site, whereas the approach used here allows a wider envelope to be explored and optimum cleaning conditions to be precisely identified.

6.10 CONCLUSIONS

The effects of actual variable cleaning recoveries for different membrane fibres fouled on full scale plants were modelled with a simple fouling model to generate a transient TMP data with variable backwash and cleaning cycles. The effect of cleaning factor variations on recovery from experimental data was incorporated into the model along with plant operating strategies, and the model used to quantify their impact on operating cost.

The study revealed:

1. Differences in cleaning factor performance, as determined by the values for the cleanant concentration C' and cleaning temperature T' and soak period P', were significant and have a bearing on the design of a process, and in particular the range of fluxes which determine the fouling rate and so cleaning cycle times.

2. Optimisation of permeability recovery has a measurable impact on operational costs, with the difference in annual operating cost between the most and least optimal chemical clean being as little as £5 to as high as £74 p.a. per L. s^{-1} of flow depending upon the plant operating protocol.

3. The extent of the operational cost reduction for the optimal cleaning regime is dependent on both the operational strategy, with respect to the basis for scheduling the chemical cleans, and the membrane type (ultrafiltration vs. microfiltration) and/or configuration (hollow fibre vs. capillary tube).

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CHAPTER 7 MODELLING HOLLOW FIBRE FLOW AND HEADLOSS DISTRIBUTION EFFECTS.

Modelling Hollow Fibre Flow and Headloss Distribution Effects. Submission to *Water Research* in preparation.

7.1 INTRODUCTION

Early attempts to describe completely the flow in hollow fibre membranes used Navier –Stokes equation for the flow along the lumen with Darcy's law being used to describe the flow through a thin walled isotropic porous media (Apelblat et al., 1974). The strong influence of the internal diameter of hollow fibres on flow for dead end operation is well known (Doshi et al., 1977). Models have been proposed for practical applications that relate flow to internal diameter, void fraction, coefficient of hydraulic permeability, length and internal/external diameter ratio. These early hollow fibre-based models were for short run, maximum mass transfer/minimum membrane area applications and were unsuitable for municipal applications (Wickramasinghe et al., 1991). The expansion of hollow fibre modules in the water industry (Rachwal and Judd, 2006) has led to more applicable model construction; Serra et al., (1998) developed an early practical axial model of fouling in dead end hollow fibres that could be used for a number of scenarios, concluding that maximum net flow rate was a combination of packing density and fibre diameter and that geometry affected the optimum design for feed conditions.

Fouling and cleaning effectiveness are thus also limited by membrane geometry. A hypothetical model using discrete boundary conditions for the axial flux profile was proposed by Carroll and Booker (2000). The model is derived Darcy/Hagan Poiseulle and predicts flow through various lengths of a fibre. Three distinct operating conditions were identified characterised by the geometry of the fibre. There is initially a decline in the permeate flowrate which is related to the non-uniform axial flux profile. In the first fouling stage the decline in permeate flowrate is independent of fibre length and surface area. Deposition is highest where the flux is at a maximum and rapid change in flux is a characteristic of Hollow Fibre (HF) membranes with deposition greatest near the pots. In the intermediary stage the rate of flux decline is related to fibre length. As the ends foul, the pressure profile changes and the flux increases towards region furthest from the inlet. The membrane surface begins to be used more effectively as time progresses. After substantial fouling, the last stage is reached when the fouling and TMP distribution along the fibre is uniform. In this last stage the axial geometry of the membrane is less important and throughput with uniform foulant deposition continues so long as driving head exists.

These transition effects are influenced by membrane geometry whether in-out CT or out-in HF. Deposition characteristics should be uniform in theory but this axial deposition model suggests there is relationship between the flux, driving head and membrane geometry that determines the pattern of foulants and thus influences the cleaning effectiveness. More recently work has looked more closely at the effect of these axial fouling stages on advanced cake filtration theory, incorporating effects of particle zeta potential and applied flux. (Chang et al., 2006; Kin-ìl et al., 2009); the uniform flux stage is approached more quickly with increasing charge, smaller particle size and higher imposed flux.

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To examine the effect of membrane design on cleanability, a study has been conducted to practically illustrate the axial relationship of flow to TMP using a mathematical model which calculates the headloss for discrete segments along a hollow fibre length for a given flow. The model is easily adaptable to different membrane internal diameters and lengths and can be related to experimental data for flow conductivity. Having verified the Hagen-Poiseulle model mathematically, this work develops and compiles it into a useable BASIC programme (Appendix D) that was verified using an analogue foulant. It was utilised and developed to examine further the geometric relationships for the different membranes being cleaned in the rest of the work

7.2 MATHEMATICAL MODEL BASIS

For a virgin membrane with DI water, flow may be considered as steady flow non-uniform flux, since the hydraulic conductivity coefficient is kept constant along the fibre. For this case, two well known formulae were used to mathematically simulate the hydraulic flow of the hollow fibre membrane. Darcy’s Law was used to simulate the flow through the fibre wall and the Hagen-Poiseulle’s Law to simulate the laminar flow through the fibre channel. The mathematical model was calibrated and verified then used to study the hydraulic performance of hollow fibre membranes, specifically the *M10* Hollow Fibre Membrane Module produced by Memcor Company and the hollow fibre membrane of *X-Norit S-225 FSFC*.

In actual conditions the hydraulic conductivity coefficient of the fibre is not constant along the length; it is varied along the fibre length. Its variation depends on the water quality and the flowrate distribution along the fibre length. The higher material rejection on the fibre surface, such as following rainy periods, the higher the reduction in the hydraulic conductivity coefficient. In this case the flow is considered as unsteady non-uniform flow.

7.3 HYDRAULIC FLOW SIMULATION OF HOLLOW FIBRE MEMBRANE UNDER CONSTANT HYDRAULIC CONDUCTIVITY

The differential form of Darcy's Law for a linear flow through a porous media is:

$$q = AK\frac{dP}{dx}$$ **Equation 7.1**

where:

q = volumetric flowrate, $L^3\ T^{-1}$,

A= cross sectional area of the porous media, L^2,

K= hydraulic conductivity of the porous media, $L\ T^{-1}$, and

$\frac{dP}{dx}$ = hydraulic gradient, $L\ L^{-1}$.

The hydraulic conductivity, K, is a measure of the ability of water to flow through a porous medium. It depends on the fluid properties and the porous medium properties through the intrinsic permeability. For water engineering purposes we can assume that for Bernoulli's modified headloss equation where the viscosity is the same and density is unity the TMP can be expressed as meters.

The flow through the hollow fibre membrane is a radial flow (Figure 7.1a). An expression for the Darcy's law for radial flow can be derived from Eq. (7.1) by transforming its Cartesian coordinate system into polar coordinates, that is:

$$q = 2\pi r \Delta S K \frac{dP}{dr}$$ **Equation7.2**

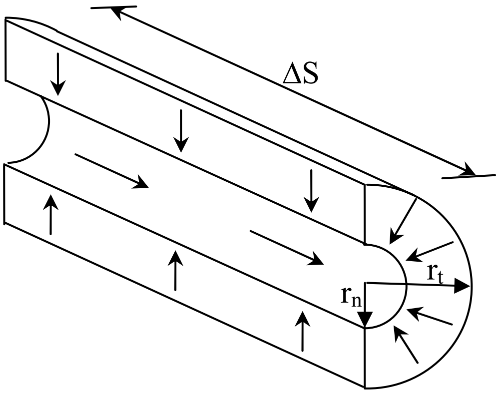
Rearranging and integrating,

$$\int_{r_n}^{r_t} \frac{q dr}{r} = 2\pi \Delta S K \int_{p_n}^{p_t} dp`$$ **Equation 7.3**

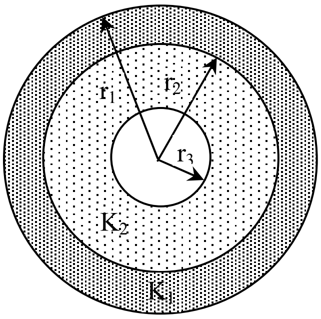
$$q = 2\pi \Delta S K \frac{P_t - p_n}{\ln(\frac{r_t}{r_n})} = 2\pi \Delta S K \frac{TMP}{\ln(\frac{r_t}{r_n})}$$ **Equation 7.4**

where

TMP	=	Transmembrane pressure ($P_t - P_n$), L.
r_t	=	the fibre outer radius, L, and
r_n	=	fibre inner diameter, L.



a- Longitudinal cross section in hollow fibre membrane



b- Cross section through a hollow fibre membrane wall

Figure 7.1 **Schematic diagrams of different cross sections in a hollow fibre membrane.**

For a hollow membrane with multi layers, an equivalent hydraulic conductivity can be substituted for the conductivity of the layers. Consider the hollow fibre membrane with two layers shown in Figure 7.1b, the radial flowrate, q, is the same for each layer and the total transmembrane pressure; equal to the sum of the transmembrane pressure of each layer, which is:

$$TMP = TMP_1 + TMP_2 \qquad \textbf{Equation 7.5}$$

Substituting an expression for TMP from Eq.(7.4) yields:

$$2\pi\Delta S\frac{q\ln(\frac{r_1}{r_3})}{K_{eq}} = 2\pi\Delta S\frac{q\ln(\frac{r_1}{r_2})}{K_1} + 2\pi\Delta S\frac{q\ln(\frac{r_2}{r_3})}{K_2} \qquad \textbf{Equation 7.6}$$

Then equivalent hydraulic conductivity is then:

$$K_{eq} = \frac{\ln(\frac{r_1}{r_3})}{\frac{\ln(\frac{r_1}{r_2})}{K_1} + \frac{\ln(\frac{r_2}{r_3})}{K_2}}$$ **Equation 7.7**

7.4 HEAD LOSSES THROUGH HOLLOW FIBRE MEMBRANE CHANNEL:

The energy loss converted to head loss along the membrane channel can be expressed in engineering terms fo TMP or headloss based on Poiseulle's Law for a laminar flow that is:

$$h_l = \frac{8Q\,\mu\,L}{\pi\,r_n^{\;4}}$$ **Equation7. 8**

in which:

h_l = the head loss, L (Energy = Kg. $m^{-1}.s^{-2}$)

Q = total discharge of the channel, L^3 T^{-1}, and

μ = the water viscosity, L.T.

7.5 APPLICATION OF DARCY'S AND POISEULLE'S LAWS TO A HOLLOW FIBRE MEMBRANE

Figure 7.2, shows a schematic diagram of a fibre membrane inserted in an actual fibre module. The pot distance, L_{pot} is required to seal the fibres so that all the flowrate of the module will be through the fibre channels of effective fibre length L. The fibre has openings at both ends and ideally the flow of the fibre membrane module is symmetrical. Therefore the flowrate calculations are carried out on one half and the whole flow for the fibre is then twice that calculated.

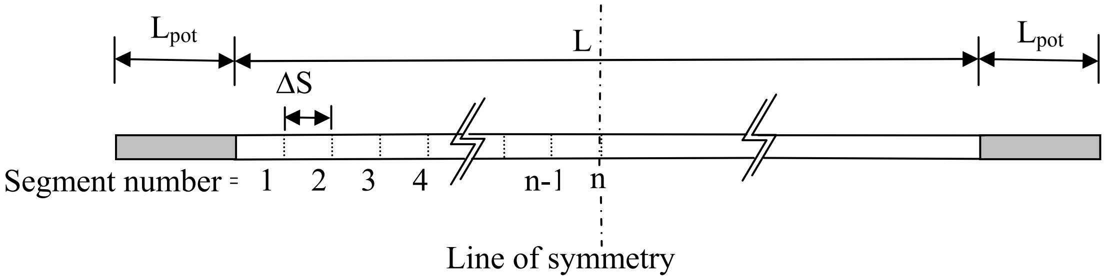


Figure 7.2 **A schematic diagram of a fibre membrane in an actual fibre membrane module.**

By dividing the fibre membrane length under consideration into n equal segments of a length equal to ΔS, then Eq. (7.4) is used to calculate the flowrate through segment:

$$q_1 = 2\pi \Delta S\, K \frac{TMP_1}{\ln(\frac{r_t}{r_n})}$$ **Equation 7.9**

The transmembrane pressure head, TMP (ΔP m), for a hollow fibre membrane is defined in this study as the head available to drive water from a point along the fibre to the outlet. By assuming that the transmembrane pressure, TMP, remains constant throughout each segment and equal to the TMP at the beginning of the segment, the TMP throughout 1st segment may be written as:

$$TMP_1 = h_{ap} - hl_{pot}$$ **Equation 7.10**

in which:

h_{ap} = the applied head, L, and

hl_{pot} = the head loss through the pot length, L.

The head loss throughout the fibre channel along the pot length may be calculated by applying Poiseulle's Law, Eq. (7.8), that is

$$hl_{pot} = \frac{8\,(q_1 + q_2 + ... + q_n)\,\mu\, L_{pot}}{\pi\, r_i^4}$$ **Equation 7.11**

Where L_{pot} is the pot length, L.

The expression for flowrate through 1st segment, Eq. (7.9), may be written as:

$$q_1 = \frac{2\pi \Delta S\, K\, (h_{ap} - \frac{8(q_1 + q_2 + ... + q_n)\mu L_{pot}}{\pi r_i^4})}{\ln(\frac{r_t}{r_n})}$$ **Equation 7.12**

Rearranging and rewriting

$$(C_2 + 1)q_1 + C_2(q_2 + q_3 + ... + q_n) - C_1 = 0$$ **Equation 7.13**

in which

$$C_1 = 2\pi \Delta S K \frac{h_{ap}}{\ln(\frac{r_t}{r_n})}$$ **Equation 7.14**

and

$$C_2 = 16 \Delta S K L_{pot} \frac{\mu}{\ln(\frac{r_t}{r_n})\, r_i^4}$$ **Equation 7.15**

In general, Eq. (7.13) may be written as:

$$(C_2 + 1)q_1 + C_2 \sum_{i=2}^{n} q_i - C_1 = 0$$ **Equation 7.16**

A similar expression can be obtained for the 2nd segment. The TMP along the segment can be written as:

$$TMP_1 = h_{ap} - hl_{pot} - hl_1$$ **Equation 7.17**

In which hl_1 is the head loss along the 1st segment, which may be obtained by using Poiseulle's Law, Eq. (7.8), that is

$$hl_1 = \frac{8(0.5q_1 + q_2 + ... + q_n)\mu \Delta S}{\pi r_i^4}$$ **Equation 7.18**

When calculating the head loss through segment 1, the flowrate through it is not fully developed, varying from 0 to q_1. It was assumed then that the flowrate through the walls of this segment varies linearly along the segment and the average was taken to calculate the head loss. The expression for flowrate through segment 1 may then be written as:

$$q_2 = -\frac{2\pi\Delta S K\,(h_{ap} - \frac{8(q_1 + q_2 + ... + q_n)\mu L_{pot}}{\pi r_i^4} - \frac{8(0.5q_1 + q_2 + q_3 + ... + q_n)\mu\Delta S}{\pi r_i^4})}{\ln(\frac{r_t}{r_n})}$$

Equation 7.19

By defining

$$C_3 = 16\Delta S^2 K \frac{\mu}{\ln(\frac{r_t}{r_n})\,r^4}$$

Equation 7.20

and arranging and rewriting

$$(C_2 + 0.5)q_1 + (C_2 + C_3 + 1)q_2 + (C_2 + C_3)(q_3 + q_4 .... + q_n) + -C_1 = 0$$

Equation 7.21

or

$$(C_2 + 0.5)q_1 + (C_2 + C3 + 1)q_2 + (C_2 + C_3)\sum_{j=3}^{n} q_j - C_1 = 0$$

Equation 7.22

A similar equation may be obtained for the remaining segments, which may be written in general form as:

$$\sum_{j=1}^{i-1}(C_2 + ((j-1) + 0.5)C_3))q_j + (C_2 + (i-1)C3 + 1)q_i + (C_2 + (i-1)C_3))\sum_{j=i+1}^{n} q_j - C_1 = 0$$

Equation 7.23

In which i is the segment number.

By applying Eq. (7.14) to the 1st segment and Eq. (7.22) to the 2nd segment through to the nth segment, the resultant is a system of linear equations with *n* unknowns represents the flowrate throughout each segment, the system may be written as shown in Table (7.1):

Table 7.1 System of Linear Equations for Segmental Flowrates.

Equation no.	q_1	q_2	q_3	q_4	------q_n	***R.H.S***
1	C_2+1	C_2	C_2	C_2	------C_2	C_1
2	$C_2+0.5C_3$	C_2+C_3+1	C_2+C_3	C_2+C_3	----C_2+C_3	C_1
3	$C_2+0.5C_3$	$C_2+1.5C_3$	C_2+2C_3+1	C_2+2C_3	---C_2+2C_3	C_1
4	$C_2+0.5C_3$	$C_2+1.5C_3$	$C_2+2.5C_3$	C_2+3C_3+1	---C_2+3C_3	C_1
⋮	⋮	⋮	⋮	⋮	⋮	⋮
n	$C_2+0.5C_3$	$C_2+1.5C_3$	$C_2+2.5C_3$	$C_2+3.5C_3$	---$C2+(n-1)C_3+1$	C_1

The above system of n simultaneous equations can be solved using any method for solving linear equations. Having obtaining the values of the flowrate of each segment, the transmembrane pressure can be calculated for each segment.

7.6 MATHEMATICAL MODEL VERIFICATION

The developed mathematical model was calibrated and verified by laboratory experimental data. Laboratory experiments were carried out on four identical sets of

HF membrane. Data for the membrane is presented in Table (7.2). Each set of fibre was inserted in a carrier pot of length 4 cm and throughput measured for a standard time with application of a constant head of DI water.

Table 7.2 Hollow Fibre Membrane Data Set

Type	*Inner diameter (mm)*	*Outer diameter (mm)*	*Number of fibres used*
PP S10V	0.39	0.65	10

The flowrate for each set was measured with its initial length and repeated for a number of segment lengths as shown in Table 7.3. The initial step toward verification was the calculation of a conductivity coefficient. The value of the conductivity coefficient is calculated using the data from the first set with an initial length of 50 cm only. This conductivity coefficient was adjusted until the predicted flowrate value matched the experimental values. The general conductivity coefficient value was found to be $4.64\text{x}10^{-7}$ cm s^{-1}.

The mathematical model was then used to generate the flowrate values of the fibre sets by changing the fibre length. Figure 7.3 shows the measured and mathematical model predicted flowrate value. A good agreement between the measured and predicted flow rates values can be noticed with a correlation coefficient of 0.996.

Table 7.3 **Measured flowrate values for fibre sets.**

Set no.	*Length (cm)*	*Measured flowrate ml min^{-1}*
1	50	18.1
	40	17.02
	30	14.84
	20	11.2
	10	5.9
2	45	18.1
	30	15.23
	15	8
3	50	18.6
	40	16.82
	30	14.84
	20	11.89
	10	6.48
4	45	18.14
	30	14.41
	15	7.84

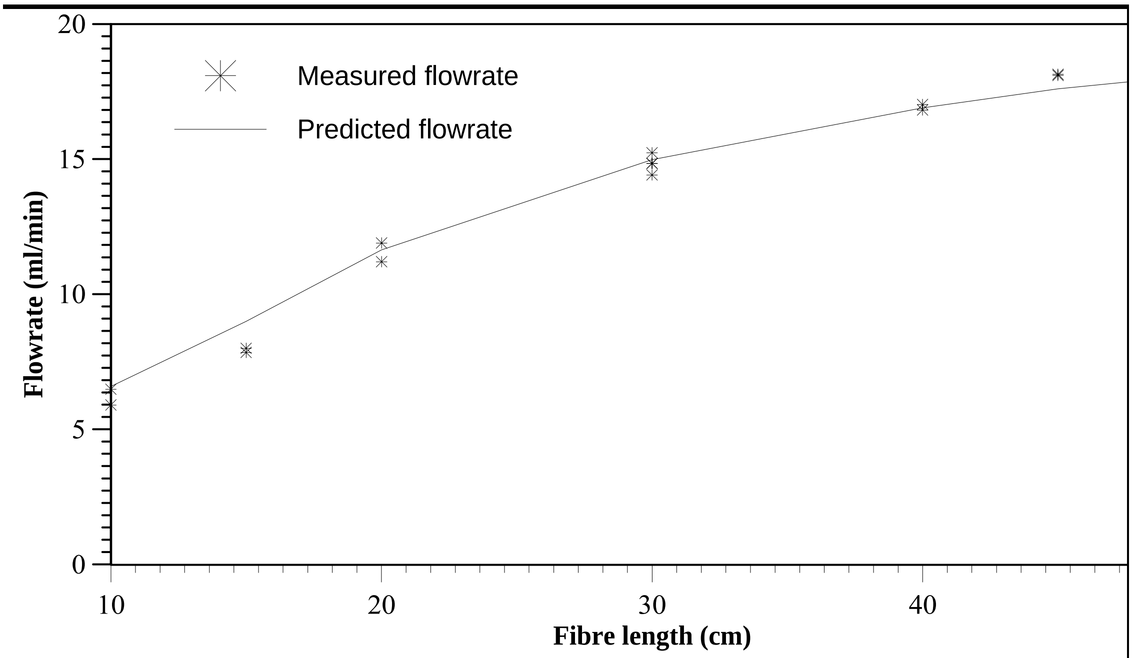


Figure 7.3 **Comparison between measured and model predicted flow rate**

7.7 APPLICATION OF THE MATHEMATICAL MODEL

The developed and verified mathematical model was used to study the hydraulic performance of commercial hollow fibre membrane.

Two sets of fibres extracted from full scale modules supplied were used: a PP microfiltration fibre from a Memcor M10 module taken from Cornhow WTW (Plant G in Appendices) and a PES capillary tube ultrafiltration membrane from Norit X-Flow module removed from N. Mymms WTW (Plant A in Appendices). Using the membrane specifications, the model was ran to generate a headloss profile based on the segmentation from the mathematical model.

7.8 MATHEMATICAL MODEL APPLICATION: MF-HF

The M10 module specifications are listed in Table 7.4.

Table 7.4 Microfiltration hollow fibre membrane module specification

Item	
Fibre inner diameter	0.25 mm
Fibre outer diameter	0.55 mm
Number of fibres per module	20 000
Effective fibre length	97 cm
Pot length	10 cm
Total effective area	33.5 m^2
Normal Module operation permeability	120 to 240 $L.m^{-2}.h^{-1}.bar^{-1}$

Source, Memcor, 2005

The virgin membrane permeability, measured with ultrapure DI water, was approximately 2,000 $L.m^{-2}.h^{-1}.bar^{-1}$ at 20°C. This permeability can be replicated in the mathematical model using a hydraulic conductivity coefficient of 2.32×10^{-5} m s^{-1}. The flowrate through each computational segment a long the fibre length, as a percentage of the total fibre flowrate, is shown in Figure 7.4. The segment length is 1 cm for all calculations. It can be seen that more that 99.3% of the throughput flowrate is concentrated in the first 10 cm of the fibre length from the potted end irrespective of overall flux.

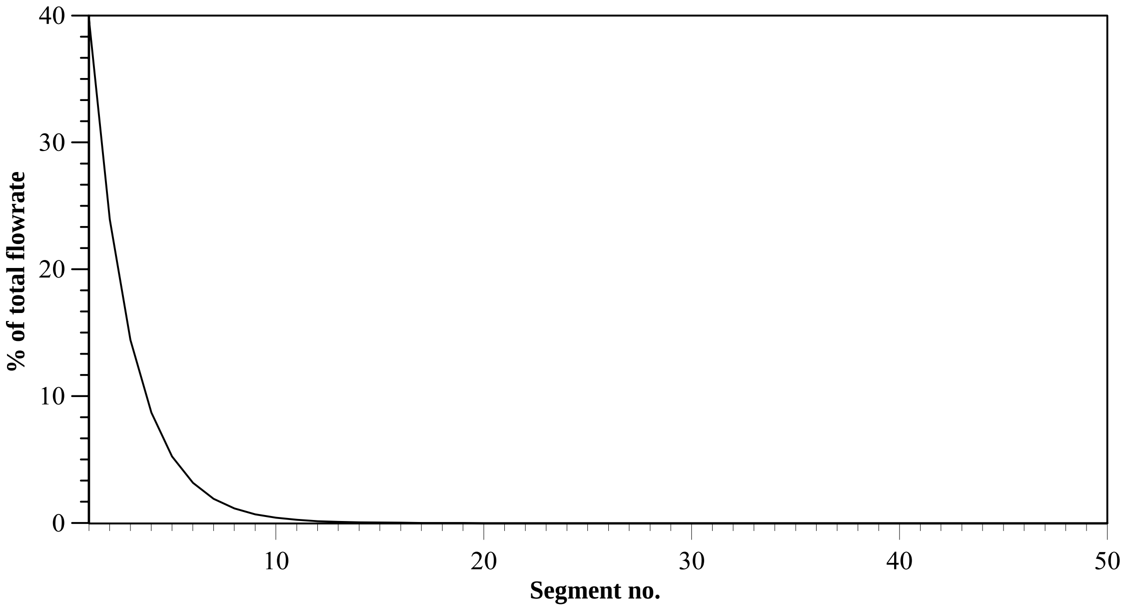


Figure 7.4 **Flowrate through each computational segment along the fibre length as a percentage of the total fibre flowrate.**

The head loss to flow relationship at the pot is given by:

$$hl_{pot} = 106.327(q_t / 2) \qquad \textbf{Equation 7.24}$$

where h_{lpot} is the head loss through the pot length.

Due to the high losses through the pot, the permeability of the virgin fibres reduces to 336.4 L m^{-2} h^{-1} bar^{-1} when placed in a full module. The flowrate of each segment along the fibre as a percentage of total flowrate is the same as in Figure 7.4.

Figure (7.5) shows the TMP as a percentage of the total applied pressure head along the fibre length. It appears that most of the applied head is used to permeate water from the first 10 cm of fibre length.

Figure 7.6 shows the TMP available at each segment along the virgin fibre length, indicating that 83.2% of the applied energy is lost through the pot length.

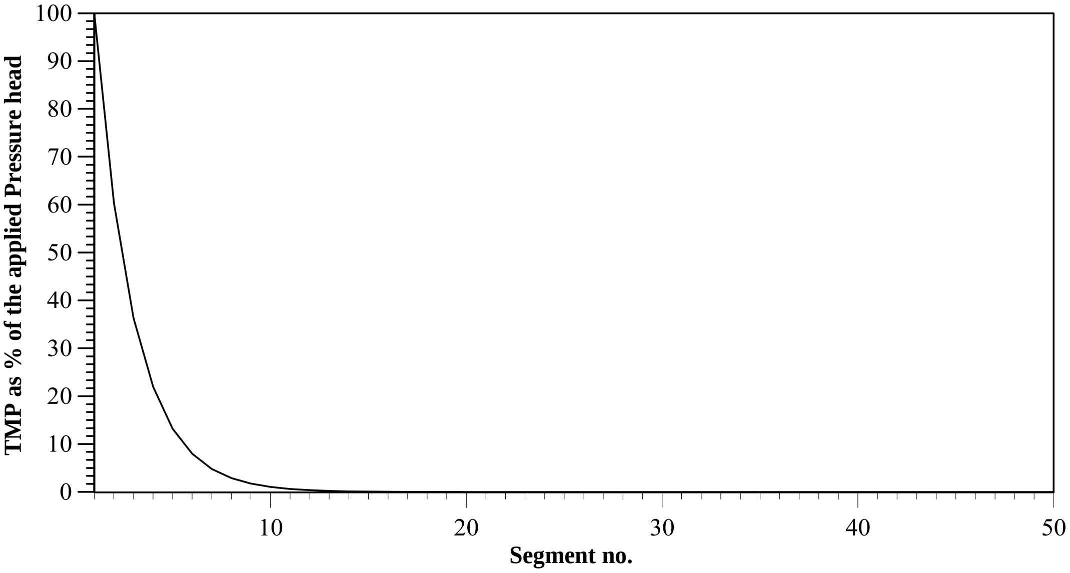


Figure 7.5 **TMP of each computational segment along the fibre length as a percentage of the applied pressure head.**

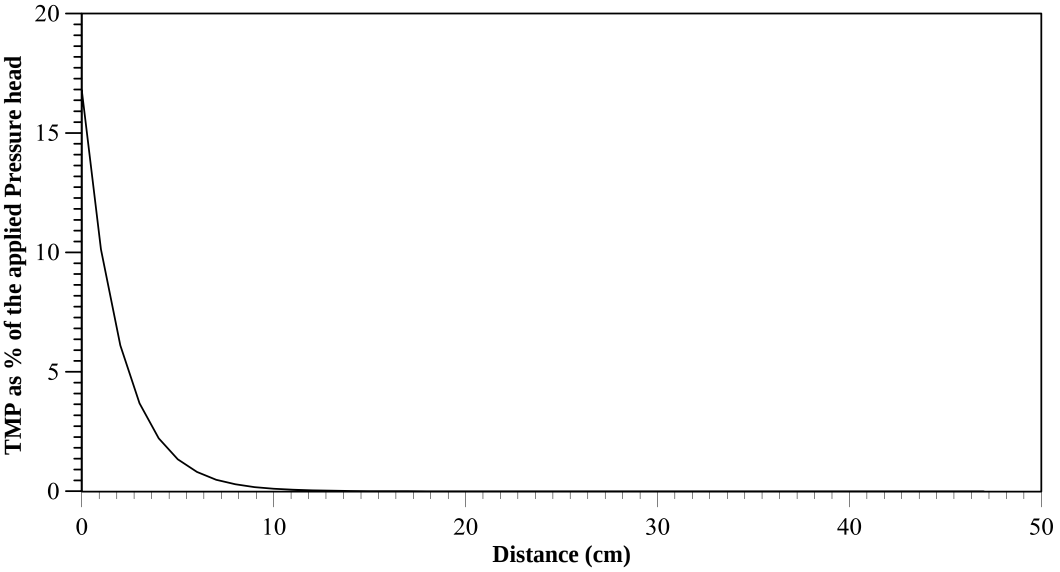


Figure 7.6 **TMP of each computational segment along the fibre length as a percentage of the applied pressure head.**

7.9 MATHEMATICAL MODEL APPLICATION: UF-CT

The ultrafiltration capillary tube fibre (X-Flow S-225 FSFC) module specifications are listed in Table 7.5 (source X-Flow, 2005).

As before the hydraulic conductivity coefficient was calculated and in this case found to be 2.57×10^{-7} cm s^{-1} ; the maximum operation permeability. The flowrate through each computational segment a long the fibre length as a percentage of the total fibre flowrate is shown in Figure 7.7. The ratio between the flowrate at the module outlet and that at its middle is approximately 1.1. Figure 7.6 shows the TMP as a percentage of the total applied pressure head, along the fibre length. Due to the large fibre diameter and the short length of the pot, the head losses calculated were small. The difference between the maximum and the minimum TMP along the fibre is lower than 10%.

Table 7.5 Specifications for Norit X-Flow S225FSFC Module

Item	
Fibre inner diameter	0.8 mm
Fibre outer diameter	1.2 mm
Effective fibre length	144.75 cm
Pot length	4 cm
Total effective area	35 m^2
Module Max operation permeability	357 L m^{-2} h^{-1} bar^{-1}

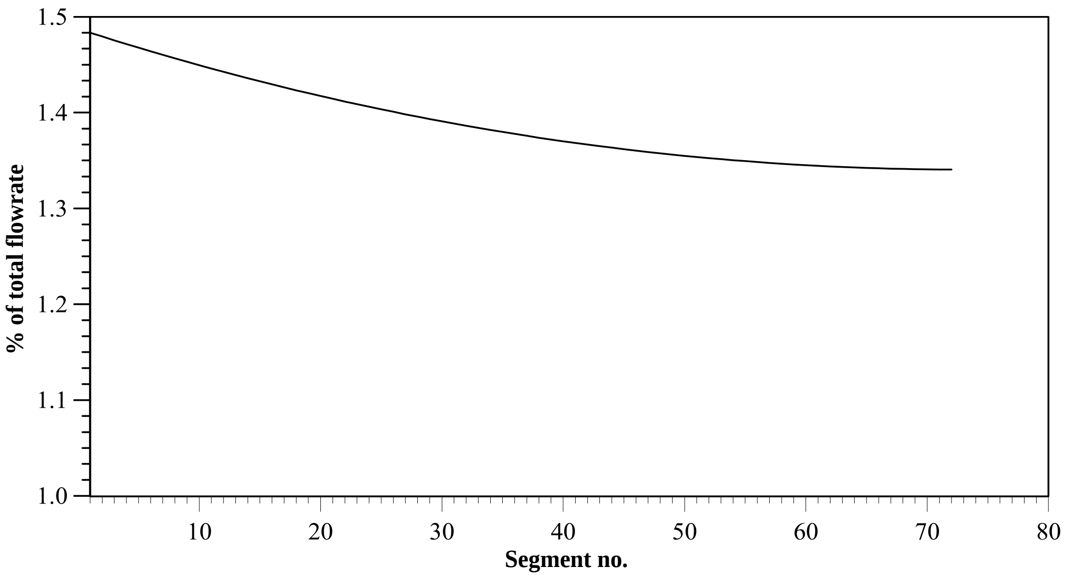


Figure 7.7 **Flowrate through each computational segment along the fibre length as percentage of total flowrate.**

7.10 HYDRAULIC FLOW UNDER VARIABLE HYDRAULIC CONDUCTIVITY (FOULING FIBRE SIMULATION)

In actual operational conditions the hydraulic conductivity coefficient of the fibre is not constant along the length; varying with operational conditions and along the fibre with ageing. The rate at which the hydraulic conductivity changes at any point along the fibre depends on the water quality thus the variation along the fibre length depends on the flowrate distribution along the fibre length.

The following scenario for either an in-out or out-in with duplex flow is suggested to explain the process of the hydraulic conductivity change along the fibre.

When a virgin or cleaned module is operated, the flowrate will distribute along the fibre length depending on the TMP along the fibre. Most of the flowrate will be concentrated at the potted ends of the fibre due to high TMP compared with other parts of the fibre and at the centre of the fibre the available TMP is less due to the head losses through the fibre channel. The hydraulic conductivity will change steadily during operation and this transient follows the flowrate distribution due to the different rejection rates along the membrane. The higher the flowrate through membrane wall at a discrete point, then the higher material rejection that caking on the membrane surface. At any point along the fibre the rate of change in the hydraulic conductivity varies with the feed water characteristics and the foulant nature. This will progress until the conductivity varies along the fibre such that the flowrate becomes distributed uniformly along the fibre length. At this flow distribution equilibrium the rate of change of the conductivity remains constant along the fibre length until terminal flux is reached.

To demonstrate this scenario, experiments were carried out on two modules with ten Memcor M10 type fibres each 1m in length built in. The initial flowrate for each module was measured, before fouling was carried out using a surrogate. The final flowrate was then measured.

The surrogate foulant was a proprietary high quality Indian ink (Colart Ltd). The products solids content is 19.5 wt.%, and as supplied had a measured suspended solids content of 200,000 mg L^{-1}. The binder for the pigment is alkali solubilised shellac and acrylic resin dispersion in water (Maginnis, 2005). This was added to make a 1% v/v solution in ultrapure water giving a concentration of 2000 mg L^{-1}. The particle size of the foulant was examined by SEM microscopy. The average particle size of the individual carbon nanospheres was 50 ηm and in pH7 DI water these nanospheres agglomerated to colloids of average size 500 ηm. The solution was kept in a stirred vessel and fed to the test membrane segments at a constant head of 2.0 bar. Figure 7.8 shows the SEM images used surrogate foulant characterisation.

The equipment and method is as given previously by Porcelli et al. (2009). Figure 7.9 shows the permeability declines with the mass of permeate collected from a set of the fibres used. The initial permeability was >400 $L.m^{-2}.h^{-1}.bar^{-1}$ and declined rapidly at 35 L m^{-2} h^{-1} bar^{-1} per minute before reaching a steady throughput of 65 $L.m^{-2}.h^{-1}.bar^{-1}$.

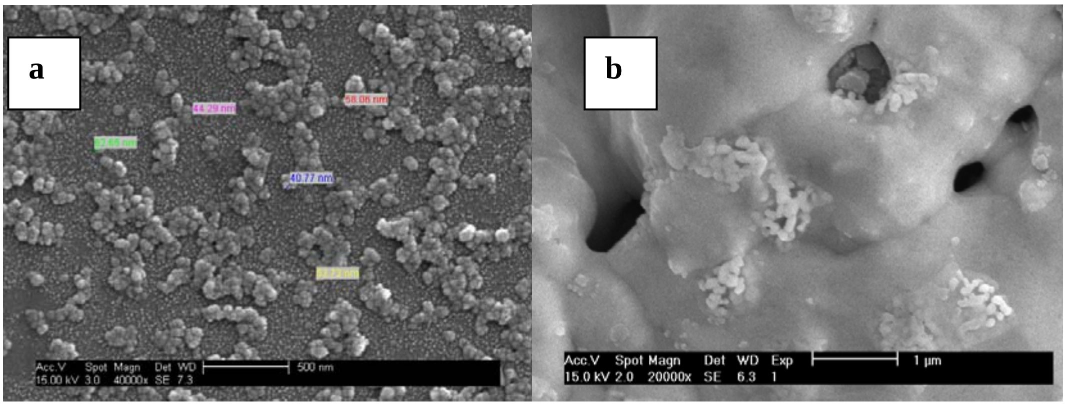


Figure 7.8 **SEM image of analogue foulant particles a) on silica glass at 40,000 X and b) in the pores of a cellulose acetate membrane at 20,000X.**

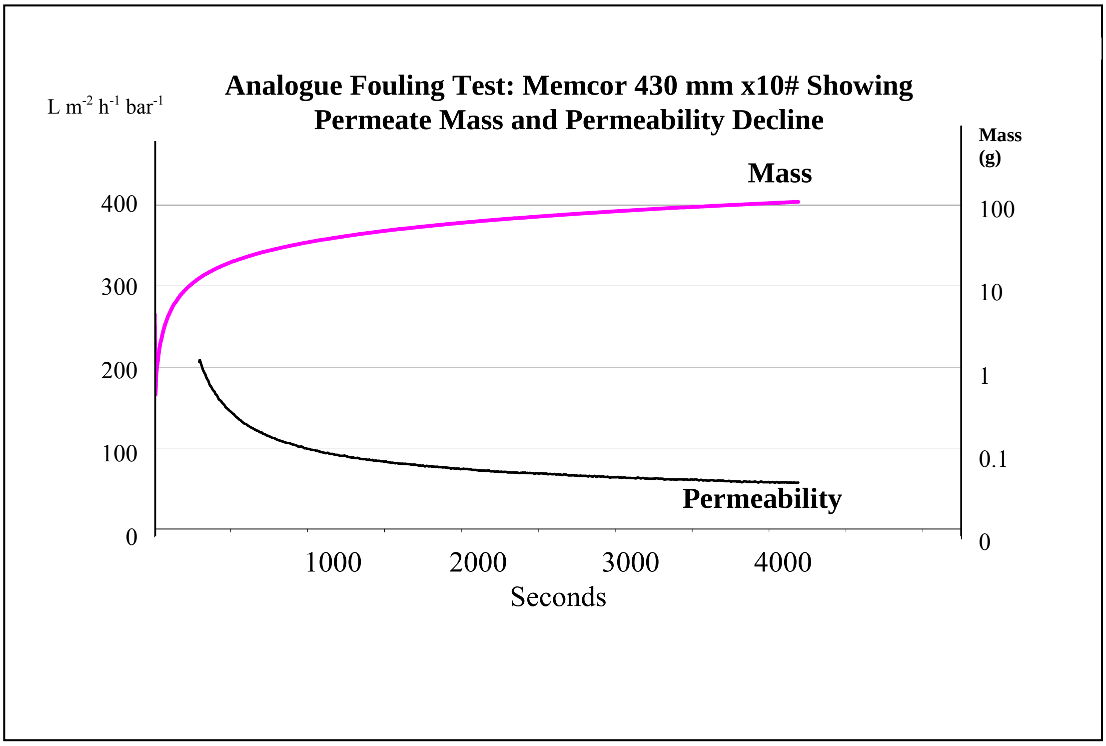


Figure 7.9 **Analogue permeability decline on Memcor MF membrane.**

The fibres were assembled into two sets and fouled with the analogue. The initial and final flowrate values for the whole fibre lengths are listed in the Table 7.6.

Table 7.6 **Analogue Fouling Test: Initial and Final Flowrate**

Set no.	Initial flowrate (mL min^{-1})	Final flowrate (mL min^{-1})
1	20.1	11.7
2	20.8	15.6

Then the fibre sets were then divided into four equal parts of 20 cm length, and the flowrate through each was measured. The effective length was reduced to 10 cm and the flowrate re-measured. Table 7.7, shows the flow rates for the segments.

Table 7.7 **Measured flowrate through parts of the fibre sets.**

Set no.	Segment no.	20 cm Length Flowrate (ml min^{-1})	10 cm Length Flowrate (ml min^{-1})
1	1	1.96	1.14
	2	6.95	4.1
	3	8.92	5.73
	4	9.68	6.27
2	1	4.16	2.96
	2	8.53	4.9
	3	9.1	5.7
	4	9.72	6.02

Figure 7.10 shows the predicted flow distribution of each computational segment along the fibre using the mathematical model for a virgin membrane fibre of the same length used in the experiments. Because of the change in the TMP along the fibre length, the flowrate distribution is not uniform along the fibre. Therefore, the material rejection along the fibre is not uniform which explains the experimental results showing variable flowrate along the fibres after the fouling process.

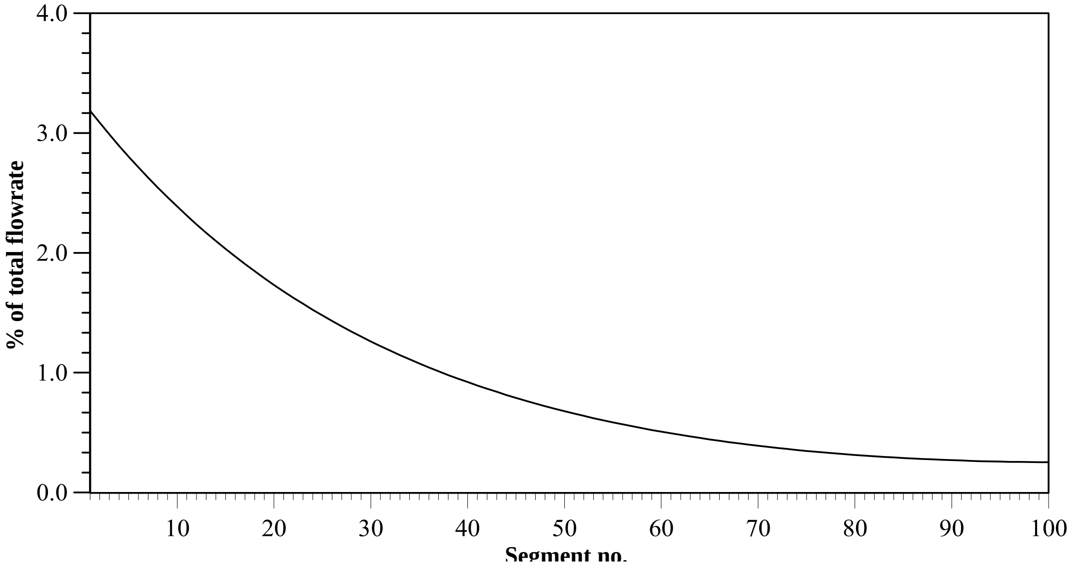


Figure 7.10 Flowrate through each computational segment along the fibre length as percentage of flow

Table 7.8 shows the corresponding hydraulic conductivity of each part, after the fouling that was predicted using the mathematical model, for the the hydraulic conductivity coefficient for two test lengths on each test set.

HF membranes have different fouling characteristics since the flowrate and headloss along the fibre depend on the design parameters (length, inside diameter, outside diameter, pot length, and the hydraulic conductivity of the fibre material. Figure 7.11 shows the mathematically predicted flow rate for variable inner fibre diameter with flowrate distribution; keeping the pot length, the hydraulic conductivity and the difference between the outer and inner diameter constant. It demonstrates how the flowrate will be distributed along fibres and therefore fouling rate along the fibre length.

Table 7.8 Hydraulic conductivity calculated for segments of two different test lengths.

Set no.	*Segment no.*	*20 cm Length Flowrate (cm s^{-1}*10-7)*	*10 cm Length Flowrate (cm s^{-1}*10-7)*
1	1	0.7	0.8
	2	2.6	2.9
	3	3.5	4.1
	4	3.8	4.5
2	1	1.2	2
	2	3.3	3.5
	3	3.6	4
	4	3.84	4.35

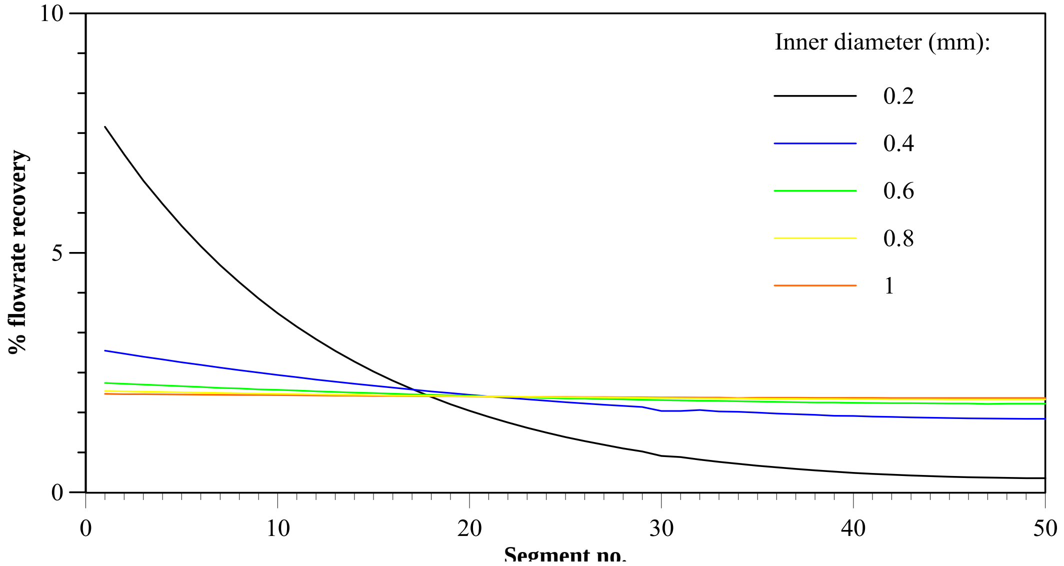


Figure 7.11 Flowrate profile for different lumen diameter

7.11 FIBRE SENSITIVITY TO FOULING

During the operation of the fibre module, fouling material is accumulated in and on the fibre wall, causing more resistance to the flow and a reduction in the TMP and thus a change in the flowrate. Depending on the hollow fibre membrane parameters, fibres differ in the relation between the flowrate and the TMP. Figure 7.12 shows the relation between the flowrate and TMP for a number of fibres of 50 cm length. The slope of the TMP relationship to flowrate for the larger Norit fibre is higher than the of Memcor MF fibres indicating better hydraulic permeability based on geometry alone.

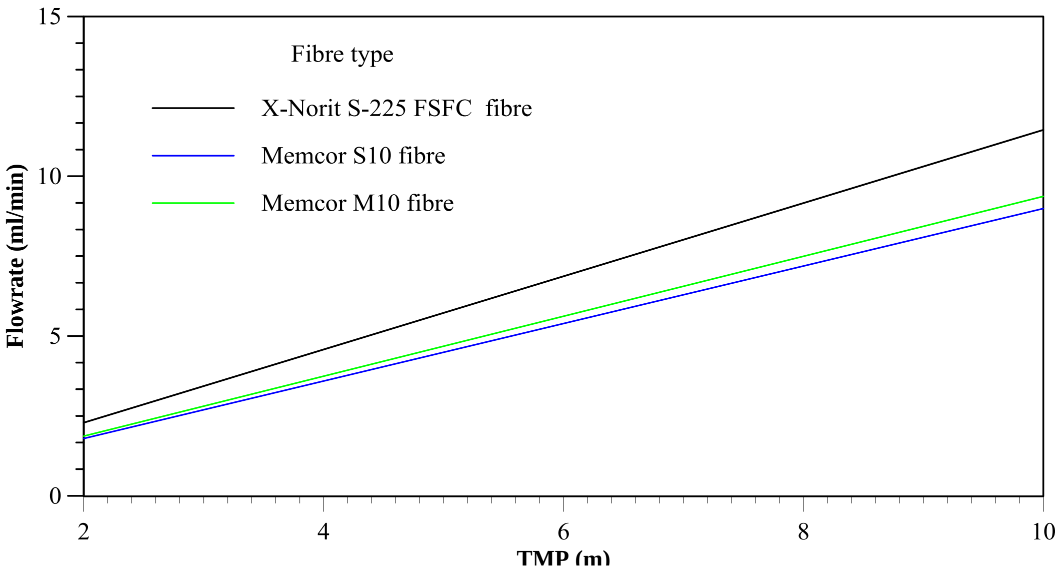


Figure 7.12 Flowrate and TMP Relationship for Different Fibres

7.12 CONCLUSIONS

1. The flow though the hollow fibre membranes under the condition of constant hydraulic conductivity can be mathematically simulated and verified applying Darcy's and Hagen Poiseulle laws. Good agreement was found between laboratory and mathematically predicted data under same conditions.
2. The transmembrane pressure (TMP), as a percentage of the total applied pressure head, and flowrate distribution along the fibre length, as a percentage of the total flowrate along the fibre length, are independent of the applied pressure head. At the initial stages of filtration the flux is concentrated at the pot ends. As a cake has not developed along the fibre and the throughput is concentrated in a smaller area, the membrane is at the most risk of integrity failure at this time. TMP cycling ranges are higher at the ends, making them more exposed to mechanical integrity failure in these regions.
3. The head losses through the pot length are relatively high percentage, such that energy efficiency decreases with decreasing fibre internal diameter. A fouling test with the surrogate foulant provided results consistent with the theory.
4. The efficiency of fibre use decreases with decreasing fibre internal diameter, as confirmed by a comparison of the geometries of the Memcor MF and the Norit UF fibre and the corresponding flow patterns. It was concluded that this could have more bearing on differences in heterogeneous fouling than raw water characteristics and development of cleaning models should reflect this.
5. The fouling condition along the fibre is not uniform, depending on the flowrate distribution along the fibre which in turn depends on the relationship between fibre design parameters. The MF fibre reaches a uniform flux faster for the same water than the CT UF fibre. The backwashing effectiveness and physical cleaning is therefore more critical for this fibre design.

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CHAPTER 8 INNOVATION IN THE UK WATER INDUSTRY AND MEMBRANE PLANT IMPLEMENTATION IN AMP 3

Porcelli, N. and Judd, S., Innovation in the UK Water Industry and Membrane Plant Implementation in AMP 3, Submitted to *Water and Environment Journal, (CIWEM)..*

Abstract

The UK water industry functions by investing in and utilising a mixture of technologies, to maintain the quality of drinking water supply along with environmental water quality standards. Regulated water companies historically have low R&D expenditure compared to other sectors which it is believed restricts development and implementation of new technologies. Capital spending is only allowed with good justification. Past studies show the motivations for innovation within such companies to be complex but driven primarily by risk management concerns than a desire for technological innovation.

This study examines organisational factors influencing uptake and implementation of membrane technology in an asset investment period. Econometric parameters for five water organisations and a survey quantifying perceived receptivity were compared. Results show that capital efficiency correlated well with receptivity perceptions; suggesting other technology adoption factors are obscured by economic focussed utility regulation. The research suggests that some of the impediments to innovation are external to the technologies and organisations themselves and to innovate the water utilities in the UK would benefit from further research to widen innovation acceptance at the operational level.

8.1 INTRODUCTION

8.1.1 Innovation and the UK water industry

The UK water industry is privatised: with 12 water and sewerage companies (WaSCs), a number of smaller water only companies (WoCs), a range of supply chain partners to these companies including engineering consultancies, contractors and equipment manufacturers, and Government regulators covering economic (The Water Services Regulation Authority, OFWAT), drinking water quality (The Department of Environment, Food and Rural Affairs (DEFRA) Drinking Water Inspoectorate, DWI) and water resources (DEFRA-Environment Agency EA) aspects of water and wastewater service provision. The industry relies on successful technology adoption for innovation but implementation success is variable (UKWIR, 2006). Prescriptive, externally developed technology is adopted, but successful implementation can be adversely affected by a lack of interpretative systems (Kaplan and Tripsas, 2003). Indeed, the industry has been subject to criticism over the past five years for demonstrating a lack of innovation. These criticisms have centred on inter alia the predominance of institutionalised beliefs in the suitability and attractiveness of large scale centralised infrastructures and relatively risk-free, tried-and-tested treatment technologies (Thomas and Ford, 2005).

Investment and R&D spend levels are organised into regulatory periods, or Asset Management Plans (AMPs), which structure investment in the sector (Table 8.1). Critiques of innovation within the UK water industry have concluded that it is economically limited by historically low R&D spending compared to other utility sectors (UKWIR, 2006 ; HMG , 2006; Cave, 2009), particularly from AMP-2 to AMP-4. The AMP structure and process tends to constrain investment to incremental and capital-intensive projects and UKWIR (2006) studies of the responses to guided questions to water professionals show that the industries concerns are related to the fragmented nature of the organisations and the mandatory separation of suppliers from the commercial and regulatory issues facing the business. Radical change may also have been avoided in relation to the provision of drinking water due to fear of increased exposure to unknown public health risks.

Table 8.1 Asset Management Period, CAPEX, England and Wales

Period	Year	Investment £-bn	Drivers	R & D Spending
AMP-1	1989 1994	25	EU DW directive, Activated carbon & ozone technology, coagulation	£36 m pa; AMP 1&2 Technologies research programme
AMP-2	1995 1999	30	EU Urban WW directive, bathing waters, Water quality at customers tap	£40 m pa; Technologies and systems
	2000 **2004**	**31**	**Cost saving AMP 4 cost efficiencies- Leakage, *Cryptosporidium***	**£30 m pa; AMP3 cost saving AMP 4 capex**
AMP-4	2005 2009	32.6	Sewer flooding capital maintenance, Common Framework.	£24 m pa; OPEX, PR09 submissions
AMP-5	2010 2014	33.6	WFD (2000), Supply and demand, Universal metering, AMP1 asset renewal. Security of supply and Water Safetyt Planning.	£40 m pa; Energy from waste, smart assets
AMP-6 & AMP-7	2015 2030	36.5	Carbon emissions, supply and demand, smart meters and tariffs, water and waste recovery and recycling, waste to energy.	Not given.

(UKWIR, 2007b)

Study period -AMP 3 2000 to 2004 highlighted.

8.1.2 Innovation theory and practice

In the academic literature innovation is regarded from a socio-economic constructivist perspective as resulting from learning activity which allows firms or organisations to go beyond the boundaries that limit new ideas (Augier and March, 2008; March, 1978). Organizational learning starts from user reflection on and adaptation to technical change, i.e. tacit operational knowledge capture (Tidd et al., 2005; Love et al., 2005), and relies on knowledge emerging as a result of the complex interaction of organisational variables (Orlikowski, 2000; Jenssen and Joorgensen, 2004).

Innovation studies have examined the socio-cognitive, socialisation or thinking interactions of the way in which learning organisations evolve (Lemon and Sahota, 2004; Nonaka, 2007; Starbuck and Hedberg, 2001). Cognitive theories relying on a structural analysis of the organisation (Carayannis, 1999; Carayannis and Alexander, 2002) have good explanatory power of innovation capacity but there are problems in application where understanding of technology perception is affected by a poor knowledge base (Edkins et al., 2007; Swan and Clark, 1992). In addition, along with other purely social theories of technological innovation, cognitive theories tend to conflate a technology's materiality (the material constraining and enabling influences of technology on individual and group behaviour) with determinism. To counter this conflation, and to better reconcile technological materiality with voluntarism in theories of organisational change, a conceptual framework has been proposed which centrally locates how technology is actually used (Orlikowski, 1992; Galliers, 2006) ; Leonardi and Barley, 2008; .

Since people interact practically with technology, dynamic socio-material models are needed (Orlikowski, 2007). "Receptivity" is a conceptual model examining inward technology transfer, where openness or readiness to accept new ideas represent success criteria (Seaton and Cordey-Hayes, 1993). Receptivity of technology transfer processes comprises: external 'awareness' or scanning; perceived utility or 'association'; adoption or 'assimilation' and implementation or 'application' (Jeffrey and Seaton, 2004). Learning is important for all parts and is influenced by the user perception in non-routine activities and organisational communications through "champions" (Trott et al., 1995).

Practical insights from receptivity models, for managing processes of technological innovation, are articulated in terms of knowledge base management (Trott, 2005). Figure 8.1 shows an adaptation of a the Trott (2005) knowledge management triangle: showing the range of socio-technical knowledge inputs and bases which must be managed within water companies and their supply chain partners, and which may influence technological innovation outcomes.

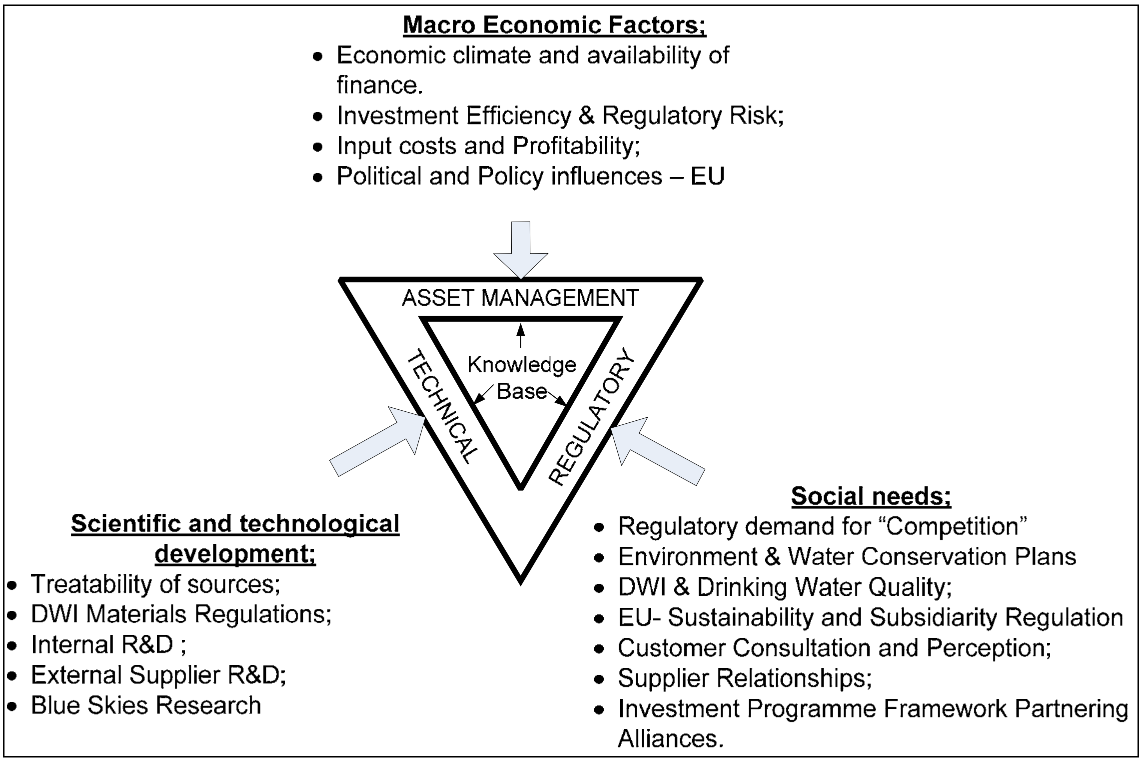

Figure 8.1 **Innovation and knowledge base management - Water Industry UK England and Wales.**

8.1.3 Aims and objectives of the study

In contrast to the focus of these models which emphasise the role of human behaviour and learning in technological innovation, regulation uses a variation of the utility econometric benchmark model (Shleifer, 1985; Helm and Jenkinson, 1997) developed to facilitate but not replicate market processes (Littlechild, 2008). The aim is to promote higher quality service delivery especially as measured in capital efficiency. This internationally recognised regulation model relies on state governed overseeing authorities and in the UK the water industry economic regulator is OFWAT. The UK regulator employs econometric models developed at privatisation in 1990, to assess and incentivise improvements to capital efficiency for service delivery by both WaSCs and WoCs (Allan, 2006; Cashman, 2006).

The UK water industry is committed strategically to increasing its innovative effort, and the regulator has expressed concern over how this may be achieved (UKWIR, 2007a). It is thus of interest to ascertain whether driving water companies (both WaSCs and WoCs) towards particular levels of econometric performance through regulation is positively related to innovation success measures in terms of the adoption and use of new technologies by those companies. If so then the more highly performing companies would be expected to also be the most innovative.

This paper presents the results of a case study of a technological innovation by a number of water companies during the AMP-3 (1999-2004) period. In this AMP period the water companies were obliged to implement drinking water quality improvements that entailed an inward technology transfer. Membrane filtration technology was prescribed by the drinking water quality regulator, the DWI, on the basis of its proven success in protecting high risk water supplies, particularly groundwater sources, from spreading illnesses arising from *cryptosporidium* oocyst contamination (Morris and Whitehead, 2006). Subsequently, operational problems and variations in uptake have led to an examination of the design, operational and implementation management aspects of membrane technology for water treatment (Judd, 2004; Presdee, et al., 2006; De Wilde et al., 2008). The reasons behind variations in uptake and in the distribution of implementation problems between water companies are not well understood. Critics have however used the lack of general implementation of membrane technology as an example of innovation failure in the UK water industry (Thomas and Ford, 2005).

This study attempts to establish, through empirical analysis of published data and reports from innovation practice:

1. possible intercorrelation between technology uptake, innovation capability, organisational receptivity and econometric performance measures, and
2. implications of this for enhancing innovation within the sector.

A basket of representative econometric performance measures were selected and reduced following a factor analysis (Section 8.2) to provide an independent set of explanatory variables. Determinants of organisational receptivity to technological innovation were then assessed using a survey instrument and combined into a single measure of receptivity to membrane technology for each water company (Section 8.3). Membrane innovation capability was characterised as uptake using a composite measure of installed plant numbers and capacity (Section 8.4). Rank order correlation analysis was then employed to examine the relationships between membrane innovation (dependent), receptivity (independent) and econometric performance (independent). The results were then interpreted for their innovation policy and management implications.

8.2 ECONOMETRIC ANALYSIS

An econometric parameter analysis was used to characterise organisation performance in a similar way to the regulatory approach, though ignoring non-business characteristics. A ranking analysis on a representative basket of industry econometric performance factors (Table 8.2) was compiled for four water and sewerage companies (WaSCo) and one water supply company (WoCo). In terms of relationship to material innovation practice these ranged from abstract financial parameters (A) to measures of operational activities (G). Econometric data (Table 8.3) were derived from OFWAT and annual company reports. The proportion of borehole sources at the time (Figure 8.2) was taken to be analogous to each companies exposure to natural *cryptosporidium* risk; this is dependent on ground and geological conditions but a dependency on ground sources presumes exposure to the risks of contamination from infiltration. Annual figures for econometrics were sourced for 1999/2000 through 2004/2005. To take account of different company business plan structures figures were expressed as the yearly average for the study period; corresponding to the AMP-3 period when membrane plants were being adopted and implemented in response to the Water Regulations of 1999. Rankings (Ro – A to G) based on the criteria are given in Table 8.2 for each parameter along with econometric rankings (R_1).

For the chosen parameters, principal component analysis (PCA) with varimax rotation was applied (Table 8.4). There were similarities between parameters D and E and also E and F. The strong influence of the quality scheme measures (D and E) to capital employed demonstrate that companies adhere to regulated business plans influencing their choice of schemes to implement. The effects of regulation can also be seen in the relationship between regulatory influence (E and F) in dictating where quality driver work is to be carried out and thus the level of capital investment in the business. Parameters D and E were thus removed their ranking from the works in the subsequent comparison with receptivity. Parameters A and B were also correlated (0.705) but not as strongly, and thus were retained.

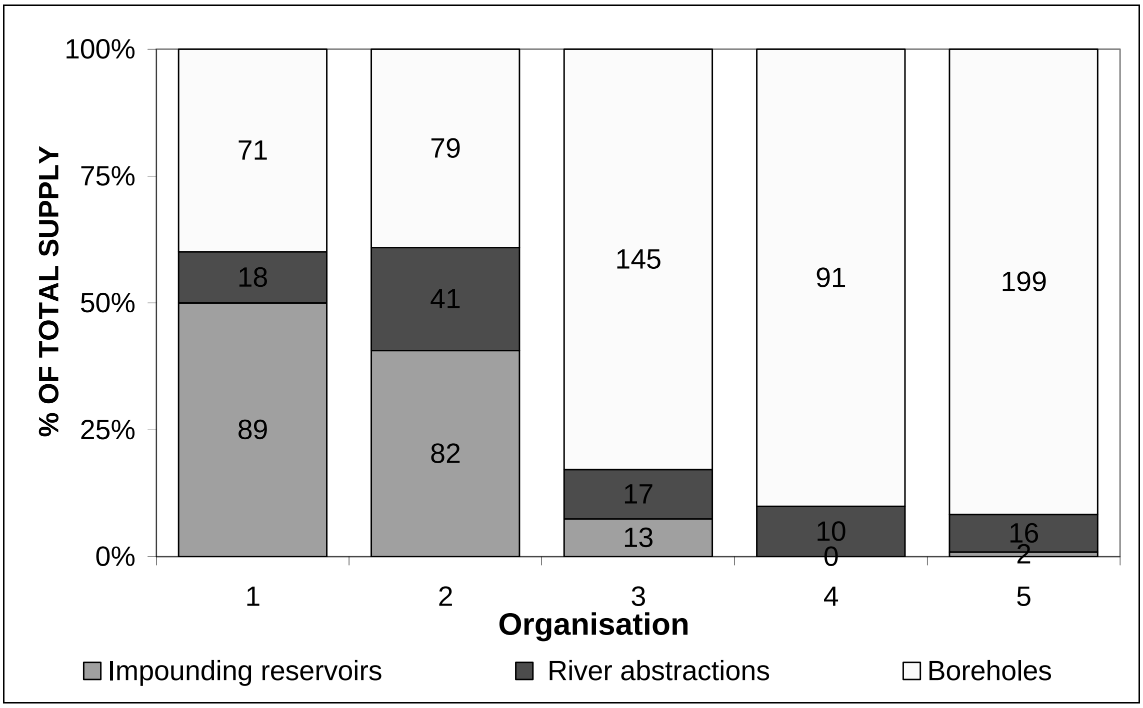

Figure 8.2 **Water source distribution for surveyed companies (UK/OFWAT RD21/07 Relative Efficiency Assessment)**

Table 8.2 **Constructed econometric factors for organisation characterisation**

A. BUSINESS TO CUSTOMER INDEX, (-).	Connected properties per £1000 of annual turnover: this indicates the service provided to customers in terms of the financial activity required and represents the organisations performance economically. A high number represents better financial performance reducing cost of capital to the business and the risk of innovation.
B. RETURN ON CAPITAL EMPLOYED, (%).	Shareholder earnings to investment level: this is a standard business investment variable and represents the returns by level of capital employed, it is directly affected by the regulator’s determination of price limits for the AMP period. A higher % indicates a higher financial return. This again reduces cost of finance and long term capital costs and indirectly lowers investment risks.
C. OPEX COSTS PER UNIT	Operational costs, in million pounds Sterling (at 2005

WATER PRODUCED, (£M per MLD).	prices), to supply a million litres daily to the water distribution network. It takes into account all parameters influencing operational costs; such as leakage and costs to treat and pump the water. A lower number may indicate operational effectiveness but takes no account of key geographic and social differences between companies. Lower inputs allows added value to be realised reducing the cost of finance and the risk of investing thus aiding innovation.
D. QUALITY SCHEME INVESTMENT TO TURNOVER INDEX, (%)	Capital investment in water quality schemes, (such as membrane plant implementation) to the whole reported annual turnover. It can be inferred that the percentage of the cost quality schemes to the turnover reflects favourably on the investment and the cultural impact on the organisation as a whole. The structures for investing in quality schemes are assumed reinforced by activity so proportionally this is an indicator of quality scheme investment capability.
E. QUALITY SCHEME INVESTMENT TO TOTAL WATER CAPITAL INVESTMENT INDEX (%).	Index of quality scheme investment (such as membrane plant implementation) to the total annual investment in water schemes only. The percentage of the investment to the total programmed investments, including maintenance and other investment in water operations is analogous to scope of the work on mandatory quality projects for this organisation. A higher value indicates a greater proportion of the quality schemes on overall drinking water operations and prospects of technology scanning affecting innovation practice.
F. REGULATORY INFLUENCE RATIO (%).	Number of DWI designated *cryptosporidium* prevention schemes to the total number of Water Quality Zones operated. A WQZ is based on treatment sources, geography and population density and allows the inspectors to monitor for legal compliance. The ratio highlights the impact of legislation and prosecution risk on the organisation and rationale driving innovation.

G. EXPOSURE TO REGULATIONS RATIO (-).	Ratio of average annual volume of water produced in MLD for *cryptosporidium*-monitored treatment plants to the total water production; this represents the proportion of the water treated considered for risk analysis by the DWI under the Water Regulations in 1999. A higher value indicates greater exposure of the organisation to regulatory risk and the necessity for adoption of membranes affected the treatment operations.

Table 8.3 Econometric performance values for UK (England and Wales) water utility organisations studied

	ECONOMETRICS WITH RANK R_0 see note 1													
	A		**B**		**C**		**D**		**E**		**F**		**G**	
ORG.	*n/a*	**R_o - A**	*%*	***R_o - B***	*£M/ML/d*	***R_o - C***	*%*	***R_o - D***	*%*	***R_o - E***	*%*	***R_o - F***	*n/a*	***R_o - G***
1 WaSC	**3.4**	***3***	**7.0**	***2***	**13.0**	***1***	**8.0**	***2***	**37.9**	***2***	**17.0**	***2***	**0.0**	***5***
2 WaSC	**3.0**	***4***	**6.4**	***4***	**14.6**	***4***	**14.2**	***1***	**50.3**	***1***	**17.6**	***1***	**0.7**	***1***
3 WaSC	**2.0**	***5***	**6.9**	***3***	**13.7**	***3***	**1.8**	***5***	**20.8**	***3***	**2.6**	***3***	**0.3**	***2***
4 WoC	**6.7**	***1***	**8.1**	***1***	**15.7**	***5***	**5.0**	***3***	**13.9**	***5***	**2.6**	***3***	**0.3**	***3***
5 WaSC	**3.7**	***2***	**6.0**	***5***	**13.2**	***2***	**2.3**	***4***	**14.5**	***4***	**0.0**	***5***	**0.0**	***4***
Mean	*3.7*		*0.1*		*14.0*		*0.1*		*0.3*		*0.1*		*0.3*	

[1] *Averages of years 1999/2000 to 2004/2005. Sources: OFWAT annual reports 1999 to 2005, company reports 2000-2005.*

Table 8.4 Correlation Matrix Showing Econometric Ranking R1

	A	B	C	D	E	F	G
A	**1.000**						
B	**0.705**	**1.000**					
C	**0.688**	**0.668**	**1.000**				
D	**-0.066**	**-0.106**	**0.237**	**1.000**			
E	**-0.428**	**-0.291**	**-0.101**	**0.915**	**1.000**		
F	**-0.266**	**-0.113**	**-0.110**	**0.888**	**0.956**	**1.000**	
G	**-0.111**	**0.028**	**0.611**	**0.623**	**0.509**	**0.324**	**1.000**
Eigen val.	3.336	2.473	0.848	0.344	0.000	0.000	0.000
% of var.	47.652	35.326	12.110	4.912	0.000	0.000	0.000

Table 8.5 Single and joint econometric ranking of water companies (R0-A – R0-G, R1)

	RANK	ECONOMETRIC				
ORGANISATION	*Total,* ***Rank*** R_1	$R_{0\text{-}A}$	$R_{0\text{-}B}$	$R_{0\text{-}C}$	$R_{0\text{-}F}$	$R_{0\text{-}G}$
1 WaSC	13 = ***1.5***	***3***	***2***	***1***	***2***	***5***
2 WaSC	14 = ***2***	***4***	***4***	***4***	***1***	***1***
3 WaSC	16 = ***3***	***5***	***3***	***3***	***3.5***	***2***
4 WoC	13 = ***1.5***	***1***	***1***	***5***	***3.5***	***3***
5 WaSC	18 = ***4***	***2***	***5***	***2***	***5***	***4***

Table 8.5 shows how each company is ranked in terms of each independent econometric parameter following the factor analysis. Rank 1 for any given parameter is given to the company with the lowest value for that parameter. If the selected econometrics used are representative indicators and drivers of water company innovation capability then high values would be expected to correlate with higher membrane innovation capability, and thus membrane uptake rates, for each econometric. To show this, a joint econometric ranking (R_1) taken as the total of the separate R_0 rankings was determined in addition to the separate econometric rankings. Higher rankings indicate an expected higher membrane innovation capability and therefore uptake rate.

8.3 MULTI-ORGANISATION RECEPTIVITY SURVEY OF MEMBRANE TECHNOLOGY ADOPTION PRACTICE

A receptivity survey of the five water organisations was carried out. The survey quantified the perception of receptivity practices, for each organisation in implementing new membrane plants over the same period as that of the econometric analysis above (AMP 3 period), and then ranked them. The procedure was based around structured interviews with the relevant operational manager responsible for overseeing the implementation of the necessary *cryptosporidium* membrane plants in the AMP-3 period. Similar methods have been used in studies into the uptake of advanced machinery in manufacturing firms to explore how strategy and culture affected successful implementation (Lewis and Boyer, 2002). These studies characterised organisations that successfully innovate as having strategies emphasising flexibility in processes and systems at operator level, adherence to quality specifications, sufficient training, pilot projects, the existence of a boundary spanning "champion" in the end user organisation and a focus on long-term technology objectives (Lewis and Boyer, 2002; Sohal, 1996; Waldeck and Leffakis, 2007).

The structured interviews designed check the respondent's agreement or disagreement with twenty six "receptivity positive" statements that represent five key areas of operational practice in relations to inward transfer of membrane technology in this period.

Areas tested were:

- an existence of cultural values such as flexibility (A1, 7 questions);
- adequacy of training provision and ability to carry out procedural and systems changes (B1, 5# questions);
- accurate treatment specification setting and the capability to meet the regulatory drivers (B2, 4 questions);
- the importance of pilot testing (B3, 4 questions) and
- values placed on having a recognised "process champion" behind the technological shift (B4, 4 questions).

Respondents for each of the five water companies were identified and contacted. They were all water company employees responsible for integrating the new technology with existing systems and for liaison with the "framework project" partners responsible for design, construction and commissioning. None were involved in technology choice or piloting decisions. On-site interviews were organised and respondents were asked to score their own organisations practices for each "receptivity positive" statement, representing successful implementation programme. Five interviews were conducted. The approach was limited to these managers as they were in the sole position to track and collate problems in the implementation process from conception to commissioning on each project. Lickert Scale scoring (Barnett, 1991) was used, with ordinal responses to each question collected with 7 being "agree strongly" and 1 "disagree strongly". The median response for each of the area was calculated for the respondents and shown as a "spider" plot (Figure 8.3).

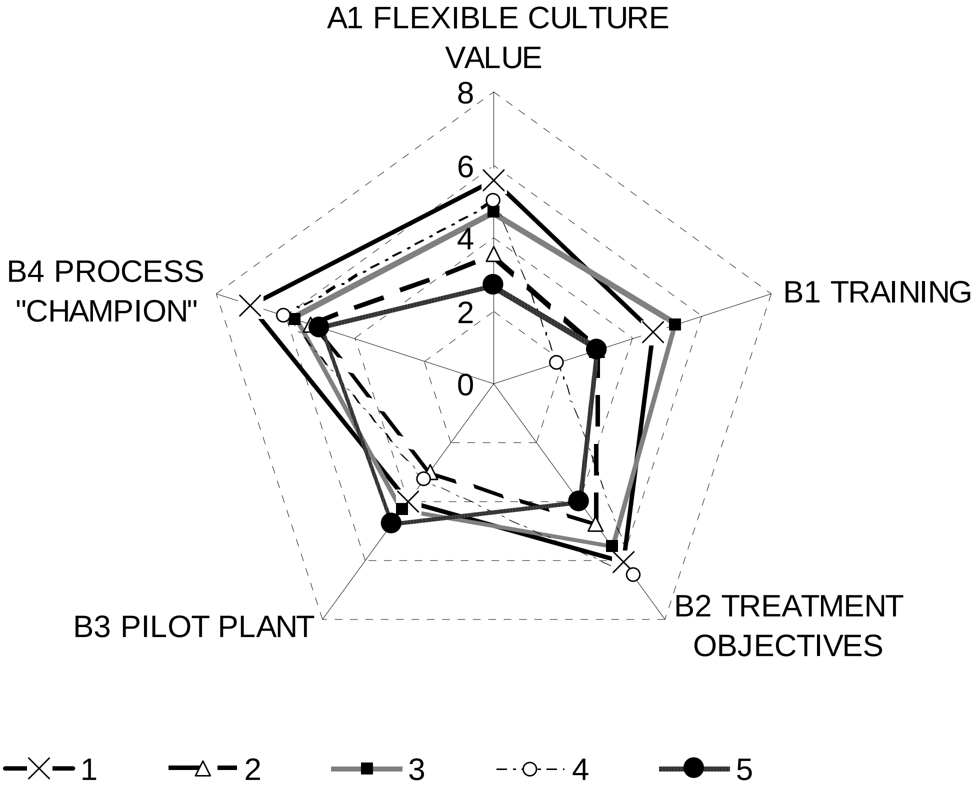

Figure 8.3 **Receptivity to membrane technology showing median response**

The total of each companies "receptivity scores" were converted to a percentage of the total positive response possible and ranked (R_2, Table 8.6). This ranking is considered to reflect the organisational "receptivity to membrane technology" as perceived from historical consideration of the success of their innovation practices for membrane technology installation in the AMP 3 period.

Table 8.6 Self- assessed receptivity rank (R2) for each water company

ORGANISATION	*RECEPTIVITY (%)**	***RANKING, R_2***
1	78%	**1**
2	56%	**4**
3	73%	**2**
4	64%	**3**
5	56%	**5**

**Average = 64%*

A comparison between each water company's econometric and receptivity-based rankings is given Table 7 using the Spearman rank correlation p:

$$p = 1 - \frac{6\sum d_i^2}{n(n^2 - 1)}$$

Equation 8.1

where n is the number of samples and d the distance between the ranking sets to be compared. The analysis presented in Table 8.7 provides an indication of how each of the seven individual econometric factors relates to water company innovation capability, as measured by receptivity ranking R2.

Table 8.7 Relationship between econometric measures ($R_{0A\ to G}$) and receptivity to membrane technology (R_2)

		ECONOMETRIC FACTOR RANKINGS R_o				
ORGANISATION	***R_2***	***R_{0-A}***	***R_{0-B}***	***R_{0-C}***	***R_{0-F}***	***R_{0-G}***
1	**1**	3	2	1	2	5
2	**4**	4	4	4	1	1
3	**2**	5	3	3	3.5	2
4	**3**	1	1	5	3.5	3
5	**5**	2	5	2	5	4
	Correlation	***-0.26***	***0.70***	***0.32***	***0.35***	***-0.26***

Four out of five of the employed econometrics were only weakly related to receptivity based on the correlations shown (Table 8.7). The one exception is Econometric R_o-B (return on capital employed), which is positively correlated with self-assessed receptivity to membrane innovation. The heavily regulated structure of the industry, where revenue growth and capital spending are coupled, poses the threat of regulatory risk; overregulation makes the industry less able to raise funds to support their investment programs. Perceived risks and inefficient use of capital may suppress innovation in some companies; in the UK regulatory bodies are apparently aware that changing capital incentives alone is insufficient to stimulate innovation (Cave, 2009).

8.4 CHARACTERISING WATER COMPANY MEMBRANE INNOVATION CAPABILITY

To characterise water company membrane innovation capability, the dependent variable in the study, a measure to reflect membrane technology uptake (or implementation) was developed. As already mentioned, each water company had to implement a number of membrane schemes in the period in response to DWI instructions. The scope and success of implementation varied between companies, and this variation in terms of the measure developed with the receptivity (R_2) and econometric rankings (R_0 and R_1) was thus appraised.

Data on membrane application (MF and UF technologies) came from the DWI annual reports for each company in the period, providing an indication of regulatory undertakings, as well as published studies of membrane plant implementation in this period (Rachwal and Judd, 2006). The number of plants implemented and operated and their capacity were compared to the totals for the organisation to generate ratios. These ratios were then multiplied to give a joint measure of membrane uptake, which was then ranked as R_3 (Table 8.8).

Table 8.8 Water company membrane implementation rankings – R3

	Plant Number			*Plant Capacity*			*Index*	*Uptake*
Organisation	Number #	Of Total #	*Ratio 1*	Capacity ML/d	Of Total ML/d	Ratio 2	*Ratio 1 x Ratio 2*	Rank R_3
1	4	121	*0.03*	95	1257	*0.08*	*0.0025*	**3**
2	2	161	*0.01*	99	1940	*0.05*	*0.0006*	**4**
3	12	164	*0.07*	137	2000	*0.07*	*0.0050*	**2**
4	4	77	*0.05*	312	700	*0.45*	*0.0232*	**1**
5	0	139	0.00	0	1200	0.00	0.0000	**5**

Table 8.9 Membrane Practice Ranks with Econometric and Receptivity Ranks.

ORGANISATION	R_3	R_{0-A}	R_{0-B}	R_{0-C}	R_{0-F}	R_{0-G}	R_1	R_2
1	3	3	2	1	2	5	2	1
2	4	4	4	4	1	1	1	4
3	2	5	3	3	3.5	2	4	2
4	1	1	1	5	3.5	3	3	3
5	5	2	5	2	5	4	4	5
Correlation with R_3		**0.1**	**0.9**	**-0.5**	**0.051**	**0.1**	**0.61**	**0.6**

Using Spearman's ranking analysis again, this uptake (and by proxy innovation capability) ranking was compared with the econometric and receptivity rankings (Table 8.9). The correlations indicate that overall the separate econometric indicators do not correlate at all well with membrane uptake, with the exception of econometric

B (return on capital employed) which is strongly correlated (r = 0.9). The correlations between the joint econometric ranking (R_1) and the receptivity ranking are almost identical (r = 0.6), with the strength of the correlation of R1 with R3 due to the single effect of R0-B.

8.5 CONCLUSIONS

A rank order analysis of the correlations between (a) water company membrane innovation capability (measured by a joint measure of installed plant capacity and numbers of membrane plants), (b) econometric parameters employed, by the UK regulators to measure and regulate water company performance, and (c) assessed water company receptivity to membrane technology was undertaken using a maximal variation case study sample of 5 different companies. The study aimed to determine whether the perceived failure of membrane technology innovation in UK water companies in terms of membrane uptake can be better explained by variation in regulatory econometric performance or by self-assessed receptivity to membrane technologies, using a set of measures derived from models of innovation which emphasise knowledge, learning and materiality.

The study concludes the following:

1. Most of the regulatory econometrics studied provided a poor explanation for variation in water company membrane uptake, with the exception of return on capital employed (Econometric B, companies ranked using R0-B) which was highly correlated with membrane uptake (r = 0.9). A proportionally higher capital spend on a new technology in an organisation is linked with implementation efficiency and organisational change.

2. Overall the organisational elements of receptivity to membrane technology studied showed correlated moderately with membrane uptake (r = 0.6), equal in strength to the joint econometric measure (R1) assessed. However if R0-B is omitted from the analysis then the

correlation between receptivity and membrane uptake is stronger relatively as the joint econometric correlation with membrane uptake drops to r = -0.051.

3. The overriding contribution of R0-B to the strength of the econometric correlation with membrane uptake requires further investigation to understand and explain. It may be that water companies able to deliver high returns on capital (Econometric B) have more effective operational and learning processes which have enabled them to better adapt to and adopt membrane technology. This is a proposition that is supported by the r = 0.7 correlation between receptivity to membrane technology and return on capital employed, but requires further work to test more rigorously.

4. The weakness of the econometric correlations with membrane uptake when B is omitted from the analysis, compared to the receptivity correlation suggests a level of tacit implementation skill within water companies that is not easily measured and therefore performance regulated.

From the above it appears that efficiency of capital employment (in terms of % return) is an adequate proxy measure of innovative capability in UK water companies, but that the other econometrics studied are not good measures. This poses questions regarding the most effective regulation of water companies for the promotion and development of innovative capability whilst developing their ability to deliver higher returns on capital. It is also clear that further research is required to understand why return on capital should be a good proxy measure of innovation capability.

It is also by no means clear in this instance that socio-technical description of the organisation can account for differences between the companies, or that there are determinist properties of the technology that promote or assist innovation uptake.

For example Company 5 did not need to implement any membrane plants for this scheme yet perceived itself as an innovator through its pilot plant work, whereas Company 4 with a similar supply basis was critically dependent on membrane technology yet perceived less reliance on training but more on treatment drivers for successful uptake. The material factors affecting innovation reception are seemingly external to the technology in this heavily economically and quality regulated industry, and this may explain attitudes to innovation.

Given the legislative implementation challenges the sector faces to reduce carbon footprints and introduce catchment scale ecological management objectives, sector specific research is needed to help identify the appropriate metrics for innovation that will expedite decision making for implementing the necessary novel mix of hard and soft technologies.

These conclusions need to be disseminated to operational managers in order that they may devise appropriate measures for an innovations implementation and ensure such that the organisation can evaluate not just the technology but its own ongoing reflexivity.

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CHAPTER 9 CONCLUSIONS

9.1 CONCLUSIONS

This thesis has furthered research by demonstrating that the cleaning science for microfiltration and ultrafiltration membranes can be advanced using statistical tools. A number of conclusions can be drawn from this work:

1. ***It was concluded that there remains little research dedicated to developing a general science of cleaning.*** In reviewing the literature (Chapter 2) it can be shown that the focus for research is on fouling mechanisms and character of foulants. Work is confined to specific applications and membranes. The cleaning science remains empirical and ad-hoc but the research here into statistical factor modelling shows that quantitative research into measuring and monitoring the changeability of cleaning effectiveness on membranes is possible.

2. ***This study demonstrated that tests into a statistical cleaning performance measurement are fruitful.*** Autopsy and permeability recovery cleaning tests were carried out on MF/UF membranes from nine plants. The cleaning tests followed a project developed protocol that, by using Box Behenken response surface methodology, quantified recovery in terms of the cleaning factors (temperature, soak time and concentration). Membrane fibres from modules fouled on various sites were tested including the analysis of foulants. From the different factor relationships we see in figures 5.2 to 5.6, the variation from optimum points for cleaning were seen to be calculated. These describe the cleanability status of a particular membrane plant's fibres at the time of removal and can provide a benchmark for design or operational studies.

3. ***It was concluded that modified factorial analysis and specifically Box Behnken (BBD) Response Surface methods are adaptable and could be applied to study changes on a single site or look at cleaning across a number of plants.*** A case study of a site using these techniques on membranes fouled at different stages of commissioning was carried (Chapter three). This showed that the optimum cleaning conditions varied from the initial to later stages of commissioning. It was ascertained that biofouling

experienced at this site was tenacious and the analysis tool allowed recommendations to be made to the operator on adapting cleaning chemicals and conditions to suit. This paper demonstrated the BBD methods utility; allowing simple comparison of the variation in intra site cleaning techniques.

4. ***In testing the protocol method on an optimum recovery matrix for a submerged MBR (Chapter four), it was concluded that as well as morphological effects on fouling and cleaning, the fibres position in the flow stream influences cleanability.*** The position of MBR fibres taken from different positions in the reactor relative to the incoming flow stream had variable recoverability and it was also seen here that there is a possible link to membrane damage and integrity failure. This work emphasises the importance of flow distribution and illustrated how cleaning recovery quantification could be adapted to diagnose with other tools such as reactor design with finite element analysis.

5. ***It was concluded that methods for quantifying the factor variation in the degree of cleanability have utility for optimising costs.*** In Chapter five the experimental cleaning efficacy varied between and also across the range of fibres tested depending on their type and condition. Using the factor analysis method here consistently, a variety of membranes (MF-HF, SMF-HF, UF-CT), fouled on different full scale plants were subjected to a number of cleaning treatments. The algorithms based on the cleaning factors for these showed how recovery across the factor envelope for each plant varied. This has implications for designers and operators as the effects were also seen to vary as the membrane ages.

6. ***It was concluded that there are significant cost implications with cleanability variation.*** For different plants the effect of variance from the optimum cleaning regime was calculated by overlaying the cleaning factors envelope from the optimum and operational scenarios on a cost model based on specific fouling resistance. Chapter six modelled these effects showing how deviations from the optimum cleaning influences the transient TMP profile. We saw in this paper that this cost effect was significant and that

there are cost benefits from optimising the cleaning factors that vary depending on membrane material, configuration and plant design. Measurement of variance in cleanability response with the tools developed in this thesis could save up to £170,000 in operational costs over a plant's lifetime.

7. ***It was concluded that membrane dimensions influence cleaning efficacy and should be considered a key factor when specifying and testing cleaning regimes.*** It is clear from the range of membrane recovery responses that fibre dimensions played a significant role in fouling and cleaning. This was investigated here in Chapter 7 where a model for flow distribution along different fibre lengths and internal diameters was built. The model was verified in the laboratory, where it was seen that the headloss distribution of a surrogate foulant along various lengths of MF fibre matched the models prediction. The ratios for TMP to applied pressure, as well as the fibre lengths to total flowrate, were independent of headloss profiles and so initially the flux and TMP profile along the fibres is dependent on the configuration of the membrane. The cleanant is applied homogeneously to a heterogeneous and axially influenced fouling zone. This has implications not only for membrane plant engineering but also for engineering the optimum cleaning regime.

8. ***In the non technical research it was concluded from studying the implementation of membrane technology, that large state regulated privatised water utilities need to develop different approaches to measuring the skills needed to innovate and thus encourage inward technology transfer.*** In Chapter eight it was asked if economic regulation shaped water company behaviour towards innovation and membrane plant uptake. Econometrics based on narrow financial measures may be restricting the development of an innovation implementation culture. A number of water companies were studied and when perceived receptivity was compared to regulatory econometrics the strongest factor was return on capital invested. Technology may be being adopted for mandatory and economic suitability instead of integration and development of the organisations tacit skills and

this is problematic. Regulators are seeking to improve water company performance to suit legislation such as the Water Framework Directive; where a combination of hard and soft investment is needed.

9.2 FURTHER WORK

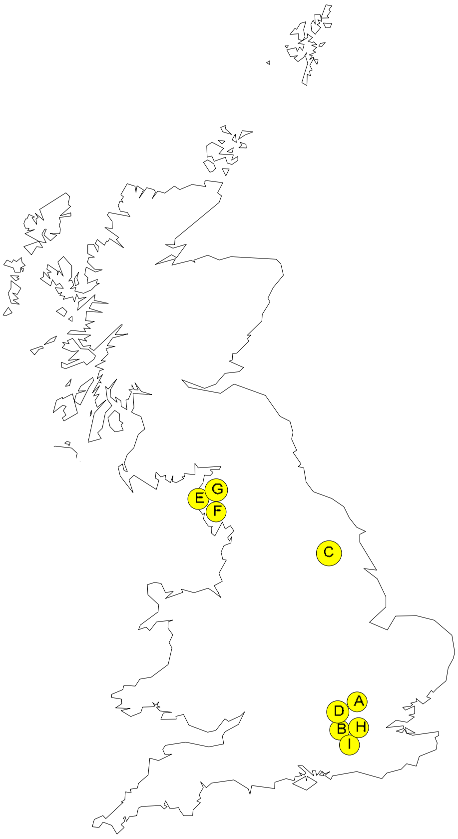
This work helps open up areas where future work can be of benefit:

1. Qualitative studies of cleaning should also seek to quantify the cleaning envelope response by working towards standardised factor algorithms for cleaning. This will help establish baselines for new plants and give an index similar to the proposed Uniform Membrane Fouling Index (UMFI). It is suggested that a Uniform Cleaning Envelope Response Algorithm (UCERA) procedure be developed that could be used in conjunction with raw water characterisation and long term monitoring of the plant TMP transient data to prescribe intuitive rather than reactive changes to CIP and maintenance cleaning regimes.
2. Membrane plants often utilise pilot studies for short periods prior to design. The incorporation of cleaning factor envelope trials on these should is seen here as an essential component of plant design. The benefit of the addition of autopsy work on pilot plant modules to compare the changing response of the fibres to fouling and cleaning regimes over time should be evident from this work.
3. Building a body of knowledge based on cleaning factor envelope studies could be incorporated with work by others into backwash and cleaning scheduling and control modelling. It has been shown here that TMP transient and resistance development are linked to cleaning response and there are economical advantages to lowering operating costs based on headloss minimisation. Optimising cleaning could be a small but necessary part of water companies' strategies to reduce carbon emissions. Studies into the marginal cost benefit of this across an organisation need to be undertaken.
4. Finally, in order to reap benefits from new technology, water companies need to manage change by developing measures to monitor inward technology transfer. The procurement of relatively new technologies such as membranes are ideal for studying the incorporation of hard and soft technology changes. A strategic objective would be to enable a learning organisation to develop where innovation is organisationally “embrained” (where an organisation

relies on the conceptual skills of it's staff to develop knowledge for itself) rather than "enforced" where learning and knowledge work processes are simply embedded in work organised as tasks.

APPENDICES

A SAMPLED PLANT LOCATION AND TYPE



	Plant Name	Location	Water Source	Type	Config	Model
A	N.Mymms	Herts. SE England	Groundwater - borehole	UF	CT	NORIT X-FLOW -XIGA- UFC M5 in S225FSFC PVC 0418
B	Dorney II	Berks. SE England	Groundwater - boreholes	MF	HF	MEMCOR CMF-1-6-112M10V
C	Kepwick	North Yorkshire	Groundwater - Borehole and multiple spring	UF	CT	NORIT X-FLOW -XIGA- UFC M5 in S225FSFC PVC 0418
D	Clay Lane	Bucks, SE England	Groundwater Borehole	UF	CT	NORIT X-FLOW -XIGA- UFC M5 in S225FSFC PVC 0418
E	Ennerdale	Cumbria, NE Lakes	Upland surface	MF	HF	MEMCOR M10 CMF-C
F	Paddy End	Cumbria, NE Lakes	Upland surface	MF	CT	MEMCOR CMF-L10V
G	Cornhow	Cumbria, NE Lakes	Upland surface	MF	HF	MEMCOR M10 CMF-C
H	Batchworth	Herts, SE England	Groundwater - Borehole	MF	HF	MEMCOR CMF-S 1x5x360 S10V
I	Chertsey	Surrey SE England	Groundwater	MF	HF	MEMCOR CMF-S 1x4x90 S10V

B PLANTS SAMPLED: MEMBRANE AND PROCESS INFORMATION

SYSTEM INFO		A
	Plant Name	N.Mymms WTW
	Type	UF
	Donor	3V
	System	Pressurised
	Model	NORIT X-FLOW -XIGA- UFC M5 in S225FSFC PVC 0418
	Type	CT
	Material	PES
	Length	1.5275
	Pot length	0.0675
	Fibre #	10000
	Id mm	0.8
	Od mm	1.75
	Pore Size Nominal (micron)	500 kDa
	MWCO (k Da)	200
	Area/module m2	35
	Membrane Charge (pH 7.0)	
	Membrane Material Character	More Hydrophilic than PP
	Membrane symmetry	Assymetric
	Membrane structure	Microporous
	OperatingpH Range	2 to 12
	Maximum TMP (kPa)	-300 to +300
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	80
	Oxidant Tolerance (HOCL)	200 ppm max. at 40 °C max. and max. 30 minutes per day
	Oxidant Tolerance (H2O2)	100-200 ppm at 40 °C max.
	Acid Tolerance (HCl)	pH > = 1
	Acid Tolerance (Citric)	2 wt%
	AlkaliTolerance (Na OH)	with EDTA = pH < = 12 1%wt
	Water Source	Groundwater - borehole
RAW WATER PARAMETERS	Turbidity (NTU)	5-25NTU
	Colour (Hazen)	
	DOC (mg/l)	
	UV254 (au)	Upt 75Abs/m
	pH	7.1
	Temp Range (degC)	10 to 11
	CaCo3 Alkalinity (mg/l)	
PRETREATMENT	Processes	Clarification (Source only), GAC & Pre-cl
	Coagulant & dose	PACL up to 4mg/l as Al
	pH	7.1
	Turbidity	Up to 5NTU
	Colour	
	DOC	
	UV 254	75Abs/m
TREATED QUALITY PEFORMANCE	Turbidity	0.05NTU
	Colour	
	DOC	
	UV254	
BACKWASH	Backwash frequency	40-80mins
	Backwash frequency (norm. interval mins)	55mins
	Backwash rate	250lmh
	Backwash Duration	50secs
CEB 1	CEB 1 type ,strength, pH	Caustic 750mg/l pH12
	Interval	18 hours
	Duration	20MINS
	Temperature of chemicals	11
CEB 2	CEB 2 type ,strength, pH	HCL 1200MG/L Ph2
	Interval	18 hours
	Duration	20mins
	Temperature of chemicals	11
CIP 1	CIP 1 type ,strength, pH	Citric pH2.5
	Interval	
	Duration	
	Temperature of chemicals	
CIP 2	CIP 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 3	CIP 3 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	36Ml/d
	Actual Operational Flow (tcm3/d)	32Ml/d
	Number of Modules in normal service	504
	Normal Operational Flux J (LMH)	109
	Operating Pressure (kPa)	
	Permeability, Kn (LMH/Bar)	250-320
	Avg. Operational TMP (kPa)	0.4bar
	Plant Rate of TMP decline (kPa/hr)	300- 180 in 8-12hours
LONGEVITY	Date Plant Commissioned	01/09/2002
	Date Membranes Commissioned	01/07/2002
	Date Membrane Sampled	01/02/2005
	Time membranes in operation (days)	946

SYSTEM INFO		B
	Plant Name	Dorney II WTW (Taplow bh E&G)
	Type	MF
	Donor	TWL
	System	Pressurised
	Model	MEMCOR CMF-1-6-112M10V
	Type	HF
	Material	PVDF
	Length	1.157
	Pot length	0.094
	Fibre #	9600
	Id mm	0.5
	Od mm	0.8
	Pore Size Nominal (micron)	
	MWCO (k Da)	
	Area/module m2	28.7
	Membrane Charge (pH 7.0)	
	Membrane Material Character	More Hydrophilic than PES
	Membrane symmetry	Asymmetric
	Membrane structure	Anisotropic
	OperatingpH Range	
	Maximum TMP (kPa)	
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	
	Oxidant Tolerance (HOCL)	
	Oxidant Tolerance (H2O2)	
	Acid Tolerance (HCl)	
	Acid Tolerance (Citric)	
	AlkaliTolerance (Na OH)	
	Water Source	Groundwater - boreholes
RAW WATER PARAMETERS	Turbidity (NTU)	0.25 < 1.4 - < 30
	Colour (Hazen)	1.0 < 4.1 < 7.0
	DOC (mg/l)	1.0 < 2.1 < 3.1
	UV254 (au)	
	pH	
	Temp Range (degC)	5 < 11.8 < 19
	CaCo3 Alkalinity (mg/l)	233 < 237 < 240
PRETREATMENT	Processes	Micro Strainer 500um
	Coagulant & dose	No coag, but some pre membrane chlorination
	pH	N/A
	Turbidity	>0.3
	Colour	N/A
	DOC	N/A
	UV 254	N/A
TREATED QUALITY PEFORMANCE	Turbidity	
	Colour	
	DOC	
	UV254	
BACKWASH	Backwash frequency	35 to 60 based on raw turbidity
	Backwash frequency (norm. interval mins)	60
	Backwash rate	
	Backwash Duration	
CEB 1	CEB 1 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	H2SO4 pH2.0 300ppm
	Interval	Every 28 days
	Duration	100 minutes
	Temperature of chemicals	15 degC
CIP 2	CIP 2 type ,strength, pH	NaOCl 0.05 %
	Interval	Every 28 days after H2SO4
	Duration	100 mins
	Temperature of chemicals	30 deg C
CIP 3	CIP 3 type ,strength, pH	Citric Acid 0.5%
	Interval	Manual every 180 days
	Duration	100 mins
	Temperature of chemicals	30 deg C
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	27 Ml/d
	Actual Operational Flow (tcm3/d)	Turb dependent
	Number of Modules in normal service	672
	Normal Operational Flux J (LMH)	71.55
	Operating Pressure (kPa)	400
	Permeability, Kn (LMH/Bar)	120
	Avg. Operational TMP (kPa)	0.6
	Plant Rate of TMP decline (kPa/hr)	3.5 to 5.5 resistance
LONGEVITY	Date Plant Commissioned	01/03/2005
	Date Membranes Commissioned	01/04/2005
	Date Membrane Sampled	07/06/2006
	Time membranes in operation (days)	432

SYSTEM INFO		C
	Plant Name	Kepwick WTW
	Type	UF
	Donor	YW
	System	Pressurised
	Model	NORIT X-FLOW -XIGA- UFC M5 in S225FSFC PVC 0418
	Type	CT
	Material	PES
	Length	1.5275
	Pot length	0.0675
	Fibre #	10000
	Id mm	0.8
	Od mm	1.75
	Pore Size Nominal (micron)	500 kDa
	MWCO (k Da)	200
	Area/module m2	35
	Membrane Charge (pH 7.0)	
	Membrane Material Character	More Hydrophilic than PP
	Membrane symmetry	Assymetric
	Membrane structure	Microporous
	OperatingpH Range	2 to 12
	Maximum TMP (kPa)	-300 to +300
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	80
	Oxidant Tolerance (HOCL)	200 ppm max. at 40 °C max. and max. 30 minutes per day
	Oxidant Tolerance (H2O2)	100-200 ppm at 40 °C max.
	Acid Tolerance (HCl)	pH > = 1
	Acid Tolerance (Citric)	2 wt%
	AlkaliTolerance (Na OH)	with EDTA = pH < = 12 1%wt
	Water Source	Groundwater - B-hole and multiple spring
RAW WATER PARAMETERS	Turbidity (NTU)	Max 1.1 , Min 0.02, Av 0.17
	Colour (Hazen)	Max 15, Min 0.7, Av 1.15
	DOC (mg/l)	Max 61.1 , Min 0, Av 6.13
	UV254 (au)	Max 114, Min 0, Av 7.0
	pH	Max 8.4 , Min 7.2, Av 7.8
	Temp Range (degC)	est 8 deg C
	CaCo3 Alkalinity (mg/l)	Max 869, Min 195, Av 250
PRETREATMENT	Processes	Mix of spring water from two discrete springs.
	Coagulant & dose	No coag, but some pre membrane chlorination
	pH	Same as raw
	Turbidity	Same as raw
	Colour	Same as raw
	DOC	Not available
	UV 254	Not available
TREATED QUALITY PEFORMANCE	Turbidity	Max 0.6, Min 0, Avg 0.1
	Colour	Max 7.5, Min 0.4, Avg 0.8
	DOC	Max 47.3, Min 0.0, Avg 2.5
	UV254	Max 637, Min 202, Avg 252
BACKWASH	Backwash frequency	Steps 0 - 5 ~ turbidity
	Backwash frequency (norm. interval mins)	Approx 1 hr, dependent on TMP build up.
	Backwash rate	75 m3 / hr
	Backwash Duration	1 min
CEB 1	CEB 1 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	Caustic 47% to pH 13
	Interval	Unable to determine - EST MONTHLY
	Duration	Soak at pH 13 for 1 hr
	Temperature of chemicals	Room temp - 15degC
CIP 2	CIP 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 3	CIP 3 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	3.4
	Actual Operational Flow (tcm3/d)	2.4
	Number of Modules in normal service	36
	Normal Operational Flux J (LMH)	75
	Operating Pressure (kPa)	0.8 - 1.2 bar
	Permeability, Kn (LMH/Bar)	From 88 to 70 over 17 hrs
	Avg. Operational TMP (kPa)	0.6 (bar)
	Plant Rate of TMP decline (kPa/hr)	
LONGEVITY	Date Plant Commissioned	01/06/2002
	Date Membranes Commissioned	01/07/2002
	Date Membrane Sampled	28/02/2007
	Time membranes in operation (days)	1703

SYSTEM INFO		D
	Plant Name	Clay Lane WTW
	Type	UF
	Donor	3V
	System	Pressurised
	Model	NORIT X-FLOW -XIGA- UFC M5 in S225FSFC PVC 0418
	Type	CT
	Material	PES
	Length	1.5275
	Pot length	0.0675
	Fibre #	10000
	Id mm	0.8
	Od mm	1.75
	Pore Size Nominal (micron)	500 kDa
	MWCO (k Da)	200
	Area/module m2	35
	Membrane Charge (pH 7.0)	
	Membrane Material Character	More Hydrophilic than PP
	Membrane symmetry	Assymetric
	Membrane structure	Microporous
	OperatingpH Range	2 to 12
	Maximum TMP (kPa)	-300 to +300
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	80
	Oxidant Tolerance (HOCL)	200 ppm max. at 40 °C max. and max. 30 minutes per day
	Oxidant Tolerance (H2O2)	100-200 ppm at 40 °C max.
	Acid Tolerance (HCl)	pH > = 1
	Acid Tolerance (Citric)	2 wt%
	AlkaliTolerance (Na OH)	with EDTA = pH < = 12 1%wt
	Water Source	Groundwater Borehole
RAW WATER PARAMETERS	Turbidity (NTU)	
	Colour (Hazen)	
	DOC (mg/l)	
	UV254 (au)	
	pH	7.1
	Temp Range (degC)	10 to 11
	CaCo3 Alkalinity (mg/l)	
PRETREATMENT	Processes	Ozone, GAC and chlorination (0.4-0.6mg/l)
	Coagulant & dose	n/a
	pH	7.1
	Turbidity	
	Colour	
	DOC	
	UV 254	
TREATED QUALITY PEFORMANCE	Turbidity	0.05NTU
	Colour	
	DOC	
	UV254	
BACKWASH	Backwash frequency	60-150mins
	Backwash frequency (norm. interval mins)	120mins
	Backwash rate	250lmh
	Backwash Duration	50secs
CEB 1	CEB 1 type ,strength, pH	Caustic 700mg/l to pH12.
	Interval	18-48 hours
	Duration	20mins
	Temperature of chemicals	11
CEB 2	CEB 2 type ,strength, pH	HCl 1200mg/l to pH2
	Interval	36-120 hours
	Duration	20 mins
	Temperature of chemicals	11
CIP 1	CIP 1 type ,strength, pH	Citric pH 2.5
	Interval	Ad hoc or every 6 months
	Duration	24 hours
	Temperature of chemicals	11
CIP 2	CIP 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 3	CIP 3 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	165Ml/dy
	Actual Operational Flow (tcm3/d)	Avg 125Ml/dy/yr
	Number of Modules in normal service	1532
	Normal Operational Flux J (LMH)	125
	Operating Pressure (kPa)	0.4bar
	Permeability, Kn (LMH/Bar)	300-330
	Avg. Operational TMP (kPa)	0.4bar
	Plant Rate of TMP decline (kPa/hr)	3 to 4 hrs cycles so 10 kPa per hour
LONGEVITY	Date Plant Commissioned	01/06/2001
	Date Membranes Commissioned	01/07/2001
	Date Membrane Sampled	01/01/2007
	Time membranes in operation (days)	2010

SYSTEM INFO		E
	Plant Name	Ennerdale WTW
	Type	MF
	Donor	UU
	System	Pressurised
	Model	MEMCOR M10 CMF-C
	Type	HF
	Material	PP
	Length	1.157
	Pot length	0.094
	Fibre #	20000
	Id mm	0.25
	Od mm	0.55
	Pore Size Nominal (micron)	0.2
	MWCO (k Da)	n/a
	Area/module m2	33.52
	Membrane Charge (pH 7.0)	Negative
	Membrane Material Character	Hydrophobic
	Membrane symmetry	
	Membrane structure	
	OperatingpH Range	2 to 13
	Maximum TMP (kPa)	
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	
	Oxidant Tolerance (HOCL)	No oxidants
	Oxidant Tolerance (H2O2)	No oxidants
	Acid Tolerance (HCl)	No oxidants
	Acid Tolerance (Citric)	No oxidants
	AlkaliTolerance (Na OH)	No oxidants
	Water Source	Upland surface
RAW WATER PARAMETERS	Turbidity (NTU)	2.44, <0.23, 0.71 (max, min, mean)
	Colour (Hazen)	8.0, 2.0, 4.28
	DOC (mg/l)	TOC results only 1.59, <0.25, 1.11
	UV254 (au)	Chlorophyll mg/l results only 8.67, 5.1, 7.17
	pH	7.9, 6.4, 6.9
	Temp Range (degC)	18.4, 3.2, 10.4
	CaCo3 Alkalinity (mg/l)	52.3*, <1.22, 3.58 (*dubious result. All other values <10)
PRETREATMENT	Processes	Microstrained water to 30 micron
	Coagulant & dose	N/A
	pH	N/A
	Turbidity	N/A
	Colour	N/A
	DOC	N/A
	UV 254	N/A
TREATED QUALITY PEFORMANCE	Turbidity	0.66, <0.01, 0.08
	Colour	3.75, <0.4, 1.32
	DOC	TOC results only 1.7, 0.7, 1.04
	UV254	N/A
BACKWASH	Backwash frequency	
	Backwash frequency (norm. interval mins)	Primary 60 mins, Secondary 30 mins, Dual 45 mins
	Backwash rate	N/A programmed in PLC
	Backwash Duration	N/A programmed in PLC
CEB 1	CEB 1 type ,strength, pH	N/A
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	N/A
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	Caustic 2.5%, 27ºC
	Interval	Every 4 Weeks primary, Weekly secondary
	Duration	30 minute soak plus other step times
	Temperature of chemicals	27 - 31 ºC
CIP 2	CIP 2 type ,strength, pH	Peroxide 5% in 2.5% caustic soln.
	Interval	As required (2x per wk in summer-Algae)
	Duration	30 min soak plus other step times
	Temperature of chemicals	27 - 31 ºC
CIP 3	CIP 3 type ,strength, pH	N/A Acid CIPs not carried out routinely
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	69 MLD 125m3/hr primary, 60m3/hr secondary
	Actual Operational Flow (tcm3/d)	65 MLD 115m3/hr primary, 60m3/hr secondary
	Number of Modules in normal service	1500
	Normal Operational Flux J (LMH)	68
	Operating Pressure (kPa)	1750 primary, 2000 secondary
	Permeability, Kn (LMH/Bar)	105 - 40 MLD =1.7 bar
	Avg. Operational TMP (kPa)	Approx 105 primary, 80 secondary
	Plant Rate of TMP decline (kPa/hr)	0.12 kPa/hr
LONGEVITY	Date Plant Commissioned	01/01/2000
	Date Membranes Commissioned	01/02/2000
	Date Membrane Sampled	01/03/2007
	Time membranes in operation (days)	2585

SYSTEM INFO		F1 PRE-COMM
	Plant Name	Paddy End primary I Pre Comm
	Type	MF
	Donor	UU
	System	Pressurised
	Model	MEMCOR CMF-L10V
	Type	CT
	Material	PVDF
	Length	1.157
	Pot length	0.094
	Fibre #	9600
	Id mm	0.5
	Od mm	0.8
	Pore Size Nominal (micron)	0.1
	MWCO (k Da)	Not known.
	Area/module m2	23.4
	Membrane Charge (pH 7.0)	Not known.
	Membrane Material Character	More Hydrophilic than PES
	Membrane symmetry	Asymmetric
	Membrane structure	Anisotropic
	OperatingpH Range	2 to 10
	Maximum TMP (kPa)	130
	Maximum permissible turbidity (NTU)	17.25 NTU
	Operating Maximum Temperature (degC)	Design temperatures: min = 5degC, average 12 degC, no maxi
	Oxidant Tolerance (HOCL)	ok
	Oxidant Tolerance (H2O2)	Not known.
	Acid Tolerance (HCl)	Assumed to be ok - acid washes permitted.
	Acid Tolerance (Citric)	Assumed to be ok - acid washes permitted.
	AlkaliTolerance (Na OH)	Not known.
	Water Source	Upland surface
RAW WATER PARAMETERS	Turbidity (NTU)	<2
	Colour (Hazen)	< 30 Hazen
	DOC (mg/l)	<2 mg/l
	UV254 (au)	Not known.
	pH	5.5 to 5.8
	Temp Range (degC)	5 to 15
	CaCo3 Alkalinity (mg/l)	<40
PRETREATMENT	Processes	pH correction, coagulation, primary pressure filtration, HOCL,
	Coagulant & dose	PACl (not currently used)
	pH	5.4 - 6.2 pre 1st stage pressure filters; 7.8 pre 2nd stage pressur
	Turbidity	Typically 0.1NTU on the outlet of 1st stage pressure filters
	Colour	Not monitored continuously but typically <5 deg Hazen
	DOC	Not monitored continuously but typically <2 mg/l
	UV 254	Not known
TREATED QUALITY PEFORMANCE	Turbidity	<0.1 ntu
	Colour	Not monitored continuously but typically <5 deg Hazen
	DOC	Not monitored continuously but typically <2 mg/l
	UV254	Not known
BACKWASH	Backwash frequency	Every 90 mins on primary membranes, 40 mins on secondary
	Backwash frequency (norm. interval mins)	As above
	Backwash rate	Low pressure air scour and air assisted drain down, backwash
	Backwash Duration	2 minutes
CEB 1	CEB 1 type ,strength, pH	NA
	Interval	NA
	Duration	NA
	Temperature of chemicals	NA
CEB 2	CEB 2 type ,strength, pH	NA
	Interval	NA
	Duration	NA
	Temperature of chemicals	NA
CIP 1	CIP 1 type ,strength, pH	10% Sodium Hypochlorite
	Interval	14 - 28 days; ratio 2 hypo cleans : 1 sulphuric acid clean
	Duration	1800 seconds soak, 1800 seconds recirc
	Temperature of chemicals	30 deg C (temp of CIP recirc water)
CIP 2	CIP 2 type ,strength, pH	30% w/w Sulphuric acid
	Interval	14 - 28 days, ratio as above
	Duration	1800 seconds soak, 1800 seconds recirc
	Temperature of chemicals	30 deg C (temp of CIP recirc water)
CIP 3	CIP 3 type ,strength, pH	0.5% Citric acid
	Interval	A Citric acid clean occurs every 6 sulphuric acid cleans
	Duration	1800 seconds soak, 1800 seconds recirc
	Temperature of chemicals	30 deg C (temp of CIP recirc water)
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	Primary membranes - 53.4m3/hr per unit; Secondary membran
	Actual Operational Flow (tcm3/d)	
	Number of Modules in normal service	120
	Normal Operational Flux J (LMH)	96
	Operating Pressure (kPa)	Feed pressure typically 100 - 140 kPa; filtrate pressure typicall
	Permeability, Kn (LMH/Bar)	214 at 0.45 bar
	Avg. Operational TMP (kPa)	Range 30 - 60 kPa (avg 45 kPa)
	Plant Rate of TMP decline (kPa/hr)	4 kPa / hr
LONGEVITY	Date Plant Commissioned	01/10/2007
	Date Membranes Commissioned	20/10/2007
	Date Membrane Sampled	07/11/2007
	Time membranes in operation (days)	18

SYSTEM INFO		F2 POST COMM
	Plant Name	Paddy End primary ii Post Comm
	Type	MF
	Donor	UU
	System	Pressurised
	Model	MEMCOR CMF-L10V
	Type	CT
	Material	PVDF
	Length	1.157
	Pot length	0.094
	Fibre #	9600
	Id mm	0.5
	Od mm	0.8
	Pore Size Nominal (micron)	0.1
	MWCO (k Da)	Not known.
	Area/module m2	23.4
	Membrane Charge (pH 7.0)	Not known.
	Membrane Material Character	More Hydrophilic than PES
	Membrane symmetry	Asymmetric
	Membrane structure	Anisotropic
	OperatingpH Range	2 to 10
	Maximum TMP (kPa)	130
	Maximum permissible turbidity (NTU)	17.25 NTU
	Operating Maximum Temperature (degC)	Design temperatures: min = 5degC, average 12 degC, no maxi
	Oxidant Tolerance (HOCL)	ok
	Oxidant Tolerance (H2O2)	Not known.
	Acid Tolerance (HCl)	Assumed to be ok - acid washes permitted.
	Acid Tolerance (Citric)	Assumed to be ok - acid washes permitted.
	AlkaliTolerance (Na OH)	Not known.
	Water Source	Upland surface
RAW WATER PARAMETERS	Turbidity (NTU)	<2
	Colour (Hazen)	< 30 Hazen
	DOC (mg/l)	<2 mg/l
	UV254 (au)	Not known.
	pH	5.5 to 5.8
	Temp Range (degC)	5 to 15
	CaCo3 Alkalinity (mg/l)	<40
PRETREATMENT	Processes	pH correction, coagulation, primary pressure filtration, HOCL,
	Coagulant & dose	PACl (not currently used)
	pH	5.4 - 6.2 pre 1st stage pressure filters; 7.8 pre 2nd stage pressu
	Turbidity	Typically 0.1NTU on the outlet of 1st stage pressure filters
	Colour	Not monitored continuously but typically <5 deg Hazen
	DOC	Not monitored continuously but typically <2 mg/l
	UV 254	Not known
TREATED QUALITY PEFORMANCE	Turbidity	<0.1 ntu
	Colour	Not monitored continuously but typically <5 deg Hazen
	DOC	Not monitored continuously but typically <2 mg/l
	UV254	Not known
BACKWASH	Backwash frequency	Every 90 mins on primary membranes, 40 mins on secondary
	Backwash frequency (norm. interval mins)	As above
	Backwash rate	Low pressure air scour and air assisted drain down, backwash
	Backwash Duration	2 minutes
CEB 1	CEB 1 type ,strength, pH	NA
	Interval	NA
	Duration	NA
	Temperature of chemicals	NA
CEB 2	CEB 2 type ,strength, pH	NA
	Interval	NA
	Duration	NA
	Temperature of chemicals	NA
CIP 1	CIP 1 type ,strength, pH	10% Sodium Hypochlorite
	Interval	14 - 28 days; ratio 2 hypo cleans : 1 sulphuric acid clean
	Duration	1800 seconds soak, 1800 seconds recirc
	Temperature of chemicals	30 deg C (temp of CIP recirc water)
CIP 2	CIP 2 type ,strength, pH	30% w/w Sulphuric acid
	Interval	14 - 28 days, ratio as above
	Duration	1800 seconds soak, 1800 seconds recirc
	Temperature of chemicals	30 deg C (temp of CIP recirc water)
CIP 3	CIP 3 type ,strength, pH	0.5% Citric acid
	Interval	A Citric acid clean occurs every 6 sulphuric acid cleans
	Duration	1800 seconds soak, 1800 seconds recirc
	Temperature of chemicals	30 deg C (temp of CIP recirc water)
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	Primary membranes - 53.4m3/hr per unit; Secondary membran
	Actual Operational Flow (tcm3/d)	Maximum design throughput not exceeded. Maximum plant fl
	Number of Modules in normal service	5 primary units are in service all the time- flow spreading to ac
	Normal Operational Flux J (LMH)	96
	Operating Pressure (kPa)	Feed pressure typically 100 - 140 kPa; filtrate pressure typicall
	Permeability, Kn (LMH/Bar)	214 at 0.45 bar
	Avg. Operational TMP (kPa)	Range 30 - 60 kPa (avg 45 kPa)
	Plant Rate of TMP decline (kPa/hr)	4 kPa / hr
LONGEVITY	Date Plant Commissioned	01/10/2007
	Date Membranes Commissioned	14/10/2007
	Date Membrane Sampled	20/01/2008
	Time membranes in operation (days)	98

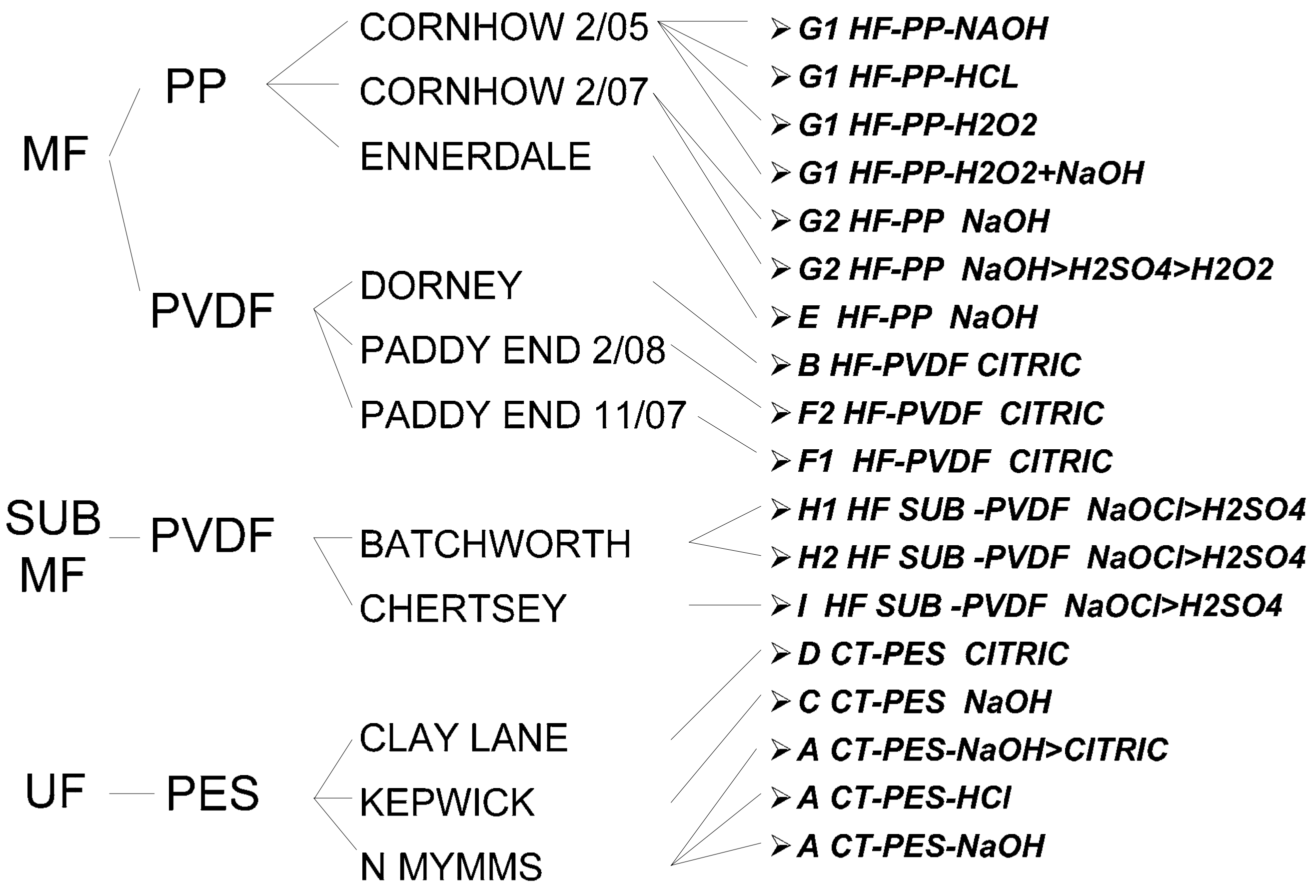
SYSTEM INFO		**G 2005**
	Plant Name	Cornhow WTW I 05
	Type	MF
	Donor	UU
	System	Pressurised
	Model	MEMCOR M10 CMF-C
	Type	HF
	Material	PP
	Length	1.157
	Pot length	0.094
	Fibre #	5000
	Id mm	0.25
	Od mm	0.55
	Pore Size Nominal (micron)	0.2
	MWCO (k Da)	n/a
	Area/module m2	33.52
	Membrane Charge (pH 7.0)	Negative
	Membrane Material Character	Hydrophobic
	Membrane symmetry	Assymetric
	Membrane structure	Anisotropic
	OperatingpH Range	2 to 13
	Maximum TMP (kPa)	200
	Maximum permissible turbidity (NTU)	500
	Operating Maximum Temperature (degC)	
	Oxidant Tolerance (HOCL)	No oxidants
	Oxidant Tolerance (H2O2)	No oxidants
	Acid Tolerance (HCl)	
	Acid Tolerance (Citric)	
	AlkaliTolerance (Na OH)	
	Water Source	Upland surface
RAW WATER PARAMETERS	Turbidity (NTU)	1.51, <0.23, 0.60 (max, min, mean)
	Colour (Hazen)	8.4, 3.0, 5.1
	DOC (mg/l)	TOC results only 1.77, 0.93, 1.23
	UV254 (au)	N/A
	pH	7.56, 6.27, 6.8
	Temp Range (degC)	17.5, 4.0, 10.5
	CaCo3 Alkalinity (mg/l)	13.4, <1.22, 3.64
PRETREATMENT	Processes	Microstrained water to 30 micron
	Coagulant & dose	N/A
	pH	N/A
	Turbidity	N/A
	Colour	N/A
	DOC	N/A
	UV 254	N/A
TREATED QUALITY PEFORMANCE	Turbidity	0.8, <0.01, 0.08
	Colour	4.0, <0.4, 1.76
	DOC	TOC results only 1.61, 0.86, 1.19
	UV254	N/A
BACKWASH	Backwash frequency	
	Backwash frequency (norm. interval mins)	Primary 60 mins, Secondary 30 mins, Dual 45 mins
	Backwash rate	N/A programmed in PLC
	Backwash Duration	N/A programmed in PLC
CEB 1	CEB 1 type ,strength, pH	N/A
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	N/A
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	Caustic 1-1.2%, 32ºC
	Interval	Every 2 Weeks primary and secondary
	Duration	30 minute soak plus other step times
	Temperature of chemicals	27 - 31ºC
CIP 2	CIP 2 type ,strength, pH	Peroxide 5% in 1-1.2% caustic soln. (50% frequency of CIP)
	Interval	As required (1 in every 4 CIP 1 s)
	Duration	30 min soak plus other step times
	Temperature of chemicals	27 - 31 ºC
CIP 3	CIP 3 type ,strength, pH	N/A Acid CIPs not carried out routinely
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	125m3/hr primary, 100m3/hr secondary
	Actual Operational Flow (tcm3/d)	125m3/hr primary, 100m3/hr secondary
	Number of Modules in normal service	90
	Normal Operational Flux J (LMH)	41 primary, 33 secondary
	Operating Pressure (kPa)	2000
	Permeability, Kn (LMH/Bar)	60 - 120 TMP
	Avg. Operational TMP (kPa)	Approx 65 primary, 60 secondary
	Plant Rate of TMP decline (kPa/hr)	0.07 kPa/hr
LONGEVITY	Date Plant Commissioned	01/05/2000
	Date Membranes Commissioned	01/06/2004
	Date Membrane Sampled	01/02/2005
	Time membranes in operation (days)	245

SYSTEM INFO		G 2007
	Plant Name	Cornhow WTW ii 07
	Type	MF
	Donor	UU
	System	Pressurised
	Model	MEMCOR CMF-M10V
	Type	HF
	Material	PP
	Length	1.157
	Pot length	0.094
	Fibre #	5000
	Id mm	0.25
	Od mm	0.55
	Pore Size Nominal (micron)	0.2
	MWCO (k Da)	n/a
	Area/module m2	33.52
	Membrane Charge (pH 7.0)	Negative
	Membrane Material Character	Hydrophobic
	Membrane symmetry	Assymetric
	Membrane structure	Anisotropic
	OperatingpH Range	2 to 13
	Maximum TMP (kPa)	200
	Maximum permissible turbidity (NTU)	500
	Operating Maximum Temperature (degC)	
	Oxidant Tolerance (HOCL)	No oxidants
	Oxidant Tolerance (H2O2)	No oxidants
	Acid Tolerance (HCl)	
	Acid Tolerance (Citric)	
	AlkaliTolerance (Na OH)	
	Water Source	Upland surface
RAW WATER PARAMETERS	Turbidity (NTU)	1.51, <0.23, 0.60 (max, min, mean)
	Colour (Hazen)	8.4, 3.0, 5.1
	DOC (mg/l)	TOC results only 1.77, 0.93, 1.23
	UV254 (au)	N/A
	pH	7.56, 6.27, 6.8
	Temp Range (degC)	17.5, 4.0, 10.5
	CaCo3 Alkalinity (mg/l)	13.4, <1.22, 3.64
PRETREATMENT	Processes	Microstrained water to 30 micron
	Coagulant & dose	N/A
	pH	N/A
	Turbidity	N/A
	Colour	N/A
	DOC	N/A
	UV 254	N/A
TREATED QUALITY PEFORMANCE	Turbidity	0.8, <0.01, 0.08
	Colour	4.0, <0.4, 1.76
	DOC	TOC results only 1.61, 0.86, 1.19
	UV254	N/A
BACKWASH	Backwash frequency	
	Backwash frequency (norm. interval mins)	Primary 60 mins, Secondary 30 mins, Dual 45 mins
	Backwash rate	N/A programmed in PLC
	Backwash Duration	N/A programmed in PLC
CEB 1	CEB 1 type ,strength, pH	N/A
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	N/A
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	Caustic 1-1.2%, 32ºC
	Interval	Every 2 Weeks primary and secondary
	Duration	30 minute soak plus other step times
	Temperature of chemicals	27 - 31ºC
CIP 2	CIP 2 type ,strength, pH	Peroxide 5% in 1-1.2% caustic soln. (50% frequency of CIP)
	Interval	As required
	Duration	30 min soak plus other step times
	Temperature of chemicals	27 - 31 ºC
CIP 3	CIP 3 type ,strength, pH	N/A Acid CIPs not carried out routinely
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	125m3/hr primary, 100m3/hr secondary
	Actual Operational Flow (tcm3/d)	125m3/hr primary, 100m3/hr secondary
	Number of Modules in normal service	90
	Normal Operational Flux J (LMH)	41 primary, 33 secondary
	Operating Pressure (kPa)	2000
	Permeability, Kn (LMH/Bar)	
	Avg. Operational TMP (kPa)	Approx 65 primary, 60 secondary
	Plant Rate of TMP decline (kPa/hr)	0.07 kPa/hr
LONGEVITY	Date Plant Commissioned	01/05/2000
	Date Membranes Commissioned	01/06/2004
	Date Membrane Sampled	01/03/2007
	Time membranes in operation (days)	1003

SYSTEM INFO		H
	Plant Name	Batchworth
	Type	MF
	Donor	3V
	System	Submerged
	Model	Memcor CMF-S 1x5x360 S10V
	Type	HF
	Material	PVDF
	Length	1.238
	Pot length	0.094
	Fibre #	14500
	Id mm	0.39
	Od mm	0.65
	Pore Size Nominal (micron)	0.04
	MWCO (k Da)	n/a
	Area/module m2	31.09
	Membrane Charge (pH 7.0)	Neutral
	Membrane Material Character	More Hydrophilic than PES
	Membrane symmetry	Assymetric
	Membrane structure	Anisotropic
	OperatingpH Range	2 to 10
	Maximum TMP (kPa)	120
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	45
	Oxidant Tolerance (HOCL)	1000 ppm max. and total on life length 1 000 000 ppm.hours
	Oxidant Tolerance (H2O2)	
	Acid Tolerance (HCl)	0.08 % w/v pH2
	Acid Tolerance (Citric)	
	AlkaliTolerance (Na OH)	NA (PVDF)
	Water Source	Groundwater - Borehole
RAW WATER PARAMETERS	Turbidity (NTU)	Usually < 0.2
	Colour (Hazen)	
	DOC (mg/l)	
	UV254 (au)	
	pH	7-7.2
	Temp Range (degC)	10 -20
	CaCo3 Alkalinity (mg/l)	300 - 330 as HCO3
PRETREATMENT	Processes	
	Coagulant & dose	
	pH	
	Turbidity	
	Colour	
	DOC	
	UV 254	
TREATED QUALITY PEFORMANCE	Turbidity	
	Colour	
	DOC	
	UV254	
BACKWASH	Backwash frequency	60 minutes
	Backwash frequency (norm. interval mins)	60
	Backwash rate	60 L/m2h
	Backwash Duration	20 s
CEB 1	CEB 1 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	Sodium hypochlorite, 300-400 ppm
	Interval	14 or 28 days
	Duration	1/2 h recirculation, 1 h soak
	Temperature of chemicals	28-30 C (in CIP water tank)
CIP 2	CIP 2 type ,strength, pH	Sulphuric acid, pH 2
	Interval	14 or 28 days
	Duration	1/2 h recirculation, 1 h soak
	Temperature of chemicals	No setting
CIP 3	CIP 3 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	38
	Actual Operational Flow (tcm3/d)	
	Number of Modules in normal service	360
	Normal Operational Flux J (LMH)	160
	Operating Pressure (kPa)	250
	Permeability, Kn (LMH/Bar)	266
	Avg. Operational TMP (kPa)	0.6
	Plant Rate of TMP decline (kPa/hr)	0.1
LONGEVITY	Date Plant Commissioned	01/11/2003
	Date Membranes Commissioned	01/11/2003
	Date Membrane Sampled	01/02/2008
	Time membranes in operation (days)	1553

SYSTEM INFO		I
	Plant Name	Chertsey
	Type	MF
	Donor	3V
	System	Submerged
	Model	Memcor CMF-S 1x4x90 S10V
	Type	HF
	Material	PVDF
	Length	1.238
	Pot length	0.094
	Fibre #	14500
	Id mm	0.39
	Od mm	0.65
	Pore Size Nominal (micron)	0.04
	MWCO (k Da)	n/a
	Area/module m2	31.09
	Membrane Charge (pH 7.0)	Neutral
	Membrane Material Character	More Hydrophilic than PES
	Membrane symmetry	Assymetric
	Membrane structure	Anisotropic
	OperatingpH Range	2 to 10
	Maximum TMP (kPa)	120
	Maximum permissible turbidity (NTU)	
	Operating Maximum Temperature (degC)	45
	Oxidant Tolerance (HOCL)	1000 ppm max. and total on life length 1 000 000 ppm.hours
	Oxidant Tolerance (H2O2)	
	Acid Tolerance (HCl)	0.08 % w/v pH2
	Acid Tolerance (Citric)	
	AlkaliTolerance (Na OH)	NA (PVDF)
	Water Source	Groundwater
RAW WATER PARAMETERS	Turbidity (NTU)	0.1 - 0.5
	Colour (Hazen)	
	DOC (mg/l)	
	UV254 (au)	
	pH	7.3
	Temp Range (degC)	12 - 15 oC
	CaCo3 Alkalinity (mg/l)	
PRETREATMENT	Processes	Pre - chlorination
	Coagulant & dose	
	pH	7.3
	Turbidity	0.1 - 0.5
	Colour	
	DOC	
	UV 254	
TREATED QUALITY PEFORMANCE	Turbidity	0.03
	Colour	
	DOC	
	UV254	
BACKWASH	Backwash frequency	Each cell will wash every 60 mins
	Backwash frequency (norm. interval mins)	
	Backwash rate	
	Backwash Duration	
CEB 1	CEB 1 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CEB 2	CEB 2 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 1	CIP 1 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
CIP 2	CIP 2 type ,strength, pH	Sulphuric acid, pH 2
	Interval	14 or 28 days
	Duration	1/2 h recirculation, 1 h soak
	Temperature of chemicals	No setting
CIP 3	CIP 3 type ,strength, pH	
	Interval	
	Duration	
	Temperature of chemicals	
HYDRAULIC PERFORMANCE	Design throughput, Q (tcm3/d)	41
	Actual Operational Flow (tcm3/d)	
	Number of Modules in normal service	450
	Normal Operational Flux J (LMH)	122
	Operating Pressure (kPa)	250
	Permeability, Kn (LMH/Bar)	205
	Avg. Operational TMP (kPa)	0.6
	Plant Rate of TMP decline (kPa/hr)	0.1
LONGEVITY	Date Plant Commissioned	01/05/2003
	Date Membranes Commissioned	01/05/2003
	Date Membrane Sampled	01/05/2008
	Time membranes in operation (days)	1827

C SAMPLED PLANTS EXPERIMENTAL RESULTS: PERMEABILITY RECOVERY EXPERIMENTS.



APPENDIX C 1 : Organisational chart showing experiment classification.

From left to right the three classes of membrane divide into material types. Then the plants are subdivided by membrane module supplied with date. Finally, experimental runs were performed on each membrane and the type of membrane (CT/HF) and chemical cleanant are given.

D MATHEMATICAL MODEL PROGRAMME LISTING

```
100  CLS
101  OPEN "data4.dat" FOR OUTPUT AS #1
102  OPEN "plot3a.dat" FOR OUTPUT AS #2
103  '__________________________

104  k = .000000464   'cm/s    Hydraulic conductivity coefficient
105  od = .065         'cm     Fibre outer diameter
106  id = .039         'cm     Fibre inner diameter
107  ds = 1            'cm     computational segment length
108  m = .001          'pa s   Viscosity
109  fbrl = 100        'cm     Fibre length
110  lpot = 3.5        'cm     Module pot length
111  hap = 2           'm      Applied head
112  nofbre = 10       '       Number of fibres under consideration
113  noffipm = 20000   '       Number of fibres per module
114  '________________________
115  hap = hap * 100
116  moda = 33.5
117  m = .001 / (1000 * 9.81) * 100
118  pi = 3.1415
119  noseg = fbrl / ds
120  DIM q(noseg), d(noseg, noseg + 1), un(noseg + 1)
121  DIM tmp(noseg), xg(noseg)
122  c1 = 2 * pi * ds * k * hap / (LOG(od / id))
123  c2 = 16 * ds * k * m * lpot / ((LOG(od / id)) * (id / 2) ^ 4)
124  c3 = 16 * ds ^ 2 * k * m / ((LOG(od / id)) * (id / 2) ^ 4)
125  FOR i = 1 TO noseg
126  FOR j = 1 TO noseg
127  IF i = 1 AND j = 1 THEN d(i, j) = c2 + 1: GOTO 100
128  IF i = j THEN d(i, i) = c2 + (i - 1) * c3 + 1
129  IF i > j THEN d(i, j) = d(j, j) + .5 * c3 - 1
130  IF j > i THEN d(i, j) = d(i, i) - 1
131  NEXT j, i
132  rhs = -c1:   GOSUB gauss
133  FOR i = 1 TO noseg
134  tmp(i) = q(i) *(2 * pi * ds * k) * LOG(od / id)
135  qtot = qtot + q(i)
136  NEXT
137  hlt = hap - tmp(1)
138  GOSUB printing
139  END
140  gauss:
141  unknown = noseg
142  np1 = unknown + 1
143  nm1 = unknown - 1
144  FOR ii = 1 TO unknown
```

```
145  un(ii) = ii
146  NEXT
147  np1 = unknown + 1
148  nm1 = unknown - 1
149  FOR ii = 1 TO unknown
150  d(ii, np1) = -rhs
151  NEXT ii
152  FOR h = 1 TO nm1
153  hp1 = h + 1
117   FOR ii = hp1 TO unknown
154  fa = d(ii, h) / d(h, h)
155  FOR j = hp1 TO np1
156  d(ii, j) = d(ii, j) - fa * (d(h, j))
157  NEXT j, ii, h
158  xg(un(unknown)) = d(unknown, np1) / d(unknown, unknown)
159  ii = nm1
118   IP1 = ii + 1
160  sum = 0
161  FOR j = IP1 TO unknown
162  sum = sum + d(ii, j) * xg(un(j))
163  NEXT j
164  xg(un(ii)) = (d(ii, np1) - sum) / d(ii, ii)
165  ii = ii - 1
166  IF ii > 0 THEN 118
167  sum = 0
168  FOR i = 1 TO unknown
169  q(i) = xg(i)
170  NEXT
171  RETURN
172  printing: PRINT "fibre lenthh="; fbrl; "cm"
173  PRINT "fibre internal diameter="; id; "cm"
174  PRINT "fibre external diameter="; od; "cm"
175  PRINT "Pot length="; lpot; "cm"
176  PRINT "Permeiblity coefficient="; k; "cm/sec"
177  PRINT "Applied head="; hap; "cm"
178  PRINT "viscosity="; m; "cm.s"
179  PRINT "Number of fibres="; nofbre
180  PRINT : PRINT
181  INPUT "Press Enter to Continue", cont
182  CLS:      pcount = 0
183  PRINT " node no.  Flowrate        TMP      %of total flow     TMP as % of total head  "
184  PRINT "
     ________________________________________________________________________________"
185  FOR i = 1 TO noseg
186  PRINT #1, i, q(i), tmp(i), q(i) / qtot * 100; tmp(i) / hap * 100
187  pcount = pcount + 1
188  PRINT USING "  ###   ##.########     ####.###      ###.###              ###.### "; i; q(i); tmp(i); q(i) /
     qtot * 100; tmp(i) / hap * 100
```

```
189  IF pcount > 40 THEN
190  INPUT "Press Enter to Continue", cont
191  pcount = 0
192  CLS
193  PRINT " node no.  Flowrate          TMP       %of total flow     TMP as % of total head  "
194  PRINT "
     _____________________________________________________________________________"
195  END IF
196  NEXT
197  PRINT : PRINT
198  PRINT "Total flowrate/fibre/sec="; qtot; "ml/sec one fibre, one end"
199  PRINT "Total flowrate/fibre/sec="; 2 * qtot; "ml/sec one fibre, two ends"
200  PRINT "Total flowrate/fibre/min="; qtot * 60; "ml/min one fibre, one end "
201  PRINT "Total flowrate/fibre/min="; qtot * 2 * 60; "ml/min one fibre, two ends "
202  PRINT "Total flowrate/min="; qtot * 60 * nofbre; "ml/min for the"; nofbre; " fibres used, one end"
203  PRINT "module Permeability="; qtot * 2 * noffipm / 1000 * 3600 / moda * (10 * 100) / (hap); "lmh/bar"
204  PRINT "module flowrate="; qtot * 2 * 20000 / 1000 * 3600 / moda; "lmh"
205  PRINT "Flowrate Ratio="; q(1) / q(noseg)
206  PRINT "Pot headloss="; hlt; "cm"
207  FOR i = 1 TO noseg
208  PRINT #2, i - 1, tmp(i) / hap * 100, q(i) / qtot * 100
209  NEXT
210  CLOSE
211  RETURN
```

E TEST MODULE PREPARATION, ASSEMBLY AND TESTING PROCEDURE

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MFI PROJECT – BENCH MEMBRANE MODULE PREPARATION

Description and photos of fabrication of end cap and modules for bench cleaning tests

N. Porcelli

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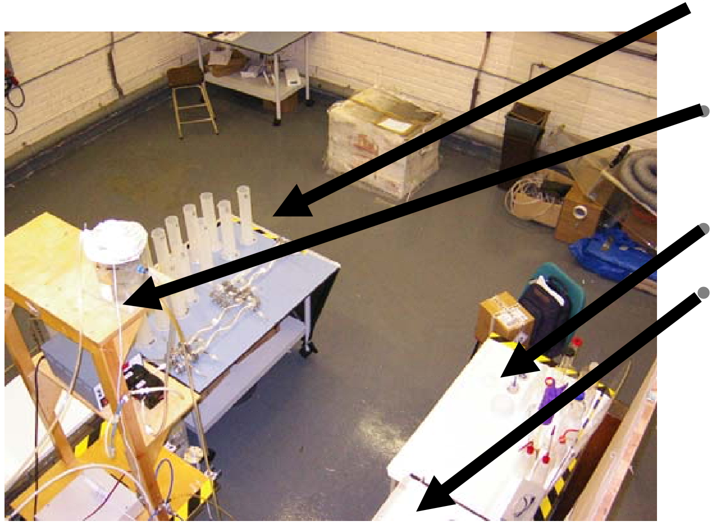
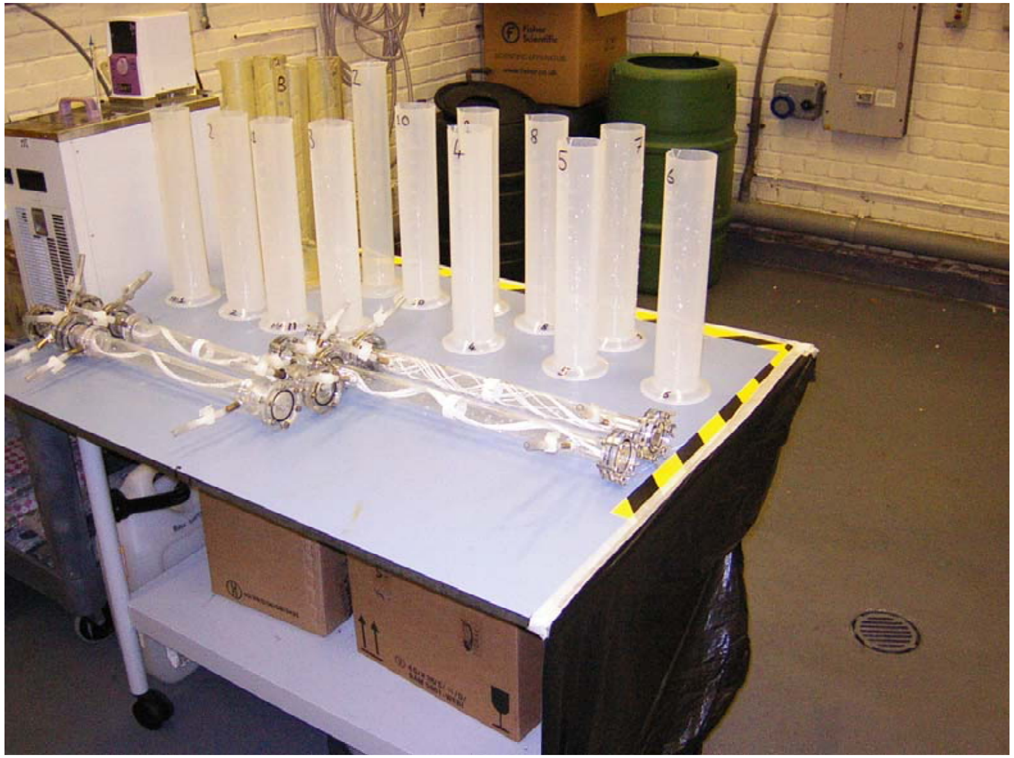
F

LAB LAYOUT



- MODULE QUEUEING BENCH
- RIG WITH DI PERMEATE HEADER
- CHEMICAL PREP BENCH
- MODULE FABRICATION AND PC BENCH

MODULE QUEUEING BENCH



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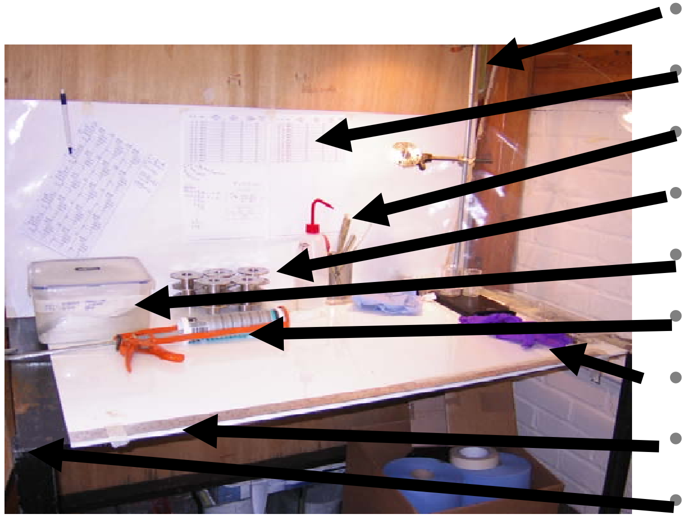
MODUE FABRICATION BENCH



- STAND
- EXPERIMENT RUN ORDER
- TOOLS
- END PIECES
- FIBRES IN TUPPERWARE
- SILICON GUN
- PPE (gloves, glasses coat)
- BENCH RULER
- END HOLDER (G clamp)

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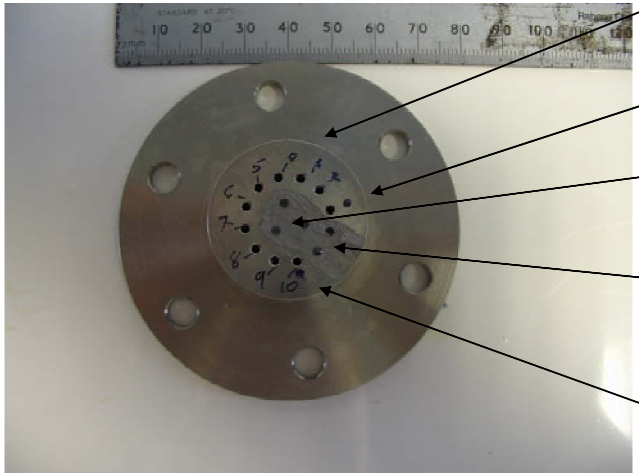
END CAP



- Fibres are inserted in the 10 holes as shown.
- The order is important – 1 through 10.
- Leave the middle holes blocked with silicon
- Leave two spare end blocked holes together for silicon gun access.
- Bottom of end piece

FIBRE POTTING

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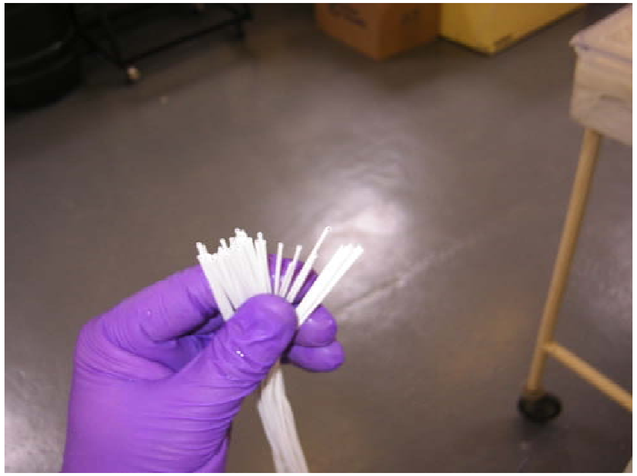

- Select a tow of about 15 wet fibres from the tupperware box.
- Should all be 600 mm long.
- Check along them to see they are not pinched. You can run your fingers along them gently for this
- Inspect visually for damage.

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THREADING FIBRES

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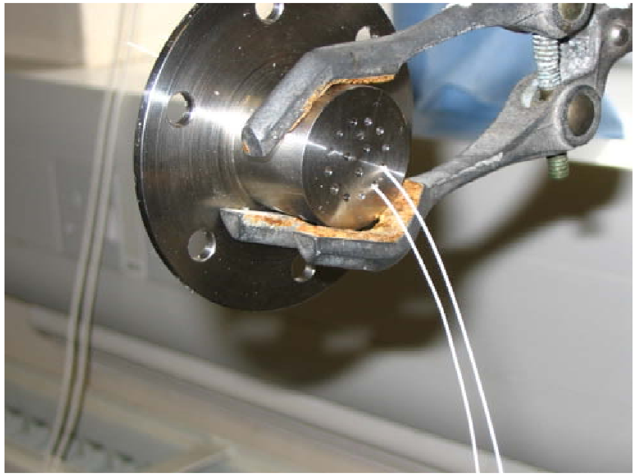

- Fibre lengths (500 -550 mm) are carefully threaded through holes, one by one.
- Care is taken not to pinch or crush the fibres.
- Keep fibres untangled.
- Pull about 75 mm through from bottom of end piece.

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FIBRE THREADING

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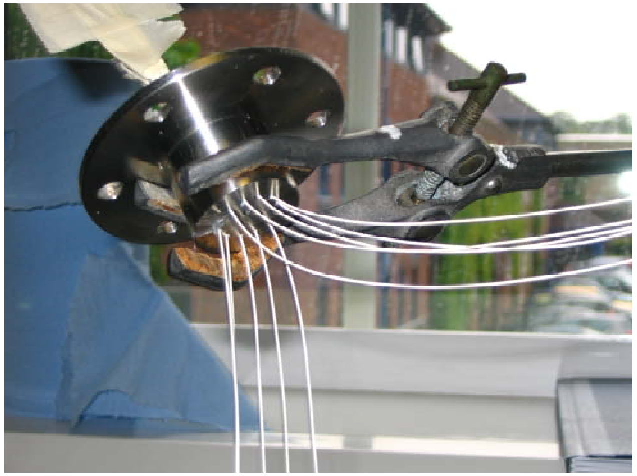

- Pull all ten fibres through.

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FIBRE THREADING

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- Leave end trailing but not tangled.

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FIBRE THREADING

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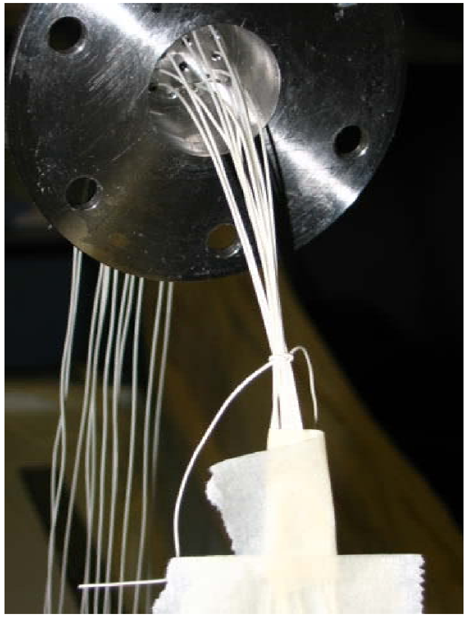

- Tie off ends using thread and tape or preferably MediClip.

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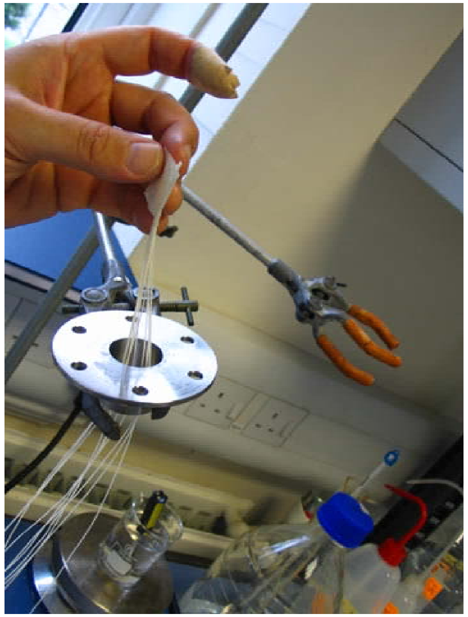
FIBRE THREADING



- End Pinching with thread and tape of MediClip.
- This helps the fibres stay in the pot when it is held upright for potting with silicon.

SILICON POTTING

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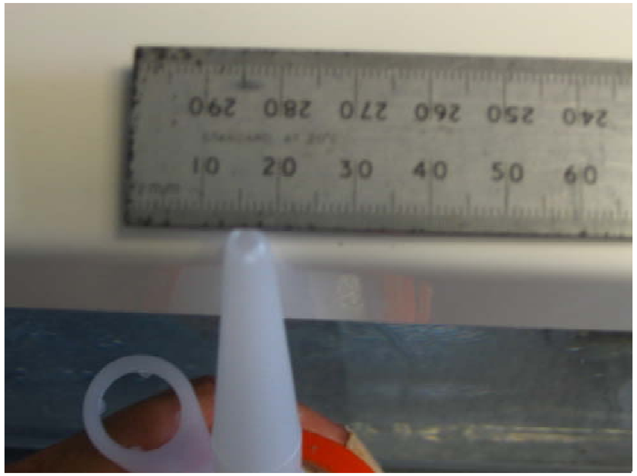

- Ensure sufficient and fresh silicon is in the gun and cut the end of the nozzle to 4-5 mm as shown.

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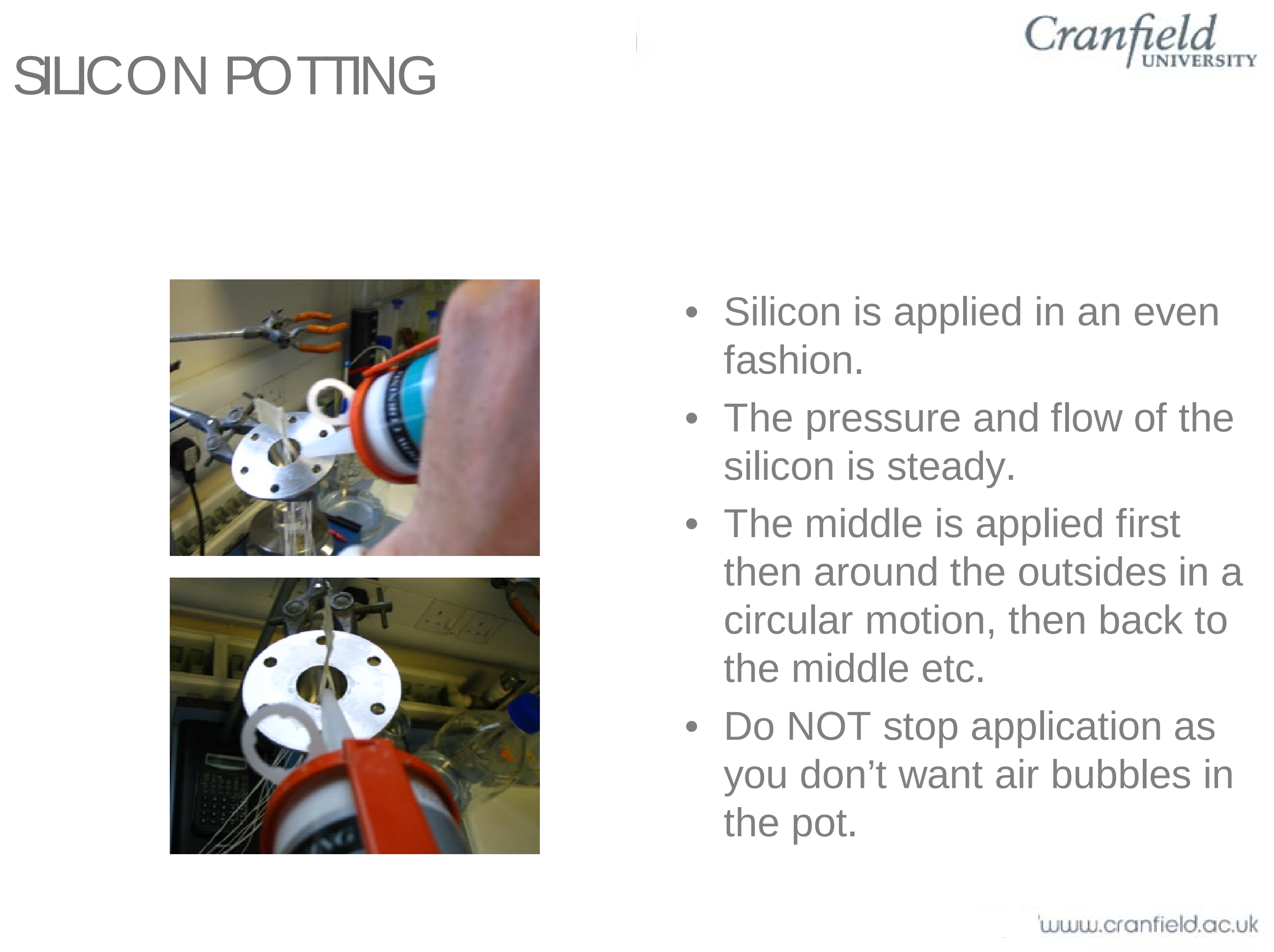

SILICON POTTING

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- Silicon is applied in an even fashion.
- The pressure and flow of the silicon is steady.
- The middle is applied first then around the outsides in a circular motion, then back to the middle etc.
- Do NOT stop application as you don't want air bubbles in the pot.

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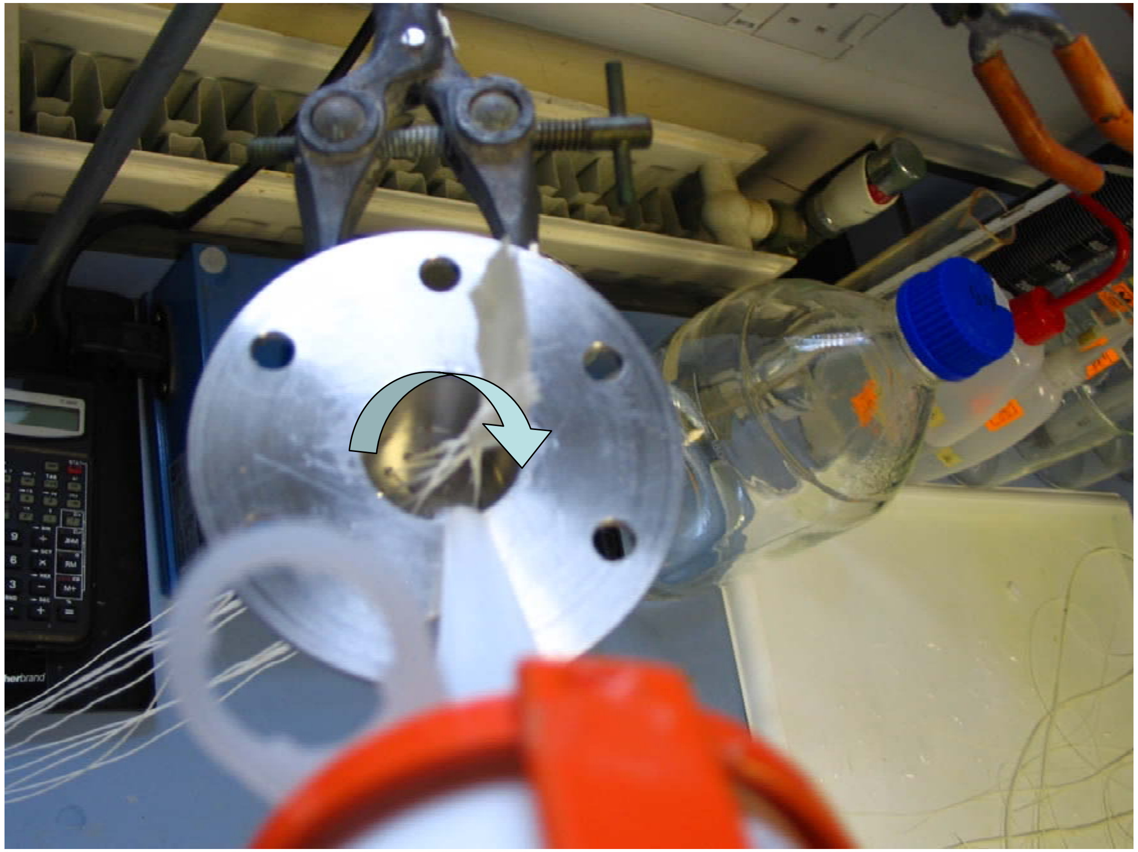
SILICON POTTING



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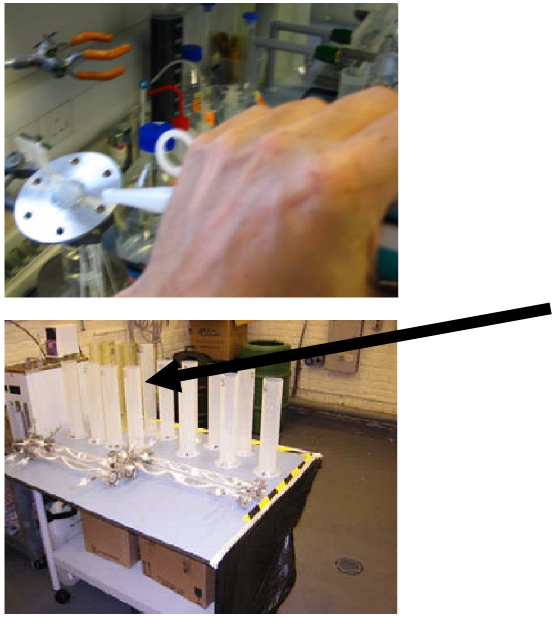
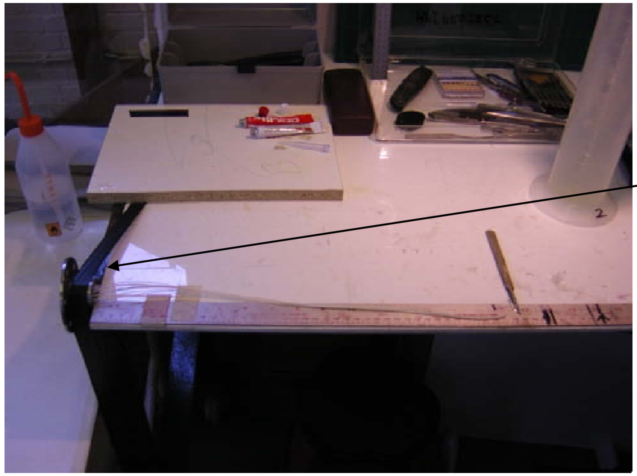
SILICON POTTING



- The pot is filled and tamped down with spatula.
- Finishing is done with finger.
- Put the fibres into their Measuring cylinder, which are filled with sample water or tap water – fibres must be kept wet.
- Potted fibres are left to cure for 24-36 hours.

FIBRE LENGTH SETTING



- After curing the fibres are cut to length. Each threaded end piece is removed from the measuring cylinder and held on the bench as shown.
- The bottom end is held against a ruler and a “G” clamp and PVC tape ensures it stays fixed. This is important as you need both hands for cutting and the fibres can be damaged easily if it falls/moves.

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FIBRE LENGTH SETTING

- A sharp scalpel with new blade is used to cut the fibres to length at the bottom end.
- Cut the fibres to ~410 mm for a 400 mm required length and 510 mm if requiring 500 mm length.

BOTTOM END SEALING

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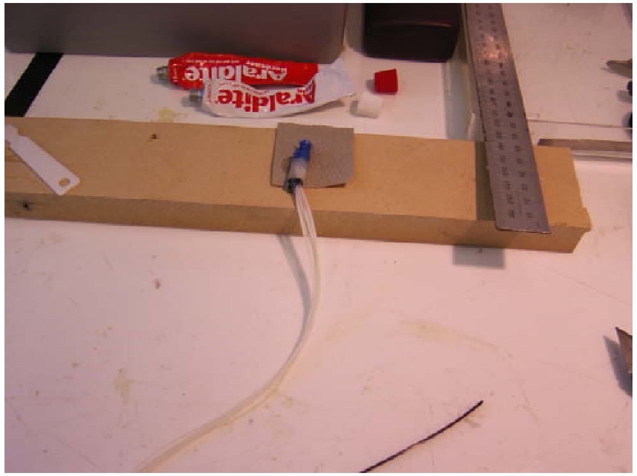

- The end is sealed using an improvised Araldite™ plug.

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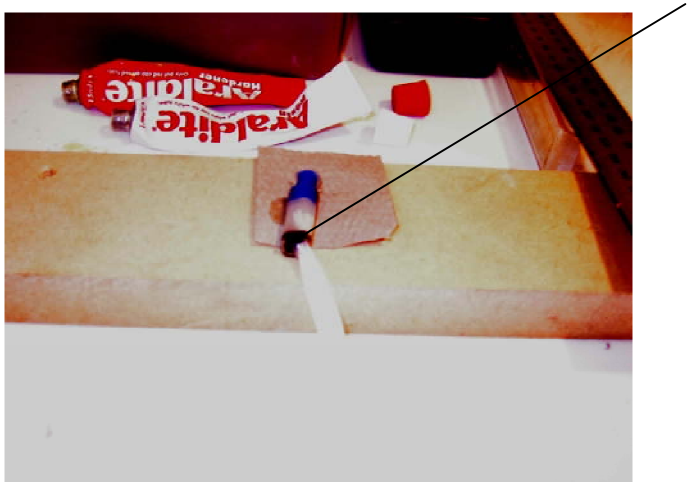
BOTTOM END SEALING



- A cable tie is used to pinch/crush seal the fibres at the required length.

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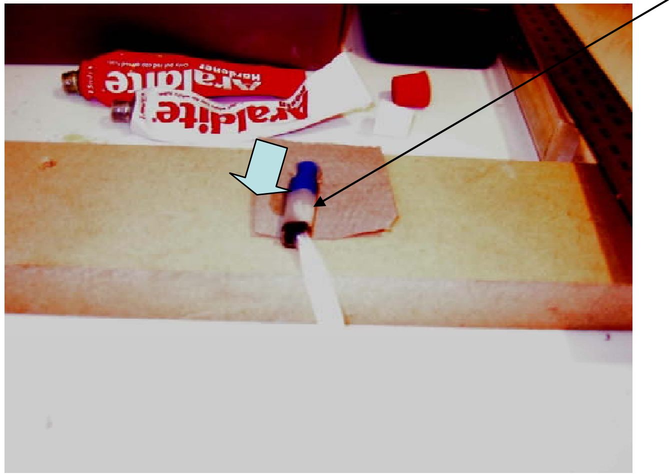
BOTTOM END SEALING



- Then a piece of hose is applied over the end and 50/50 mixed araldite inserted.

BOTTOM END SEALING

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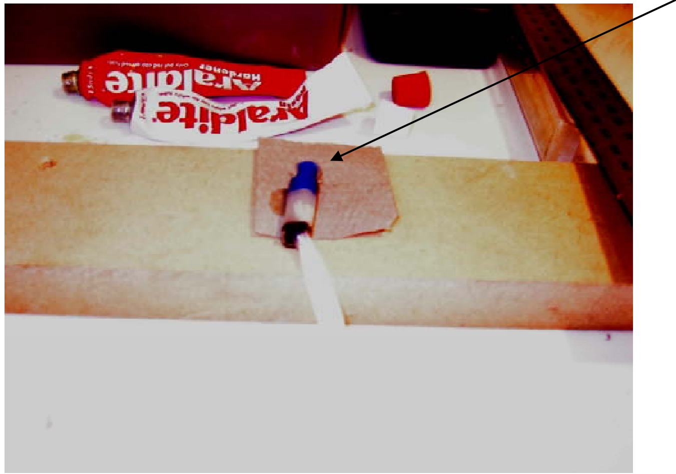

- A plug is inserted to push the araldite up.

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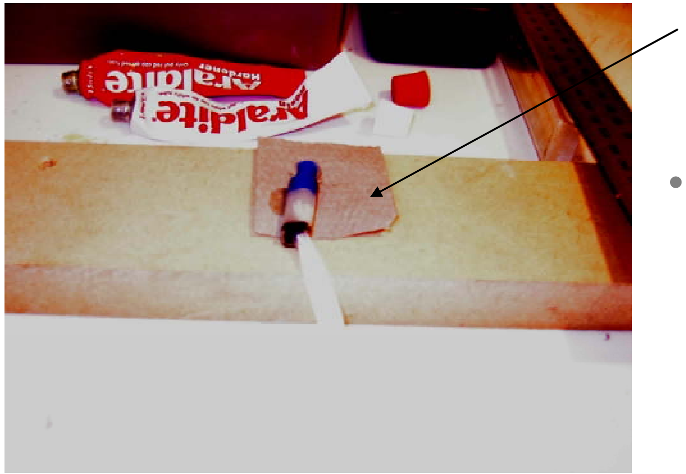
BOTTOM END SEALING



- Paper is used to stop the araldite sticking to the bottom of the volumetric flasks
- The potted fibre is then transferred back to the flask and allowed to cure for 12 hours.

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TOP END FIBRE CUTTING

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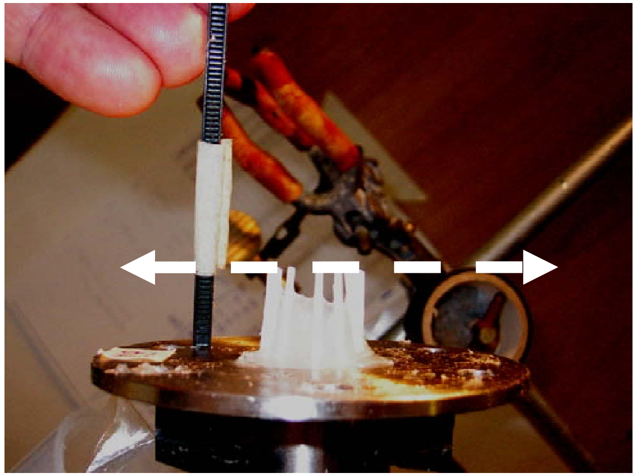

- The top of each fibre assembly is then cut.
- Use the marked stick to set the length of 15 mm.
- Pot length is important to the hydraulic profile of the module so they must be the same length.
- Ensure scalpel blade clean and sharp before cutting.

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TOP END FIBRE CUTTING

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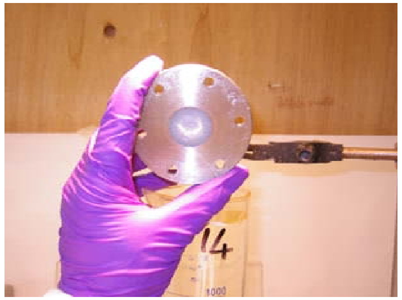

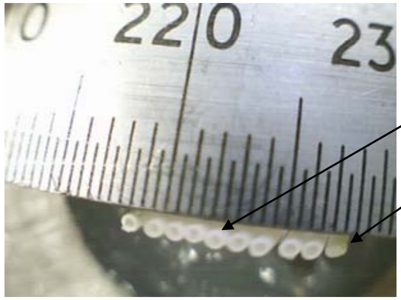

- The cut should be clean and even. Use deliberate stroke with sharp scalpel to avoid pinching the fibres or deforming the ends from circular form.
- OK
- Deformed

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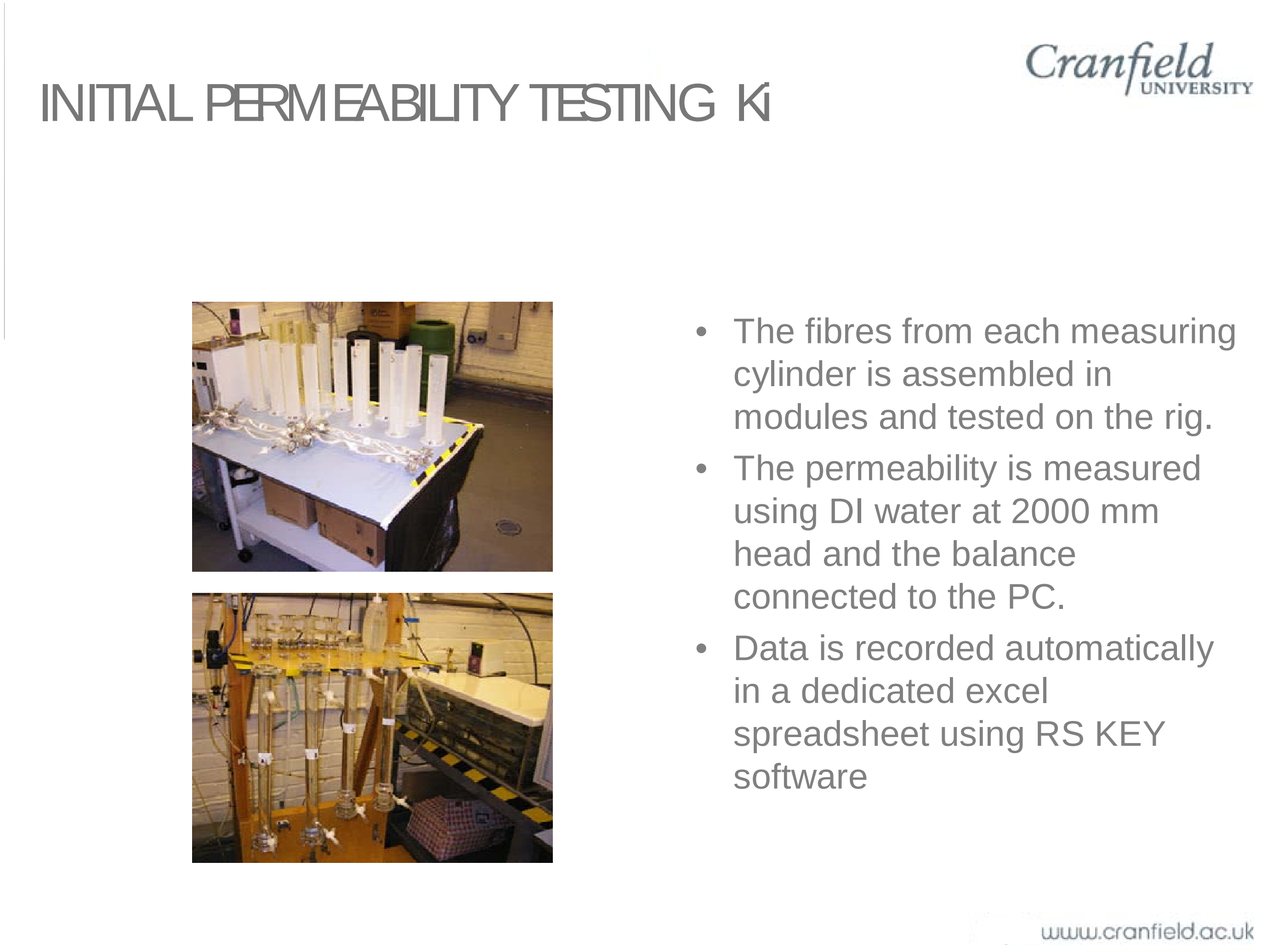

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INITIAL PERMEABILITY TESTING Ki

- The fibres from each measuring cylinder is assembled in modules and tested on the rig.
- The permeability is measured using DI water at 2000 mm head and the balance connected to the PC.
- Data is recorded automatically in a dedicated excel spreadsheet using RS KEY software

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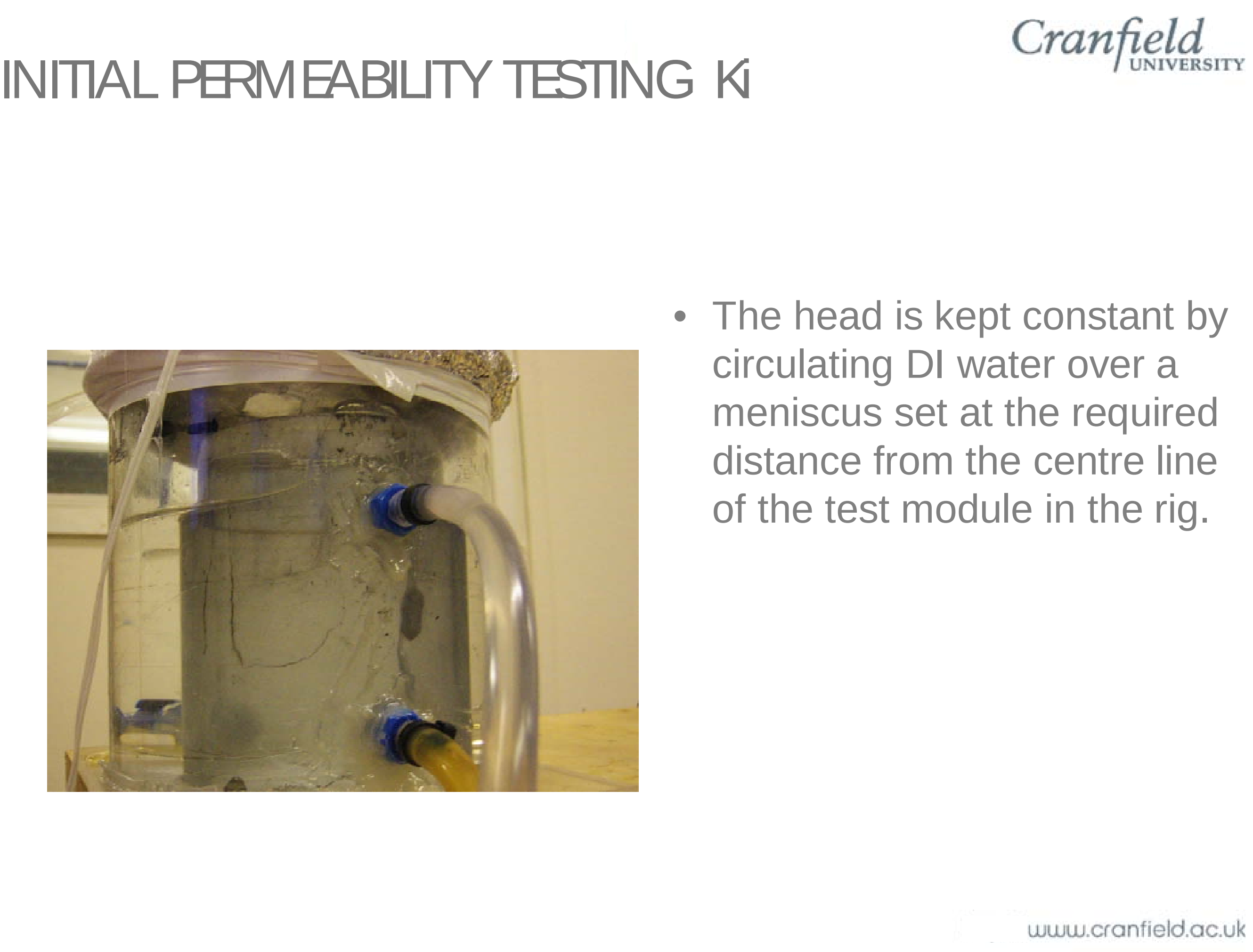

INITIAL PERMEABILITY TESTING Ki

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- The head is kept constant by circulating DI water over a meniscus set at the required distance from the centre line of the test module in the rig.

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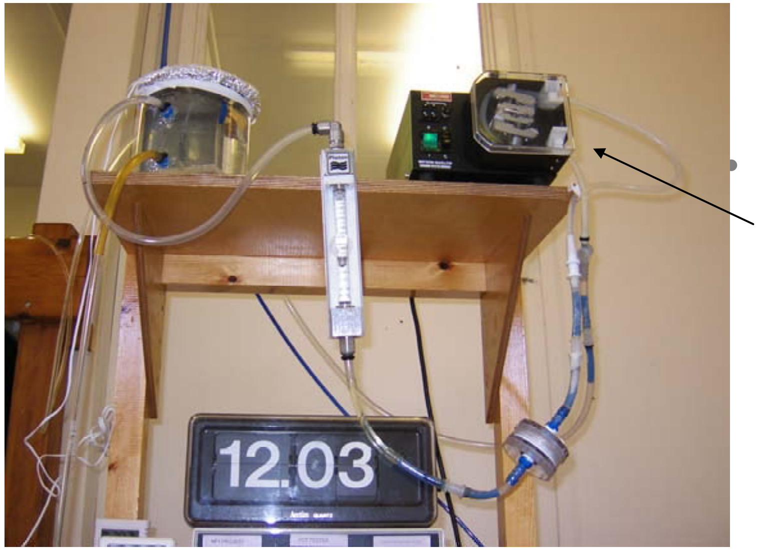
INITIAL PERMEABILITY TESTING Ki



- During the test period a peristaltic pump keeps the meniscus level.

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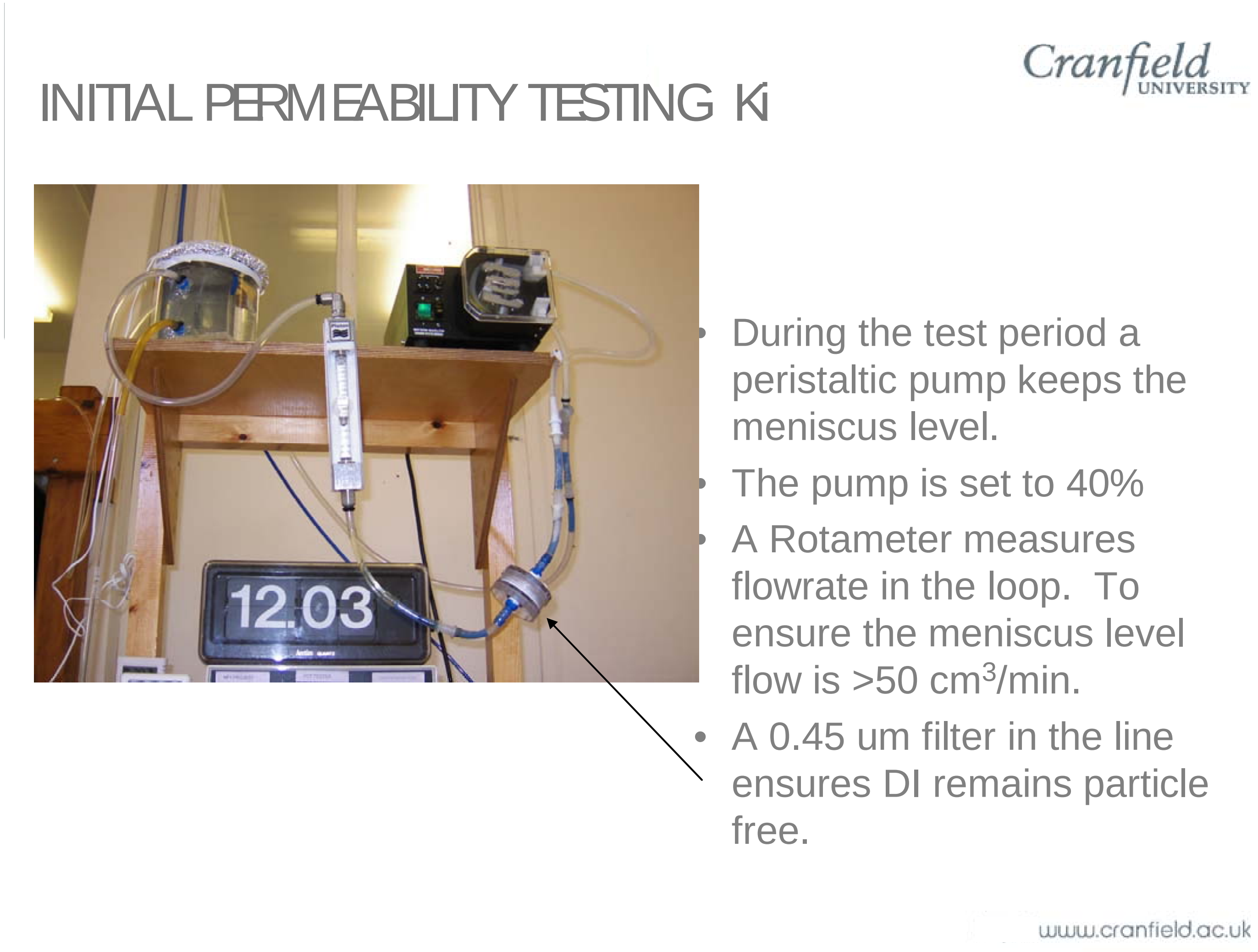

INITIAL PERMEABILITY TESTING Ki

- During the test period a peristaltic pump keeps the meniscus level.
- The pump is set to 40%
- A Rotameter measures flowrate in the loop. To ensure the meniscus level flow is >50 cm^3/min.
- A 0.45 um filter in the line ensures DI remains particle free.

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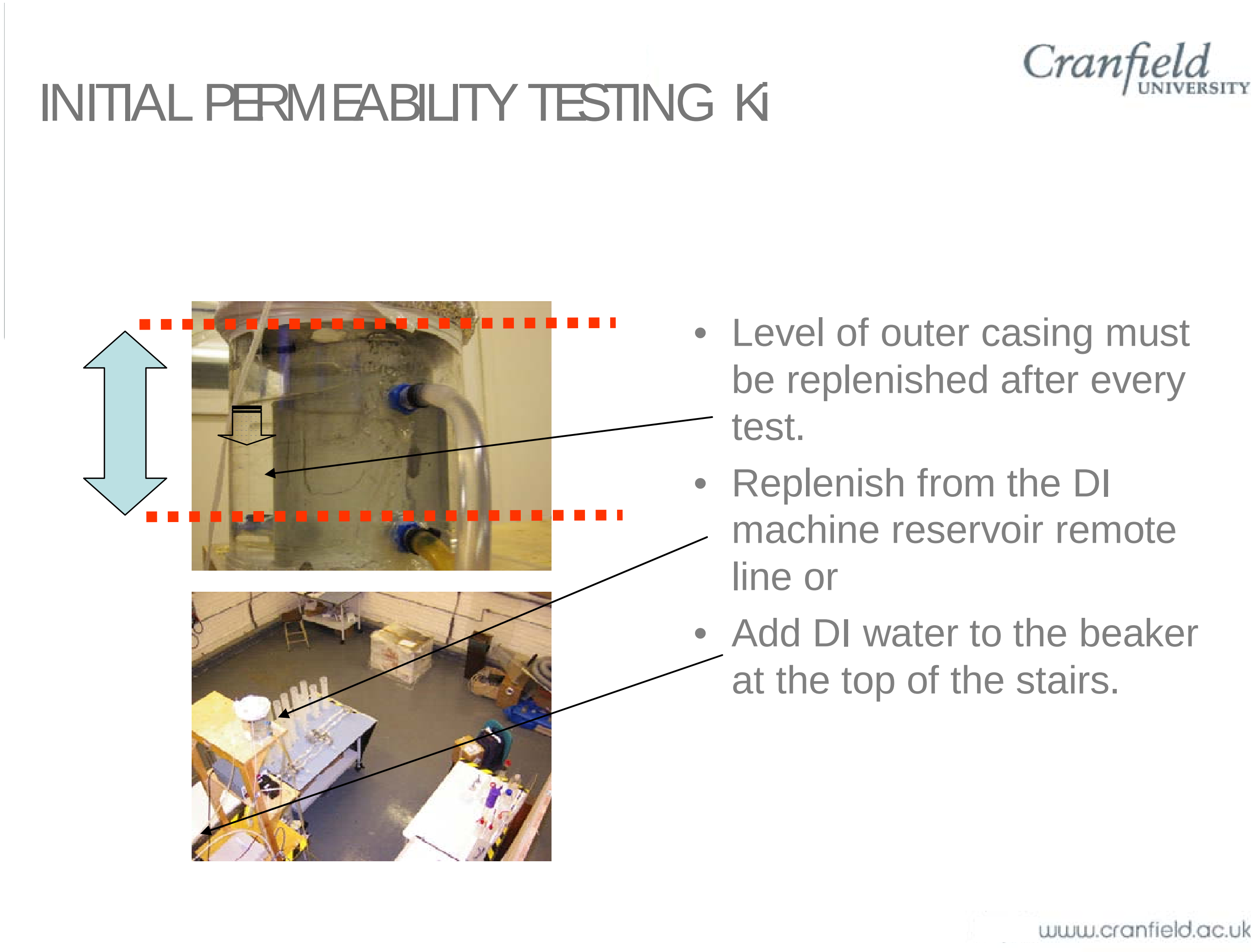

INITIAL PERMEABILITY TESTING Ki

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- Level of outer casing must be replenished after every test.
- Replenish from the DI machine reservoir remote line or
- Add DI water to the beaker at the top of the stairs.

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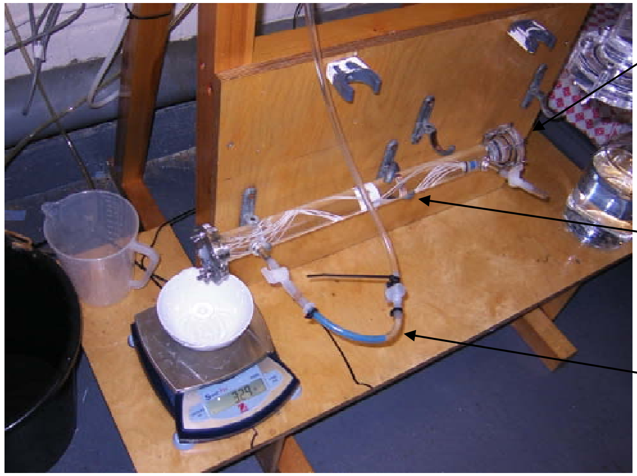
INITIAL PERMEABILITY TESTING Ki



- Each end cap is assembled in one of the four modules and placed in the rig brackets.
- A bracket clamp is put over the barrel of the module to hold it in place.
- DI water lines are connected.

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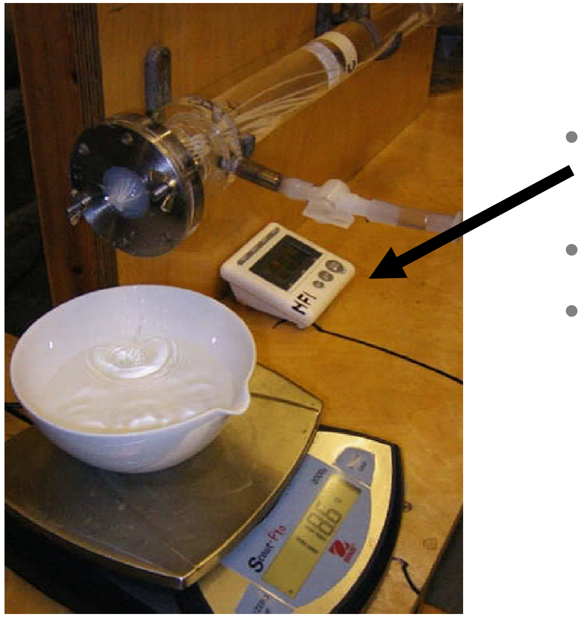
INITIAL PERMEABILITY TESTING K_i

- Balance and crucible beaker are aligned under the open fibre ends.
- The DI water taps are turned on and the module purged of air.
- Allow the flow to establish for 2-3 minutes.

INITIAL PERMEABILITY TESTING Ki

Cranfield UNIVERSITY



- A timer is also used to time the experiments.
- Ten minute runs are used.
- The PC logs the data automatically

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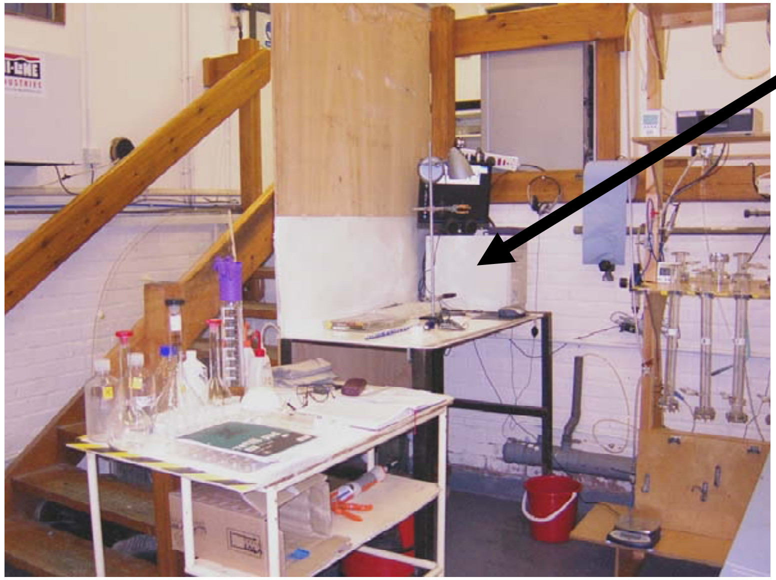
INITIAL PERMEABILITY TESTING Ki



- On PC a new excel spreadsheet from template is opened.

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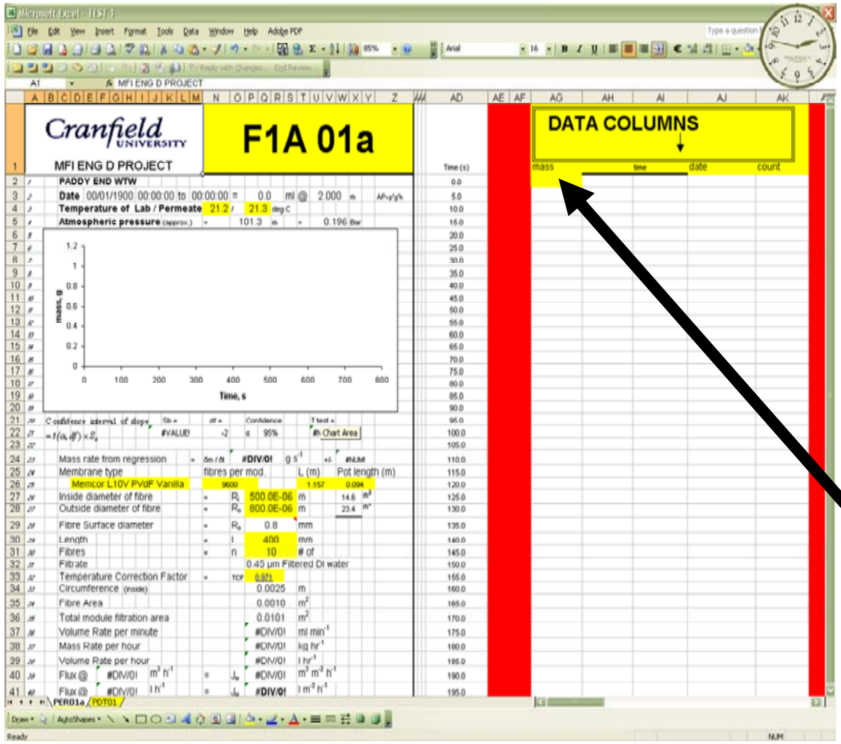
INITIAL PERMEABILITY TESTING Ki



MFI ENG D PROJECT

F1A 01a

PADDY END WTW

Date 00/01/1900 00:00:00 to 00:00:00 = 0.0 ml @ 2.000 m

Temperature of Lab / Permeate 21.2 / 21.3 deg C

Atmospheric pressure (approx.) = 101.3 m = 0.196 Bar

Confidence interval of slope Sb = #VALUE! df = -2 Confidence α 95% T test =

Mass rate from regression = δm / δt #DIV/0! g s⁻¹ +/- #NUM

Membrane type: Memcor L10V PVdF Vanilla; fibres per mod. 9600; L (m) 1.157; Pot length (m) 0.094

Inside diameter of fibre = R_i 500.0E-06 m 14.6 m^2

Outside diameter of fibre = R_o 800.0E-06 m 23.4 m^2

Fibre Surface diameter = R_s 0.8 mm

Length = l 400 mm

Fibres = n 10 # of

Filtrate 0.45 µm Filtered DI water

Temperature Correction Factor = TCF 0.971

Circumference (inside) 0.0025 m

Fibre Area 0.0010 m^2

Total module filtration area 0.0101 m^2

Volume Rate per minute #DIV/0! ml min⁻¹

Mass Rate per hour #DIV/0! kg hr⁻¹

Volume Rate per hour #DIV/0! l hr⁻¹

Flux @ #DIV/0! $m^3 h^{-1}$ = J_w #DIV/0! $m^3 m^{-2} h^{-1}$

Flux @ #DIV/0! $l h^{-1}$ = J_w #DIV/0! $l m^{-2} h^{-1}$

DATA COLUMNS

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- Membrane and experiment data entered
- Temperature of lab and permeate entered in yellow boxes.
- Membrane type and module configuration entered.
- Save with run name – eg F1A-per01a through F1A-per16a
- Select data entry box- “AG2”
- Start or switch to the “**RS KEY**” software.

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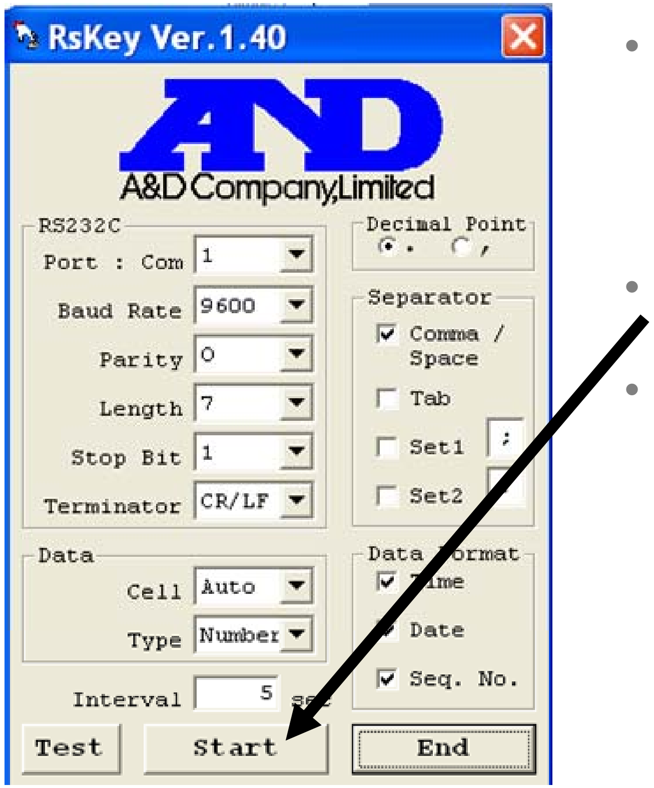
INITIAL PERMEABILITY TESTING Ki



RS232C
Port : Com 1
Baud Rate 9600
Parity O
Length 7
Stop Bit 1
Terminator CR/LF

Data
Cell Auto
Type Number

Interval 5 sec

Decimal Point
● . ○ ,

Separator
☑ Comma / Space
☐ Tab
☐ Set1 ;
☐ Set2

Data Format
☑ Time
☑ Date
☑ Seq. No.

- Check RS KEY software communications settings before starting – AS SHOWN
- If satisfied push the start button.
- You should be returned to the Excel spreadsheet at the date point.

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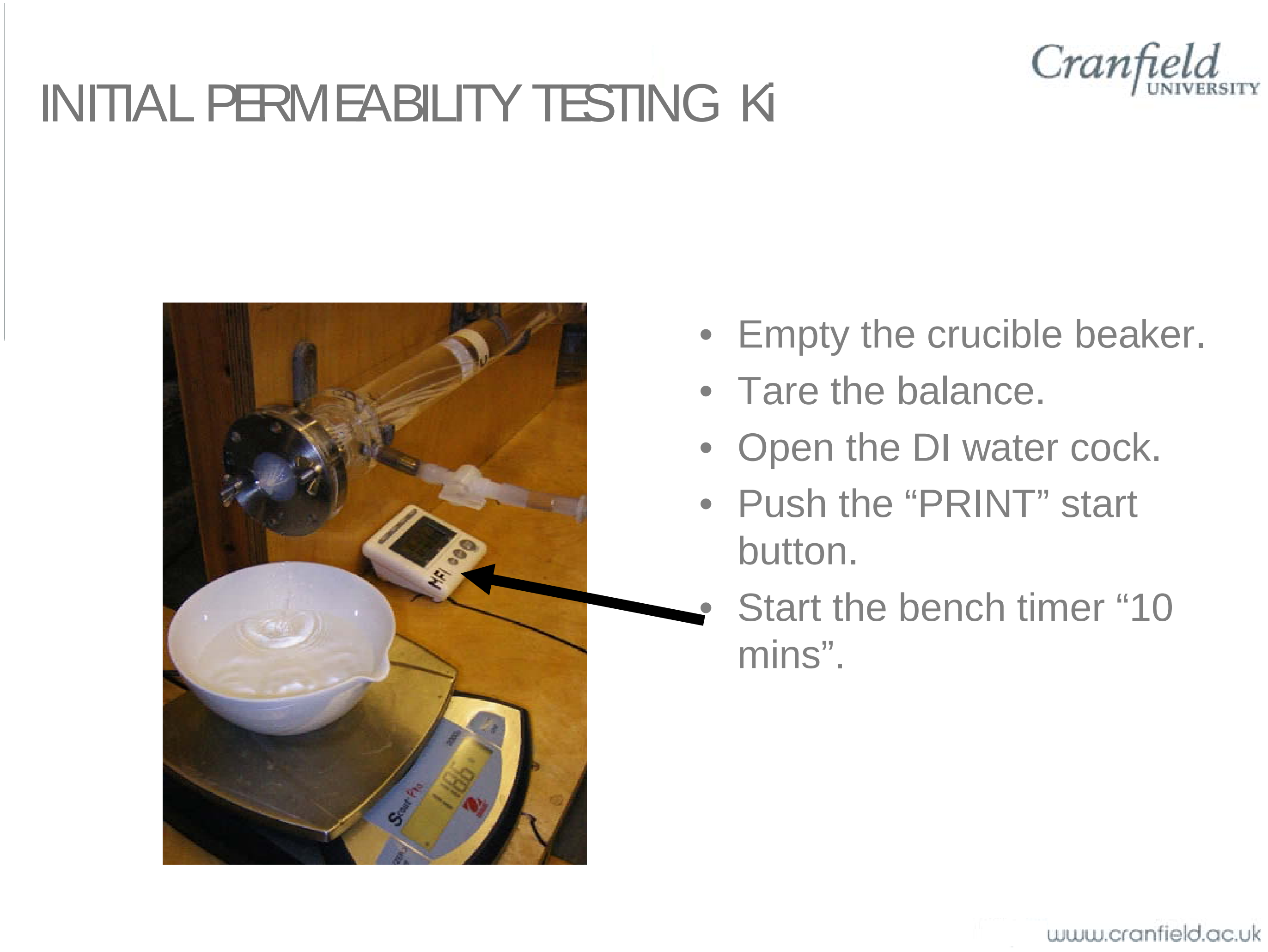

INITIAL PERMEABILITY TESTING K_i

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- Empty the crucible beaker.
- Tare the balance.
- Open the DI water cock.
- Push the “PRINT” start button.
- Start the bench timer “10 mins”.

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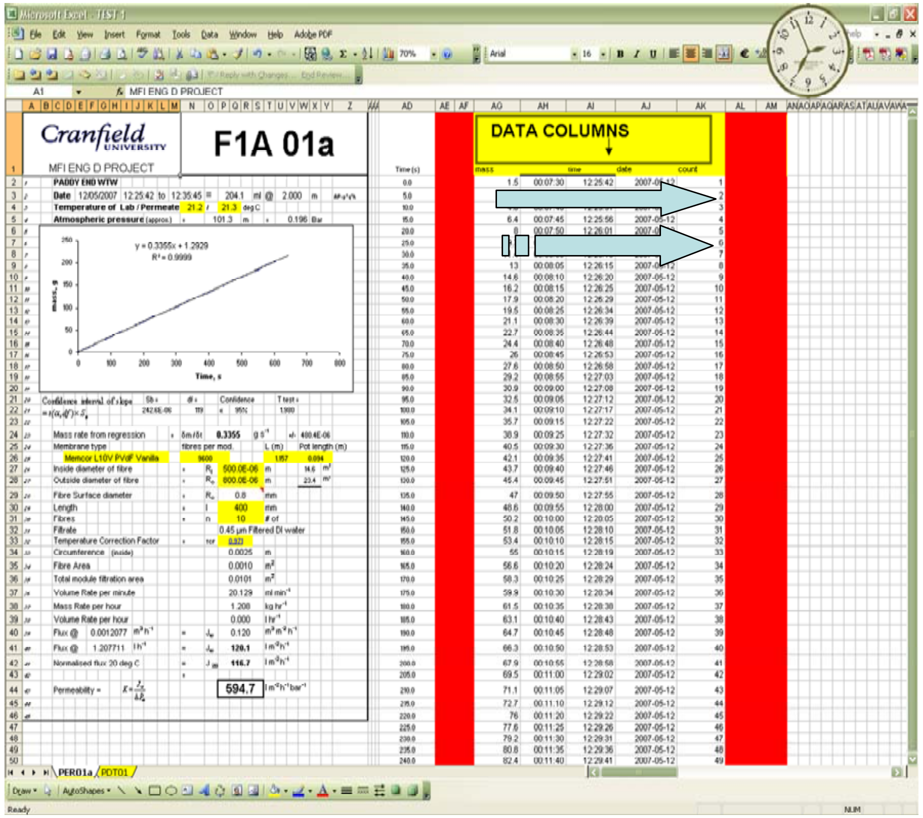
INITIAL PERMEABILITY TESTING Ki



MFI ENG D PROJECT

F1A 01a

PADDY END WTW

Date 12/05/2007 12:25:42 to 12:35:45 = 204.1 ml @ 2.000 m

Temperature of Lab / Permeate 21.2 / 21.3 deg C

Atmospheric pressure (approx.) = 101.3 m = 0.196 Bar

Confidence interval of slope $= t(\alpha, df) \times S_b$ Sb = 242.6E-06 df = 119 Confidence = 95% T test = 1.980

Mass rate from regression = δm/δt 0.3355 g s⁻¹ +/- 400.4E-06

Membrane type	fibres per mod.	L (m)	Pot length (m)
Memcor L10V PVdF Vanilla	9600	1.157	0.094

Inside diameter of fibre = R_i 500.0E-06 m 14.6 m²

Outside diameter of fibre = R_o 800.0E-06 m 23.4 m²

Fibre Surface diameter = R_o 0.8 mm

Length = l 400 mm

Fibres = n 10 # of

Filtrate 0.45 µm Filtered DI water

Temperature Correction Factor = TCF 0.921

Circumference (inside) 0.0025 m

Fibre Area 0.0010 m²

Total module filtration area 0.0101 m²

Volume Rate per minute 20.129 ml min⁻¹

Mass Rate per hour 1.208 kg hr⁻¹

Volume Rate per hour 0.000 l hr⁻¹

Flux @ 0.0012077 m³h⁻¹ = J_w 0.120 m³m⁻²h⁻¹

Flux @ 1.207711 l h⁻¹ = J_w **120.1** l m⁻²h⁻¹

Normalised flux 20 deg C = J_{20} **116.7** l m⁻²h⁻¹

Permeability = $K = \frac{J_w}{\Delta P_a}$ **594.7** l m⁻²h⁻¹bar⁻¹

DATA COLUMNS

Time (s)	mass		time	date	count
0.0	1.5	00:07:30	12:25:42	2007-0	1
5.0					2
10.0					3
15.0	6.4	00:07:45	12:25:56	2007-05-12	4
20.0	8	00:07:50	12:26:01	2007-0	5
25.0					6
30.0					7
35.0	13	00:08:05	12:26:15	2007-0	8
40.0	14.6	00:08:10	12:26:20	2007-05-12	9
45.0	16.2	00:08:15	12:26:25	2007-05-12	10
50.0	17.9	00:08:20	12:26:29	2007-05-12	11
55.0	19.5	00:08:25	12:26:34	2007-05-12	12
60.0	21.1	00:08:30	12:26:39	2007-05-12	13
65.0	22.7	00:08:35	12:26:44	2007-05-12	14
70.0	24.4	00:08:40	12:26:48	2007-05-12	15
75.0	26	00:08:45	12:26:53	2007-05-12	16
80.0	27.6	00:08:50	12:26:58	2007-05-12	17
85.0	29.2	00:08:55	12:27:03	2007-05-12	18
90.0	30.9	00:09:00	12:27:08	2007-05-12	19
95.0	32.5	00:09:05	12:27:12	2007-05-12	20
100.0	34.1	00:09:10	12:27:17	2007-05-12	21
105.0	35.7	00:09:15	12:27:22	2007-05-12	22
110.0	38.9	00:09:25	12:27:32	2007-05-12	23
115.0	40.5	00:09:30	12:27:36	2007-05-12	24
120.0	42.1	00:09:35	12:27:41	2007-05-12	25
125.0	43.7	00:09:40	12:27:46	2007-05-12	26
130.0	45.4	00:09:45	12:27:51	2007-05-12	27
135.0	47	00:09:50	12:27:55	2007-05-12	28
140.0	48.6	00:09:55	12:28:00	2007-05-12	29
145.0	50.2	00:10:00	12:28:05	2007-05-12	30
150.0	51.8	00:10:05	12:28:10	2007-05-12	31
155.0	53.4	00:10:10	12:28:15	2007-05-12	32
160.0	55	00:10:15	12:28:19	2007-05-12	33
165.0	56.6	00:10:20	12:28:24	2007-05-12	34
170.0	58.3	00:10:25	12:28:29	2007-05-12	35
175.0	59.9	00:10:30	12:28:34	2007-05-12	36
180.0	61.5	00:10:35	12:28:38	2007-05-12	37
185.0	63.1	00:10:40	12:28:43	2007-05-12	38
190.0	64.7	00:10:45	12:28:48	2007-05-12	39
195.0	66.3	00:10:50	12:28:53	2007-05-12	40
200.0	67.9	00:10:55	12:28:58	2007-05-12	41
205.0	69.5	00:11:00	12:29:02	2007-05-12	42
210.0	71.1	00:11:05	12:29:07	2007-05-12	43
215.0	72.7	00:11:10	12:29:12	2007-05-12	44
220.0	76	00:11:20	12:29:22	2007-05-12	45
225.0	77.6	00:11:25	12:29:26	2007-05-12	46
230.0	79.2	00:11:30	12:29:31	2007-05-12	47
235.0	80.8	00:11:35	12:29:36	2007-05-12	48
240.0	82.4	00:11:40	12:29:41	2007-05-12	49

- Run data will automatically fill in the DATA COLUMNS row by row.
- After 10 minutes 600 readings the timer should sound and the run is completed.

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CLEANING OF FIBRES

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- Put stands in bath
- Fill bath
- Start cooling and heating controllers and set to desired temperature
- Put test tubes in stands when required. Pull chemicals through membranes from tube using suction.
- Time cleaning run using timer

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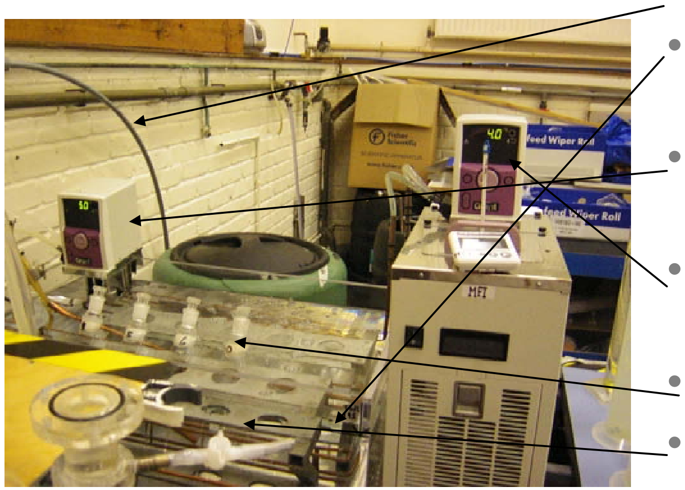
CLEANING OF FIBRES



- Potable supply line
- Water Bath – tapwater filled to line
- Heater circuit Temp. controller
- Cooler circuit Temp. controller
- Cleaning Tubes
- Stands

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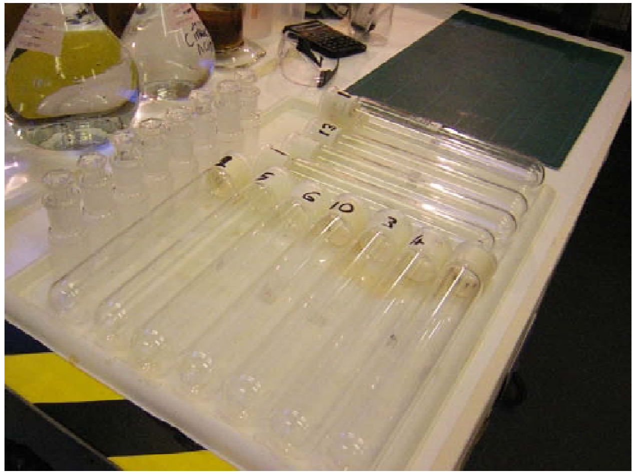
CLEANING OF FIBRES



- Prepare the numbered cleaning tubes.
    - Acid clean and rinse.
    - Numbers correspond to run order.

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CLEANING OF FIBRES

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- Prepare and label three concentrations of the chemical cleanant to be tested.
    - C1 1000 ml
    - C2 2000 ml
    - C3 1000 ml

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CLEANING OF FIBRES

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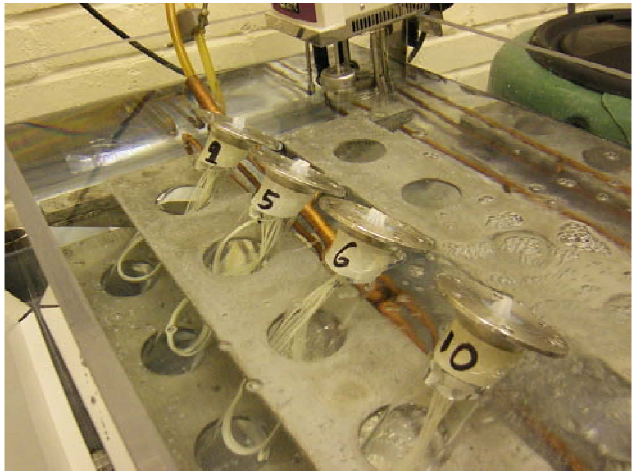

- The fibres are removed from their measuring cylinders and placed in cleaning tubes.
- Fibres are carefully folded to go into to flasks.
- Tubes are then placed in the preheated bath for the required time as per experiment run sheet.

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CLEANING OF FIBRES

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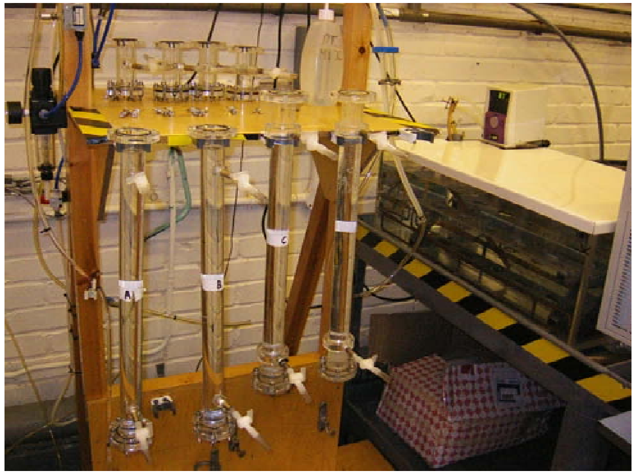

- Remove cleaned fibres from the cleaning tube.
- Rinse the chemicals off the outside of the fibres.
- Transfer the fibres to the test module.
- Connect and run the DI water through the fibres.
- Re-Run permeability test ensuring correct ID used

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CLEANED PERMEABILITY TESTING Kf

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- Run permeability tests on each cleaned membrane set and calculate permeability recovery from the initial fouled state relative to the virgin permeability.

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