Evaluation of granular media to wastewater treatment; advanced primary treatment utilising modified pumice, titanomagnetite and ceramic nozzles.

A Thesis

Submitted in partial fulfilment of the requirements

for the Degree of

Doctor of Philosophy

in Materials and Processing Engineering

at the University of Waikato

by

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KO TE TANGATA

The University of Waikato 2002

Abstract

The treatment of wastewater to higher standards is becoming increasingly important as legislation and public perception demand reduced impact of human activity on the environment. Filtration of primary treated effluent (PTE) and secondary treated effluent (STE) offers a possible sustainable technology for meeting or partially meeting these demands.

Granular media filtration, if used at all, is typically incorporated in wastewater treatment as a polishing step following secondary biological treatment and flocculation. The use of filtration following primary sedimentation has not been practised widely. The objective of this research was to investigate the feasibility of filtering PTE using New Zealand's indigenous volcanic deposits as filter media in a recently developed dual media filtration system incorporating porous ceramic nozzles. The two media types studied included titanomagnetite (TM) and Silicon Sponge (SS). Titanomagnetite is an ironsand concentrate with a very fine grain size (0.07 to 0.18 mm). Silicon Sponge is abraded river pumice and was available in two sizes: SS 14/24 (0.6 to 1 mm) and SS 7/14 (0.4 to 2 mm). The surface of TM grains is smooth and that of SS is highly porous due to the vesiculated structure of pumice, while maintaining a low abrasive factor.

In order to test filtration behaviour of TM and SS media, bench-scale experiments, using glass columns were performed. A model effluent was prepared from reconstituted dairy farm manure. Suspended particle sizes in farm dairy effluent (FDE) were similar to those measured for PTE. Most of the bench trials used FDE. Filtration trends were easy monitored by turbidity reduction. About 50% of FDE turbidity was removed by 100 mm TM columns. TM performance was improved by the addition of TM fines but flows were reduced. Unmodified TM retained most particles larger than 5 μm. This would likely to remove protozoa such as giardia and cryptosporidium. The effect of pH on turbidity removal by TM beds was significant. Turbidity removal increased from 40 to 80% upon the lowering of the effluent pH from 6.5 to 3.5. Raising the pH above 6.5 had no effect. No pH-induced change in particle size distribution of the effluent was observed. A pH of 3.5 is below the isoelectric point of TM and it is likely that the enhanced turbidity removal was due to electrostatic effects.

In bench–scale filtration of PTE at pH 6.5, a 100 mm depth of TM removed 30% of the turbidity. Increasing the bed depth for TM did not lead to significant further improvement in filter performance. In the case of filtration by a fine SS (14/24), 25% turbidity removal was achieved using a 500 mm bed. Increasing the bed depth to 1800 mm increased turbidity removal to 50%. In glass column experiments, a 1000 mm deep bed of coarse SS (7/14) removed 30% of the turbidity. An increase to 35% occurred when the bed depth, using this medium was increased to 1600 mm. The same media when used in larger pilot plant filters removed 50 % of the turbidity. A particular feature of the TM medium was its ease of fluidisation at flowrates < 10 m/h. This characteristics make it a suitable medium for pulsed bed operation. A feature of the SS medium was its large hydraulic conductivity and high solids holding capacity.

On the basis of the results obtained from the bench trials, evaluation filter units incorporating the essential features of the porous ceramic dual media (PCDM) system were constructed. The evaluation systems were equipped with under drains that incorporated a variety of purpose designed titanomagnetite ceramic (TMC) nozzles held in place by a tube bolt that also delivered air and water during filtration and backwashing. Investigation of the flow characteristics of the TMC nozzles showed that the major flow-regulating factor in the nozzle system was the water orifice in the bolt stem. Resistance in up-flow was consistently larger than resistance in down-flow. This was attributed to differences in the hydrodynamics of flow in each direction. Re-designing the bolt could lead to a 30% reduced flow-restriction during backwashing.

Biofilm growth on the TM and SS filtration media and the TMC used to manufacture the ceramic nozzles was studied by prolonged immersion in partially digested effluent. SS media showed a high degree of vesicular space and under aerobic conditions, affinity for loose attachment of organic material. TM showed most anaerobic film growth, accompanied by formation of a red iron oxide deposit. Significantly little biofilm growth on the ceramic material was observed. Such growth, if it occurred, could block the nozzles.

Pilot plant evaluation trials using PTE showed that for SS, filtration efficiency increased steeply with bed depth up to a bed depth of 500 mm, irrespective of the grade of SS used. At depths greater than 500 mm, filtration efficiency increased more slowly with depth. Run times were typically of 50 h duration by which time flow rates under a constant head of 2700 mm had decreased from 12 m/h to 1 m/h. Bed loadings were as high as 100 kg/m^3 at the top of the bed with a mean of 22 kg/m^3. In the case of TM, a minimum bed depth of 100 mm was required in order to accommodate the TMC tube nozzles. Deeper beds blocked after very short run times. Flowrates of 1 m/h or less and run times of 1 h or less were observed.

For both types of media, solids were easily dislodged during backwashing.

Pulsed backwashing greatly reduced backwash water requirements and in the case of SS beds yielded sludge with solids content suitable for anaerobic digestion.

Flow-dynamics through PCDM filter nozzles showed increased resistance in upflow being caused by designs in the nozzle-bolt. Vena contracta effects provide an explanation.

PCDM filtration using SS as the major medium offers a viable technology for the continuous removal of much of the TSS from PTE. A potential energy saving can be envisaged from the reduced aeration required for reduced BOD loadings going to secondary biological treatment. In addition, the concentrated backwash water from bed backwashing, can be transferred directly to anaerobic digesters. Reduced CO_2 generation and an increase in methane production would off-set installation and running costs of the filter in about 15 years.

Acknowledgements

I wish to express my greatest appreciation to Prof. A. Langdon for his professional support throughout the extended period of my study.

The help and support of Mr. P. Smyth, Mr. T. Hill and Mr. G. Jamieson from Works Filter Systems Ltd of Hamilton, New Zealand who provided filtration equipment for the pilot plant, the Foundation for Research, Science and Technology and the Department of Materials and Process Engineering at Waikato University under Dr. J. Swan, both for their financial support is acknowledged gratefully.

There are many individuals who contributed help during the experimental work: Dr. C. Fee, and Dr. M. Walmsley for suggestions and help in hydrodynamics, Dr. C. Haarfoot for explanation with regards to biological assays, Dr. N. Kim for help with analysis of Arsenic, Dr.R. Briggs, Prof. M. Hochstein and Ms P. Cooke for explanations in mineralogy, Mr. A. Harris for assistance with scanning electron microscopic photography, Mr. Y. Choo and Mr. A. White for their active involvement with pilot plant studies. Also Mr. W. Potts from the Hamilton City Council and Mr. G. Mc Graw from the Matamata Piako District Council for accommodating me and my colleges at their wastewater treatment plants. I also enjoyed the generous help throughout this study from my friend and college, Mrs. Z. Yang. Further help during the writing process was provided by Ms. R. Nilawongse. Ms. K. Wilkinson did most of the editing for me and both, Mr. M. Lay and Mr H. Yuan helped me with the final formatting. The reliable and helpful support provided by the departmental assistants, Mrs. J. Liang, Mr. Y. Zhang, Mr. B. Nichol and Mr. P. Ewart, is gratefully acknowledged.

To M. Wallace, thank you

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Nomenclature

ρ	density (g/cm^3)
μ	dynamic viscosity ($N\ s/m^2$)
ε	porosity
λ	wavelength (nm)
ρ_a	absolute density (g/cm^3)
ΔC_i	concentration change during filtration (mg/L)
ρ_e	envelope density (g/cm^3)
ρ_f	density of fluid (g/m^3)
α_f	friction factor
ρ_g	grain density (g/cm^3)
Δh	headloss or hydraulic head (m)
λ_i	filtration coefficient
α_p	particle shape factor
ΔQ	differential volumetric flowrate (m/sec) ρ= density (g/cm^3)
μ	dynamic viscosity ($N\ s/m^2$)
ε	porosity
λ	wavelength (nm)
ρ_a	absolute density (g/cm^3)
ΔC_i	concentration change during filtration (mg/L)
ρ_e	envelope density (g/cm^3)
ρ_f	density of fluid (g/cm^3)
α_f	friction factor
ρ_g	grain density (g/cm^3)
Δh	headloss or hydraulic head (m)
λ_i	filtration coefficient
α_p	particle shape factor
ΔQ	differential volumetric flowrate (m/sec)
ρ_{wm}	wet media density (g/cm^3)
A	filter section area
A_a	area of vena contracta
AAS	Atomic absorption spectroscopy
AD	anno domini

A_j	area of jet of streamlines
A_o	area of orifice hole
BOD	biological oxygen demand (mg/L)
CBOD	carbonaceous biological oxygen demand (mg/L)
C_i	concentration of solids (mg/L)
d	grain diameter (m)
D{4,3]	equivalent volume diameter (m)
d_{10}	effective grain size (m)
d^3	surface area of particles (m)
D_o	bolt orifice diameter (m)
FDE	farm dairy effluent
G	acceleration of gravity (m/sec^2)
g	gram
h	hour
HMWW	Hamilton municipal wastewater
HWWTS	Hamilton wastewater treatment station
IEP	isoelectric point
km	kilometre
kW/h	kilo watt hour
$-K_p$	coefficient of conductivity
K_{vc}	coefficient of vena contracta
L	filter bed depth (m)
L_e	expanded filter bed (m)
m	meter or loading rate (m^3/m^2)
m/h	volumetric flowrate through a given filter area ($m^3/m^2/h$)
m/m	mass by mass
min	minute
mL	milliliter
mm	millimeter
mol	mole
M-tonne	mega tonne
mV	millivolt
MWWTS	Morrinsville wastewater treatment station
NTU	Nephelometric turbidity units
P	gas pressure (N/m^2)
P_0	saturation vapour pressure (N/m^2)

pa	per annum
PAC	poly aluminiumchloride
PC	Porous ceramic
PCDM	porous ceramic dual media
PEF	primary effluent filtration
PFS	poly ferric sulfate
ppm	parts per million
psi	pounds per square inch
PTE	primary treated effluent
Q	volumetric flowrate (m^3/sec)
q	velocity of flow (m/s)
q_{mf}	minimum fluidisation velocity
Re	Reynold's number
RMA	Resource Management Act
sec	second
SS	Silicon Sponge
SBR	sequence batch reactor
STD	standard deviation
STP	Standard temperature pressure
STE	secondary treated effluent
t	time
TM	Titanomagnetite
TSS	total suspended solids
UC	uniformity coefficient
UV	ultra - violet
v	velocity of flow (m/sec)
VSS	volatile suspended solids
$V\varepsilon$	bed porosity
$V\varepsilon_{Le}$	porosity of expanded bed
$V\varepsilon_{mf}$	bed porosity at minimum fluidisation velocity
v/v	volume/volume
v/w	volume/weight
V_B	bed volume (mL)
V_g	volume of grains (mL)
v_{mf}	minimum fluidisation velocity

V_p	volume of pores (mL)
V_{STP}	volume of gas adsorbed at standard temperature and pressure
w/w	weight/weight
WFS	Works Filter Systems, Ltd

Chapter 1

Introduction and Literature Review

1.1 *General Introduction*

In New Zealand, there are currently 26 sewage plants serving communities with populations greater than or equal to 20,000 people and 243 plants, serving communities with populations less than 20,000 people (Ministry of Health, 2003). Of the larger plants with a mean flow of $3x10^4m^3$/day, 10 have some form of tertiary treatment and 17 have secondary treatment. A third of the plants rely on primary treatment alone. Of the 243 plants serving smaller communities with a mean flow of $1.3x10^3m^3$/day, only about a third of them have secondary or tertiary treatment installed.

The larger plants treat mean BOD levels of 138 mg/L and mean TSS levels of 115 mg/L. Similar levels can be expected for the smaller plants. Research has shown that primary settling removes about 50 - 60% of TSS and 20 - 30% of BOD. Excessive loads of BOD and TSS leave treatment plants and enter surface waterways if only conventional primary treatment is used. Reduction of BOD and TSS in the wastewater will be required for ecosystem safeguards and adverse effect mitigation requirements of the RMA.

The Resource Management Act 1991 and increasing public pressure for improving the quality of the receiving water will require upgrading of these facilities. While many plants are likely to install aerobic digestion as a secondary treatment, the possibility exists that improvements in filtration technology will make feasible the use of filtration as an addition to the options available in wastewater treatment. It might be used as an additional primary treatment to further reduce suspended solids, or as a tertiary treatment to improve effluent quality particularly relating to turbidity.

For the treatment of drinking water, granular medium filtration is commonly used in combination with flocculation to meet health regulation standards. The application of this technology during the early stages of wastewater treatment is less common. The performance of conventional gravity filters in primary wastewater treatment has been poor due to high concentration of suspended solids, which clogs the surface of the filter bed and results in short filter runs and frequent backwashing. However, coagulation, flocculation and settling are commonly used in conjunction with filtration as part of tertiary treatment.

Recently, Hill and Langdon (1991 and 1993) reported on an improved filtration technology. Porous ceramic dual media (PCDM) filtration involves the use of porous ceramic nozzles in combination with dual media consisting of silica sand and Silicon Sponge, a modified pumice. PCDM technology has been shown to produce high flow rates and long filtration run times in the treatment of municipal drinking water (Liu and Langdon, 1997). The improved efficiency of PCDM filtration technology, particularly when used in conjunction with novel New Zealand media (pumice and titanomagnetite) may make filtration a viable option for an additional primary treatment step after primary clarification. Solids removed by such filtration might then either be digested to produce methane or processed into pellets for agricultural use or land-fill disposal. The increased solids removal would reduce the subsequent load associated carbon dioxide emissions during secondary treatment, and if additional solids were anaerobically digested, would provide a direct energy return. In the case of tertiary treatment a fine media such as titanomagnetite may allow efficient filtration without the use of flocculating chemicals.

This chapter reviews the relevant technical and political issues relating to the management of municipal wastewater. Aspects of filtration have been of interest since the late 19th century and a brief overview account of the development of filtration technology will be included.

New Zealand is a geologically young country and has abundant resources of novel granular minerals, particularly pumice and titanomagnetite. These materials have not been fully studied in filtration applications. Their usefulness in wastewater filtration will hence be a major topic of the thesis. The chapter will conclude with the objectives of the present investigation and an outline of the structure of the thesis.

1.2 *The regulatory framework for wastewater in New Zealand: the Resource Management Act (1991)*

The relatively low population density and the island status of New Zealand has made it possible in the past for the impact of poor wastewater management to stay less obvious than it would be in more developed, densely populated and industrialised countries. New Zealand's population has only reached 4 millions in 2003 (Statistics New Zealand, 2003). The Resource Management Act 1991 (RMA) drew together several interconnected statutes into a single Act with the purpose of promoting the sustainable management of natural and physical resources. Local authorities were given redefined functions to give effect to the Act including control of the use of land for the purpose of the maintenance and enhancement of the quality of water in water bodies and coastal water. In addition a general duty was prescribed for any person to avoid, remedy or mitigate any adverse effect on the environment arising from any activity carried out by that person irrespective of administration or functions of local authorities. New administrative policies of the authorities (backed by substantially increased penalties) coupled with the general duty has required private and commercial operators to re-evaluate their effects on natural resources and to control and improve the degree of pollution resulting from wastewater disposal.

In the Resource Management Act (Section 5) sustainable management is defined as meaning:

"Managing the use, development and protection of natural and physical resources in a way, or at a rate, which enables people and communities to provide for their social, economic and cultural well being and for their health and safety while sustaining the potential of natural and physical resources to meet the reasonable foreseeable needs of future generations; and safeguarding the life- supporting capacity of air, water, soil, and ecosystems; and avoiding remedying and mitigating any adverse effects of activities on the environment."
(New Zealand Government, 1991).

Local authorities, in carrying out their functions under the Act, are now in the process of rewriting regulations and updating consents involving the use of natural resources.

1.3 *Current wastewater problems*

Contaminants, such as biological oxygen demand (BOD) and total suspended solids (TSS) are one subject of attention. Increased use of land irrigation to treat effluents leads to the potential risk of spreading airborne pathogens. The recreational use of rivers has prompted a classification scheme where concentrations of pathogens are defined. Wastewater treatment technologies will have to perform to protect the public from disease causing micro-organisms. As can be seen from Figure 1-2 above, filtration may provide a solution to remove bacterial contamination. In addition to municipal waste, those responsible for managing water quality must take account of New Zealand's very large population of pastoral grazing agricultural livestock. The population of dairy cows of 1.5 million concentrated predominantly in the highly fertile areas of the Waikato and Southland produce excreta with large amounts of BOD and nitrogen (Environment Waikato, 1998).

BOD and TSS levels are also a matter of attention for primary agricultural processing industries. These cover dairy, cropping, meat, hides and wool, as well as a substantial forestry industry with its production of pulp and paper.

In the case of a livestock slaughtering plant with a daily throughput of up to 800 cattle, primary effluent concentrations reach about 2500 mg/L TSS and 3000 mg/L BOD, about 10 times the concentration found in domestic effluents. Each carcass requires between 1.4 - 2.2 m^3 water (Pattle Delamore, 2000). High TSS and BOD levels also appear in effluent from the pulp and paper industry. These are typically 40 mg/L TSS and 20 mg/L BOD (Fletcher Challenge, 1998). Another major source of high TSS and BOD loadings are dairy sheds, where about 1×10^4 mg/L TSS is produced as an average loading in wash water after milking (Longhurst et. al. 2000). In dairy product processing factories, TSS levels are similar to those in domestic effluent but in addition, they often contain a high BOD content (300 - 750 mg/L). For domestic effluent, initial levels of 150 - 500 mg/L TSS and 100 - 900 mg/L BOD can be expected (Degrémont, 1991). If these levels were to be reduced to the levels found in natural waters, TSS $\leq$ 30 should be achieved. However current background levels of BOD and TSS in the Waikato river are rather high. They range from 0.2 mg/L upstream to 1.1 mg/L downstream; and TSS ranges from 84 mg/L upstream to 120 mg/L downstream of Hamilton (Charles, 2001).

For the purposes of the present study, two important wastewaters, representing the range of non industrial wastewaters in New Zealand, will be considered: municipal wastewater and farm dairy effluent.

1.3.1 Municipal wastewater

Municipal wastewater production has been estimated at 260 L per person/day (Degrémont, 1991). Sewage treatment plants receive wastewaters from a variety of sources. They include domestic wastewater, industrial wastewater and infiltration or accidental inflow of stormwater. The percentage of wastewater components varies with local conditions and the time of the year (Tchobanoglous and Burton, 1998). The industrial proportion of wastewater generally keeps pace with population growth and joins the sewage after special treatment at the source when required.

Composition of municipal wastewater

The important components of municipal wastewater are: non-degradable suspended solids, degradable organic material, colour, dissolved organic and inorganic compounds and microbes. Levine (1985) listed the different parameters found in wastewater as those shown in Table 1-4 and described them by their size in Figure 1.2.

Characteristics used to describe wastewater

General – temperature, pH, conductivity, suspended solids, odour, alkalinity, acidity
Metals - Fe, Zn, Cu, Cd. Pb, etc.
Inorganic - Sulfate, Phosphorus, Nitrogen
Organic – organic and volatile acids, carbon, methane, phenol, oil and grease
Agricultural – eg. pesticides
Organisms – viruses, bacteria, fungi, yeasts, etc.
Visual – colour, turbidity
Oxygen Demand - (BOD and COD)
Solids: Total, dissolved, suspended, volatile and fixed.

Particles in municipal wastewaters are mostly colloidal in nature and negatively charged, thus repelling each other. Typical surface charge ranges between - 10 and - 18 mV. In the presence of microbial activity, these particles form biofilms and slimes derived from the bacteria and algae present in the wastewater (Adin, 1994).

Suspended solid concentrations in untreated domestic effluent were estimated at 720 mg/L (Metcalf & Eddie, 1991). Most particles in domestic effluent were sized < 1 μm, however the largest volume of particles was occupied with particles of > 12 μm. Most of the larger particles were filterable, but only between 20 - 50% of the smaller particles were filterable. There were few particles of the size range between 1 - 12 μm (Levine (1985).

Levine (1985) devised a graph (see Figure 1-1) to describe components found in wastewater by their size.

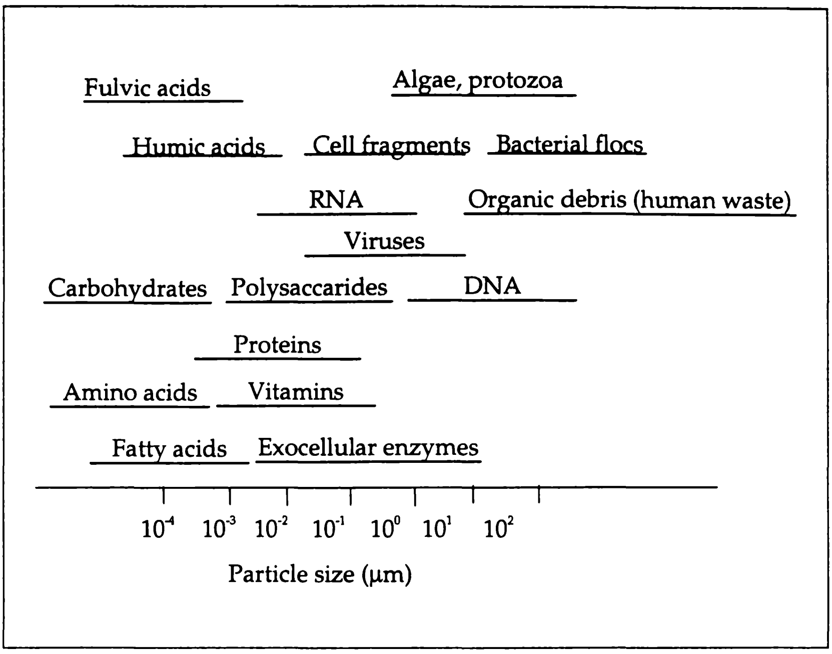


Figure 1-1 Sizes and types of organic constituents that may be present in settled municipal wastewater.

Particle size in effluent is subject to seasonal variations. To study seasonal effects on filtration performance, pilot plant studies for at least six months have been recommended by Kawamura (1999). Roth (1979) described one particular effect on particle size in secondary treated effluent, following plant overload or heavy rainfall. During heavy storm and rainfall, or at high plant nutrient loading rates, particle sizes of < 80 to 150 μm with a median value of 25.7 μm (d_{50}) were present during normal biological plant operation. However, the particle size decreased to a median value of 4.2 μm (d_{50}) during the storm.

1.3.2 Farm dairy effluent

Dairy cow numbers and herd sizes in New Zealand have experienced a steady increase from about 4 million beasts in 1995 to 4.3 million in the 1999/2000 season (Statistics New Zealand, 2001). However, dairy farm numbers have decreased, resulting in an average herd size of between 200 - 300 cows. Milk production has become more efficient through this intensification, but increasing amounts of effluent are being generated from each dairy shed. During the milking process, the daily production of diluted cow manure amounts to about 50 L per cow (Nadarajah, 1996). Unless this waste is treated, it represents an unacceptable burden to waterways and soils. Inadequate management of farm dairy effluent (FDE) can lead to a decline in surface water quality (Hickey et.al., 1989), an increase in ground water nitrate levels (Askew,

1985) and nuisance effects from odour and spray drift (Vanderholm, 1984). Another contaminant is the colour caused by dissolved solids from undigested feed matter. Concerns about environmental degradation and a move towards sustainable farming systems have led to closer attention being paid to FDE. Treatment of FDE in New Zealand is mainly through land application, or through oxidation ponds. Regulations have been imposed to limit the amount of nitrogen applied to land from FDE.

Masters (1993) described a more intensive treatment method for sandy soils in Western Australia. It comprised three steps. An anaerobic digestion pond with up to 20 days holding time was followed by slow sand filters in combination with a wetland and land-based irrigation.

A slow sand filter (vegetative filter strip) was used as the only treatment by Schwer and Clausen (1989), where it showed satisfactory results for both nitrogen and phosphorus reduction, however it was sensitive to hydraulic loading rates.

Data gathered on farm dairy effluent in New Zealand found an average dry matter content of 0.9%. Nitrogen content in FDE is related to the amount of dry matter (Barton, 1994) with levels up to 400 mg/L. The solids in FDE are made up of about 10% excreta, 4% teat wash water, 86% wash water plus other foreign matter (Gibson, 1995). Average values of phosphorus and potassium were 70 mg/L and 370 mg/L, respectively (Longhurst et.al. 2000). Goold (1980) reported calcium, magnesium and sodium levels of 177 mg/L, 39 mg/L and 54 mg/L, respectively.

1.4 *Wastewater treatment technologies*

The processes of treating municipal sewage have been broadly classified as primary, secondary and tertiary. The particular processes used in a given situation depend on the volume to be treated, the location of the outfall, the dilution factor, the potential hazard to users receiving the water, and in many cases, the cost of the project. An outline of the process options involved is shown in Figure 1-2. Products from wastewater treatment comprise sludge and treated effluent. The largest volume of sludge produced follows primary treatment. Further processing can result in the production of biogas and fertiliser.

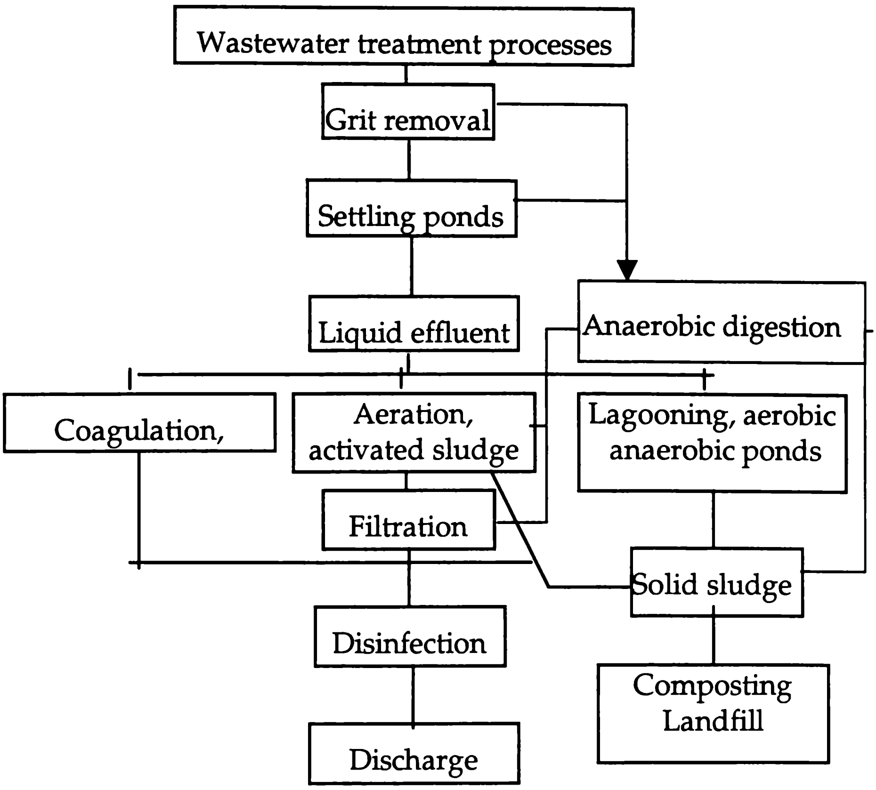


Figure 1-2 Wastewater treatment technologies.

1.4.1 Primary treatment

Primary treatment includes mechanical processes, which remove solids. Metal screens stop the larger solids and sands and small stones settle in a grit chamber. The water then passes into a settling tank, where its rate of flow is sharply reduced and smaller particles settle as sludge. Scum at the top can be removed by skimming. The quantities of TSS removed by these processes are about 10% after screening and 50 - 60% after settling (Degrémont, 1991).

1.4.2 Secondary treatment

Secondary treatment systems can consist of a variety of technologies including: stabilisation basins, trickling filters, activated sludge and rotating contactors.

Some of these systems rely on the effects of biofilms (Bryers and Characklis, 1989). For example, in trickling filters or packed bed reactors, fixed biofilms on the media surface remove carbon and nitrogen (see also Section 1.5). Biomass supported particle systems were developed more recently, incorporating a fixed lattice or foam structure in order to better define growth and detachment conditions for biological particles. Biological fluidised bed reactors comprise a fluidised bed of solid biofilm support media (0.2 mm to 2.0 mm), which can move freely in an upward flow of water, or a water and gas mixture.

In New Zealand, activated sludge systems are commonly used for secondary treatment. It involves biological processes using aerobic or anaerobic bacteria. Reactions of aerobic bacteria are generally faster than anaerobic bacteria. In addition, the wider range of organic substrates that are digested by such organisms make them more popular then anaerobic bacteria. Mechanical aeration is provided, to supply the micro-organisms with the oxygen they need. In addition, mixing achieves homogenisation to ensure close contact between the live medium the polluting elements and the oxygenated water.

A recent improvement to secondary treatment has involved the use of pure oxygen instead of air. More bacteria could so be supported in a smaller space using oxygen (Chhatwal, 1996).

During clarification of the aerobically treated effluent, about 90% of the biological matter and 90% of TSS along with about 50% of the nitrogen and 10% of the phosphate is removed. Part of the sludge removed re-enters the system to maintain higher bacterial activity.

Anaerobic systems are used less frequently to treat high strength organic wastes and sludges. Compared with aerobic processes they use less energy. However, separation between the liquid phase and the biological floc is very difficult. Products after anaerobic systems include methane and sludge. Methane is generated from methane producing bacteria (obligate anaerobes).

Both aerobic and anaerobic treatment processes can be divided into suspended growth and attached growth processes. In suspended growth processes the micro-organisms grow and degrade organic substrate while suspended. Activated sludge processes, treatment ponds and lagoons fall into this category. In attached growth processes, the micro-organisms are stabilised by attachment to a solid medium. The degradable material is incorporated into the biofilm, containing the decomposer bacteria. In these systems, aeration is achieved by spraying the wastewater over porous beds or contacting it with rotating surfaces (Reible, 1999).

Following the secondary treatment, the clarified effluent may receive tertiary treatment and chlorination prior to discharge. Sludges from both primary and secondary treatment are combined for further treatment.

1.4.3 Sludge processing

Sludge removed during primary treatment is either dried or digested in anaerobic tanks, reducing its volume to one-third. Filters, centrifuges or driers are used for these separation processes before any further treatment (Chhatwal, 1996). Sludge from primary (and also secondary) treatment can be heat dried and sold as fertiliser. However its application as fertiliser for agricultural purposes is limited in some countries by current legislation, governing heavy metals, phosphate and nitrogen concentration in soils and ground water. Where these limitations inhibit land based applications, alternatives for dealing with sludge comprise composting and thermal treatment. During composting, conditioning with lime or iron often precedes aerobic stabilisation or bio reduction (Leschber, 1997). Biological end products are stored in land fills or used in landscaping projects. During thermal treatment, sewage sludge is combined with solid waste for incineration. The energy consumed during thermal drying of de-watered sludge is never greater than 50% of the energy released during the incineration of the dry material (Lungwitz, 1997). Thermal end products can be integrated with products used in the building or roading industries (Honecker, 1997).

1.4.4 Tertiary (advanced) treatment processes

Water leaving secondary treatment is often not of sufficient quality to be used as a drinking water source. Most of the original phosphates and nitrates as well as any persistent herbicides or insecticides or disease causing bacteria and viruses are still present after secondary treatment. Eutrophication of lakes is often a sign of high nutrient levels present in wastewater after secondary treatment (Chhatwal 1996).

Tertiary treatment is able to remove virtually all the remaining contaminants but often at high costs. The four major tertiary treatment processes are filtration, flocculation, ion exchange and membrane separation. Their application depends on the degree of separation required and on the nature of contaminant. Other treatment steps include extensions of biological treatment and physico-chemical separation techniques such as adsorption, distillation and reverse-osmosis.

1.4.5 Flocculation as tertiary treatment

Chhatwal (1996) described flocculation, followed by settling of the separated floc. In a first step, lime (CaO) was added to make the water alkaline and to precipitate the phosphorus. In a second step, metallic hydroxides were added, aluminium or iron salts being the most common. In either case, inorganic phosphorus (as phosphate) was precipitated as phosphate salts of Fe^{+3}, Al^{+3}, or Ca^{+2} and the organic phosphorus compounds were adsorbed onto hydroxide floc (precipitate) formed by these same cations in alkaline solutions.

1.4.6 Flocculation - filtration

For filtration of effluent containing less than 60 mg/L TSS, flocculation in combination with filtration has been recommended. Pilot plant studies were used to determine the best technical parameters (Degrémont, 1991). Filter configuration ranges from bed depths of 350 - > 1000 mm with grain sizes between 0.4 to 2 mm, with the larger grain sizes related to the deeper beds.

Diamadopoulus and Vlachos (1996) compared coagulation and settling of secondary effluent prior to filtration with in-line addition of coagulant during filtration. Less coagulants were used when they were added in-line as a filter aid without prior settling. However, headloss became the limiting factor leading to short filter runs. In-line addition of < 0.3 mg/L Fe^{3+} or 2.5 mg/L Al^{3+} resulted in 80% reduction of Phosphorus and TSS (Boller 1977, 1980, 1981, 1984 and 1993; Nacheva, 1996; Stephenson, 1996; Namasivayam, 1992; Edwards, 1993; Diamandopoulos, 1996; Cikurel, 1996).

1.4.7 Ion exchange and electrodialysis

Ion exchange technology is used to remove ionic substances such as certain metals. A negatively charged cation exchange resin containing sodium (Na^{+}) can be used to remove lead or cadmium ions from wastewater. Anionic species such as I-, or NO_3- are removed by a positively charged anion resin, containing chlorine (Cl^{+}) (Chhatwal, 1996). Electrodialysis relies on the selective passage of electrically charged ions through membranes under the influence of an applied field.

1.4.8 Membrane filtration processes

Membranes remove organic and inorganic contaminants from liquid waste streams sourced from industrial processes or other treatment processes as by-products. Specialised applications deal with hazardous waste streams, where contaminants are removed based on their size (molecular weight) or polarity. The chemical characteristics of a membrane can be used to separate non-polar constituents in a waste stream from polar constituents.

Membrane processes can also be used as a final polishing tool to remove contaminants from a relatively dilute waste stream. For example, a membrane system can be implemented as a final polishing step on a bio-reactor. The bio-reactor can be designed to cost – effectively treat the bulk of the organic contamination, while the membrane can be designed to treat the aqueous phase prior to discharge.

In addition to its use as a separation tool, membrane technology is also used to concentrate contaminants from a dilute aqueous stream to a level that could support efficient biomass for bio remediation technologies (Cheremisinoff, 1998).

1.4.9 Other tertiary treatment steps

Other tertiary treatment steps range from extension to biological processes so that they become capable to remove nitrogen and phosphorus, to physico-chemical separation techniques such as adsorption, distillation and reverse osmosis. Activated carbon filters removed between 70 – 80% of refractory organic compounds, which usually resisted bacterial action but presented a problem to aquatic life (Cheremisinoff, 1998).

1.4.10 Effective size range of particle removal technologies

Figure 1-3 below shows the application of various treatment options for removal of a range of particle sizes (Levine, 1985).

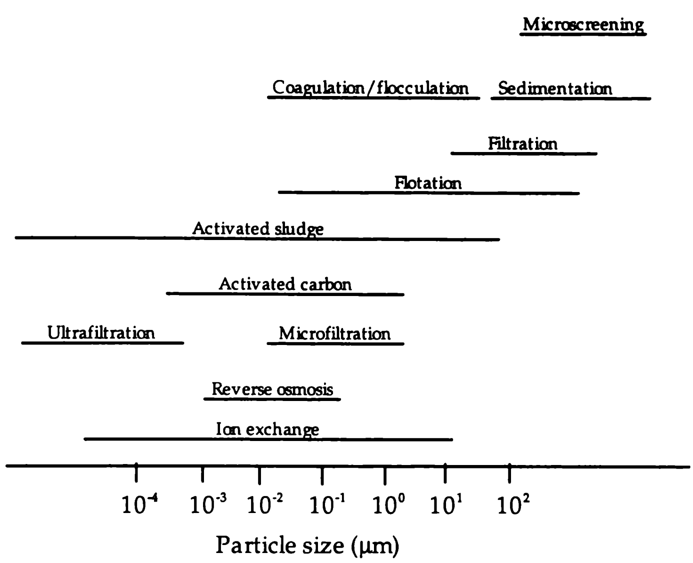


Figure 1-3. Size ranges over which various wastewater treatment operations are effective.

It is clear that granular filtration, possibly in conjunction with flocculation and coagulation is potentially able to remove most size fractions found in wastewater.

1.5 *The filtration option for wastewater treatment*

Primary effluent filtration (PEF) after settling and screening of municipal and/or farm dairy effluent may be a suitable treatment option for wastewaters containing up to 60 mg/L TSS. With average filtration efficiencies between 50 - 70% (see Section 1.11), these levels could be reduced to 20 - 30 mg/L. Current resource consent conditions for the Hamilton wastewater treatment plant specify a median TSS concentration in the discharge of 30 mg/L (Environment Waikato, 1998). If however secondary treatment is still deemed necessary, primary filtration, by removing a significant fraction of degradable material, offers a means of reducing the size and cost of the system required. Filtration of PTE has the further advantage of removing the larger and less readily degradable material, making the filtered effluent more digestible.

Farm dairy effluent containing between 0.02 - 2.6% dry matter (average 0.45%) could have this proportion reduced by settling, followed by primary filtration. A decrease in dry matter prior to land-based applications leads to reduced nitrogen loading and more likely to a compliance with current regulations for land based applications.

Previous studies have shown that filtering primary treated effluent (PTE), prior to secondary treatments allow a 10-fold reduction of costs of the overall treatment through energy savings attributed to a reduction of residual TSS and BOD by between 60 – 70%. In addition, solids captured by PTE filtration thickened better than the solids from anaerobic digesters. The smaller unfiltered particles presented an easier digestible material and the recovery of energy through digester gas utililsation proved to be a viable option (Garzonetti, 1984).

Following PTE filtration through a 1800 mm deep bed (d = 1.5 - 2.2 mm) over 450 mm gravel, a sufficient reduction of TSS was achieved to enable UV disinfection of the treated effluent. A cost comparison between chemically assisted settling of primary effluent and PTE filtration with and without chemicals, favoured the first option with only half the net value of investment required (Cooper - Smith and Rundle, 1998). Earlier, PTE filtration without chemicals was reported, using a shallow bed (d = < 0.4 mm, depth = 450 mm) with pulsed backwashing (Matsumoto et.al. 1982).

PTE filtration, with addition of two polyelectrolytes (Vitro PQ-01 and Prosifloc A-252 at 0.2 mg/L to 0.5 mg/L) was reported by Landa et.al. (1997) who recommended optimum design parameters as: 1000 mm bed depth, (d = 1.2 mm; UC: 1.6) and a flow rate of 10 m/h. Particle size analysis was carried out before and after filtration. Filtration becomes effective for particle sizes > 1 μm and is most effective for particles > 12 μm. Results from image analysing showed the effect of filtration on the composition of small particles in the effluent. Particles in unfiltered effluent showed an average length of 1.3 μm (range: 0.99 μm - 1.61 μm) and width of 0.53 μm (range: 0.45 μm - 0.65 μm). Following filtration through a shallow pulsed bed, the average size was reduced to 1.13 μm length (range: 0.66 μm - 1.56 μm) and 0.49 μm width (range: 0.3 μm - 0.61 μm). Not only small particles were affected by filtration. The major volume based effect was in the size range > 8 μm with an average of 57% removed for particles between 8 μm and 12 μm and 75% for those > 12 μm. In some cases, an increase in the concentration of filterable solids was observed for a particular size range. This might have been possible due to aggregation, or shearing of particles during filtration (Levine, 1985). Removal of larger particles, was also confirmed by Darby et.al. (1991) who used a 220 mm deep bed (d = 1.4 mm and 1.7 mm) for secondary effluent filtration. Particles between 25 μm - 40 μm were efficiently removed throughout the whole filter run, but increased removal occurred over time. Smaller particles were not removed until filter ripening occurred.

The economics of advanced primary treatment by either improved settling or filtration has gained in interest. The biofiltration process used as "Biostyr" has been reported in 1999 (Flakentoft et.al 1999). Odegaard (1998) studied methods to increase particle separation in primary settling tanks by using small doses of coagulants together with modified settling tanks. He found that most particles can be found in the size range between 1 – 8 •m but that the overall size distribution was influenced by sewer conditions, prior to the effluent plant. During flat reticulation, pre-dominantly anaerobic conditions in the sewer can lead to biodegradation and a decrease in particulate matter. Improved efficiency of pre-treatment was achieved by additions of polymers and metal coagulants. Solid removal in primary settling tanks increased from typically 50% to as much as 80%. In more recent publications Odegaard and Liao (2002 and 2003) achieved further increases in SS removal, using coarse, floatable plastic media on primary effluent and additions of low doses of coagulants. Even without coagulants, 60 – 70% of solid removal was achieved.

1.6 *History of filtration*

One of the first human cultures to use filtration as a means of water purification was the early Egyptian civilisation. Large clay jugs were filled with water and matching saucers were placed underneath them to capture the cleaned percolated water that escaped from the inside of the jug. The first documented case of a filter known to service an entire city was that of Paisley, Scotland in 1804 (Slowey, 1989). James Simpson built filters for the England based Chelsea Water Co. in 1929 to improve water quality from the River Thames (Toohil et.al. 1999). The first documented wastewater treatment "filter" was a plant built in London during 1829 to reduce pollution of the river Thames (New Encyclopaedia Britannica, 1995). A trickling filter comprising large tanks filled with stones from 100 - 250 mm average diameter was flooded with the settled wastewater. After contact of several hours, each tank was drained and much organic matter remained on the surface of the stones enmeshed in the massive bacterial growths that occurred there. Tanks were rotated on a fill-and-draw basis. Because of the labour involved with this manipulation and a desire for something better, the next big step was toward a so-called trickling or sprinkling filter (see Section 1.8).

By 1832, the first American water treatment plant, incorporating gravity sand filtration was constructed. The operation of early sand filters was limited by the amount of solids accumulating on the media. After a certain time, these filters blocked up. To avoid the need for frequent scraping of the surface of these filters, under drain systems were developed, allowing backwashing to remove the solid build up. This technology has remained largely unchanged for more than 100 years. Only after 1960 was research carried out to gain more detailed understanding of the theoretical principles governing filtration efficiencies (Slowey, 1989).

1.7 *Principles of filtration and flow through granular media*

The purpose of this section is to describe filtration in general with special reference to wastewater treatment. Technologies have changed over the years. Modifications affecting the filtration efficiency and operational considerations such as backwashing and functions of the underfloor system are reviewed. The range of filter media used varies between countries. More and less common media types are described.

The chapter starts by explaining the different filtration mechanisms. These mechanisms will provide the baseline during the planning of the experiments and in the interpretation of the results. Relevant construction details of wastewater filters follow, where problem areas, such as the underdrain and the nozzles are explained.

1.7.1 Terminology

Filtration is defined as the passage of fluid through a porous medium to remove matter held in suspension. In water purification, the matter to be removed includes suspended silt, clay, colloids and micro-organisms. The filter medium can be regarded as a barrier with an interconnected pore structure, through which the particles must pass if they have not been collected on the surface. Filtration mechanisms are complex and consist of a combination of factors. These include straining, impaction, adhesion, interception, diffusion, chemical adsorption, physical adsorption, and flocculation within the medium and gravity settling. Generally these mechanisms are additive, however they depend on the particle size of the solids to be removed (Mackie, 1993 and Rushton et.al. 1996).

Straining is the simplest of the collection mechanisms and occurs when the particle diameter is larger than the pore through which the fluid-flow streamlines pass. The grain size and grain size distribution plays an important role in this mechanism as narrower passages are found with smaller grained collection media. Straining and gravity settling in the medium pores or at the filter surface generally remove particles greater than 0.5 μm where a "cake" is formed.

Sedimentation occurs only in down-flow filtration with particles settling vertically through the flow streamlines, as they distort around the pores.

Interception occurs when the suspended particle radius is greater than the flow streamline thickness. In such instances, the suspended particle will contact the filter grain in the absence of any repulsive mechanisms.

Whenever fluid flow streamlines change directions, it is possible that particles may be less able to change direction because of their greater density. The effect is then called *inertial impaction and bounce*. If a collection surface is nearby, the particles may make contact with the grain surface or become trapped in another flow streamline.

Diffusion is only applicable for particles of diameter less than 1 μm and occurs through bombardment between the suspended particles and the surrounding fluid molecules. Particles may acquire sufficient momentum to change directions, leading in some instances to the particle contacting the grain surface.

Hydrodynamic interaction can be considered by applying the Reynold's number to the type of flow. Modifications to the Reynold's number have been proposed to account for the differing sizes of the various flow channels found in deep filter beds. During normal filtration, conditions are usually *streamline*. During backwashing, there is some advantage in achieving turbulent condition with a Reynold's number > 2000.

Chemical and physical adsorption can occur as a result of modifications to the surface charge of the filter media or by addition of chemicals. Further interactions of a physical nature include electrical double layer interactions creating attractive or repulsive energy depending on charge (Rushton et.al. 1996). The development of *biofilms* within the media can also enhance solids removal (Tchobanoglous et.al. 1970).

Literature about a flow simulator (Buganos et. al. 2001) provided further explanations to transport and adhesion mechanisms in a granular filterbed.

1.7.2 Hydrodynamic modelling

Generally a water filter is a bed of sand of either uniform or varying particle sizes. The filter bed is porous with a typical void fraction in the range of 30 - 50% (Reible, 1999).

The flow rate of a clean liquid passing through a porous bed was first reported in 1856 as Darcy's law. This law states that the driving force for volumetric flow (Q) is the head lost (Δh) in passing clear water through the filter bed (L) with a surface area (A). This loss of head can be expressed as a function of several variables such as: porosity of the filter bed (ε), average mean diameter of filter grain size (d), velocity of flow through the filter medium (q), dynamic viscosity (μ), fluid density (ρ_f) and, acceleration of gravity (m/sec^2). Darcy's Law (Darcy, 1956) is expressed as Equation 1-1.

$$\frac{Q}{A} = q = -K_P \frac{\Delta h}{\Delta L} \tag{1-1}$$

or

$$-K_P = \frac{\mathrm{q\,L}}{\Delta\,\mathrm{h\,t}} \tag{1-2}$$

The negative sign of the coefficient of conductivity ($-K_p$) indicates that the flow of water is in the direction of decreasing head. Darcy's law is experimentally based and its range of validity is determined by the magnitude of the porous medium's Reynold's number.

A commonly used model relating (Δh) q and d is stated in the Carmen - Kozeny Equation. This equation was derived from the Darcy - Weisbach relationship for Δh in circular pipes, where the roughness factor of the surface from the inside of a pipe was incorporated with the Reynold's number to obtain Δh in either horizontal or sloping directions. In the Carmen-Kozeny equation (1-3), the value corresponding to the length of the pipe was redefined as the "length over which headloss occurs", which corresponds to L. The value corresponding to the pipe diameter was replaced by an equivalent term equal to four times the value of the hydraulic radius of the flow passage. If a unit volume of the filter medium is considered, the volume available for flow is essentially equal to ε of the bed. For the entire filter bed, the channel volume is obtained by multiplying ε by the total volume occupied by the bed. Considering the granular material is usually non-spherical, it was necessary to correct the equation by insertion of a dimensionless particle shape factor α_p.

$$\Delta h = \alpha_{\mathrm{f}}\ \rho_{\mathrm{w}} \frac{L}{\alpha_{\mathrm{p}}\,\mathrm{d}} \cdot \frac{1-\varepsilon}{\varepsilon^3} Q^2 \tag{1-3}$$

or

$$\Delta h = 180 \frac{L \ v_w}{\alpha_p d^2} \frac{(1-\varepsilon)^2}{\varepsilon^3} \quad (1\text{-}4)$$

where:

Q = volumetric flowrate

ε = porosity of a filter bed

α_p = shape factor for filter grain, (1: uniform; 0.82: rounded sand, 0.75: average sand)

α_f = friction factor

ρ_w = density water

d_p = filter grain diameter

v_w = mean fluid velocity

Empirical equations have been developed, incorporating the filtration behaviour with Δh and q. These take into account the fact that during the early stages of filtration, the top part of the filter will remove most of the larger particles leaving the smaller particles to be treated in the lower part of the filter, resulting in Equation 1-5 (Mackie and Bai, 1993).

$$\frac{\Delta C_i}{\Delta L} = -\lambda_i C_i$$
$$C = \sum_{i=n}^{n} C_i \quad (1\text{-}5)$$

where:

C_i = volumetric concentration (TSS)

ΔC_i = concentration change during filtration

$-\lambda$ = filtration coefficient

A mass balance for the particles removed per unit volume of filter bed is given in Equation 1-6, (Boller, 1980).

$$Q\,t = q \int_0^t \frac{\Delta C}{\Delta L} \Delta t \quad (1\text{-}6)$$

where:

t = time

Q = mass flow

q = flow velocity

An empirical model based on the time-space removal of particulate matter in the filter bed was developed by Tchobanoglous and Burton (1998). This resulted in the Equation of Continuity (1-7), based also on earlier work by Iwasaki (1937) and Ives (1963) where suspended - solids mass balance for a section of filter A and a thickness ΔL was measured in the direction of flow.

$$\left(\frac{\Delta C}{\Delta t} + \varepsilon\,(t)\frac{\Delta C}{\Delta t}\right)\Delta V = q\,(C) - q\,\left(C + \frac{\Delta C}{\Delta L}L\right) \tag{1-7}$$

Because these models are empirical in nature, the empirical coefficients are applicable only to the filter or the influent suspension for which they were developed (Kau, 1995).

When the filtration of wastewater is considered, these equations have proven to be inadequate. The reason they do not apply is related to the characteristics of the liquid, principally the high concentration of suspended solids, which tend to clog the filtering medium in a short period of time thereby altering the characteristics of the porous medium (Rengel - Aviles, 1986).

1.7.3 Under drain and nozzles

With the development of rapid gravity sand filters, underfloor (under drain) systems have been developed which allow water distribution for backwashing operations. Under drains with an even distribution of the water and air outlet holes are an essential requirement for backwashing of a properly operational rapid sand filter, at the same time, they should preclude the passage of the filter media. In most cases, the system consists of porous plates or nozzles with holes (water orifices) for water flow in both directions. The system includes supporting gravel at the base of a filter bed.

Slowey (1989) has described three types of under drain. They included porous plate bottoms, false floor systems with nozzles attached to an under drain plenum and air grids installed at the media interface and systems of laterally aligned pipes, perforated or fitted with nozzles for direct distribution of air and water from the base of the filter.

Kawamura (1999), Dohmann et.al. (1996) and Lombard and Haarhoff (1995) warned about false floor under drain systems, that fail when bolts fixing the nozzles come loose at the grouting. Tightening of nozzle bolts has to be carried out carefully, because over-tightening can strip the plastic threads, which will eventually lead to blow out. A limitation with filter nozzles is the number of nozzles that can be fitted in the defined floor space of a filter. However nozzle bolts are often favoured, because of the relative ease of installation by simply screwing them into the false floor.

The Reynold's number identifies the type of flow through water orifices in the nozzle bolt (or stem), ie.

$$R_e = \frac{\rho\, v\, D_o}{\mu} \qquad (1\text{-}8)$$

where:

D_o = diameter of orifice

Water orifices in the nozzle bolt play a primary function as a hydraulic control mechanism. Their position in the filter nozzle is usually in the upper region of the bolt. Details of a nozzle bolt are shown in Section 5.2. The diameter of these orifices is usually smaller than the bolt tube interior diameter and the restriction provided is larger than that of the nozzle dome. Small imperfections can have a sizeable effect on the fluid velocity.

Formation of mudballs, channelling and cracks in the filter bed are all signs of inadequate backwashing processes, associated with an uneven distribution of air and water (Slowey, 1989). A variety of reasons for problems with filter nozzle performance have been described (Barth, 1995; Lombard and Haarhoff, 1995; Geering, 1996 and 1997 and Schell, 1992). Restriction to flows can be caused by partially blocked orifice holes, holes which did not get fully pressed through the stem during manufacturing and nozzle dome slots partially blocked by small filter sand grains or by manganese and iron deposits. In addition, the bond between the dome and the base can be broken, allowing media to enter the under drain plenum. One of the downstream effects of excessive filter run times is penetration of solids deep into the filter bed. Subsequent increased backwash head can lead to nozzle breakages. Chemical and thermal stability of the material used for nozzles should therefore be matched to the extreme pressures a present during backwashing.

Geering (1996 and 1997) overcame the problem of blocked slots in the dome, with a new type of nozzle dome. This dome consisted of perforated lamellae and an increased number of slots aligned in horizontal direction. Additions of chlorine (10mg Cl_2/L) during backwashing prevented formation of manganese-and iron oxides which otherwise formed deposits as hydroxides in the slots of the nozzle domes (Schell, 1992).

1.8 *Granular filtration technologies*

The different systems of operation in granular filtration technology relate to the flowrate and the flow direction during the filtration process.

Specific types of filters, operating under slow flows include trickling filters, roughing filters and slow sand filters. For filters under rapid flow, the types include single and multi media rapid gravity or up-flow filters.

1.8.1 Trickling filters

A trickling filter is not a filter in the usual sense, but a bio-reactor consisting of a large shallow concrete tank filled with medium sized stones over whose surface the settled wastewater is allowed to trickle, draining from the top to the bottom of the unit. Such filters are operated intermittently so that air has free access to films of zoological growths formed on the stones. It is this biofilm that accomplishes the oxidation of material. Microbiological activity rather than physical straining through the granular media are responsible for the removal of solids in these filters. Plastic based media (polyethylene or polypropylene) have been manufactured to replace the rock media. They come in various shapes such as strips, wheels or blocks. Strips hanging from the top can eliminate the need for an under drain to hold blocked media (Riddell, 2002). Cleasby (1991) has listed limitations to optimum performance. They include: low turbidity not exceeding 5 Nephelometric Turbidity Units (NTU), absence of algal blooms and low levels of iron (< 0.3 ppm) and manganese (< 0.05 ppm). Slow sand filters and roughing filters, are both suitable candidates as pre-cursors for trickling filters.

1.8.2 Roughing Filters

Roughing filters usually work in up-flow and in horizontal direction. They can be used to treat wastewater containing up to 150 mg/L TSS. Gravel size in roughing filters ranges between 3 to 25 mm (Cleasby, 1991 and Boller, 1993). Bed length is between 1 to 4 m and flowrates are typically between 0.75 to 1.5 m/h. The removal of TSS reached as much as 98% for particles > 1 μm. In roughing filters, solids settle vertically to the lower layers and can be removed by frequent drainage.

1.8.3 Slow sand filters

Slow sand filters can achieve up to 80% TSS removal from wastewater, using a filter bed depth of 600 mm and a grain size of 0.28 mm (Kim et.al. 1999). Dosage flowrate is usually around 0.1 m/h, applied every 6 h at a load of 0.08 m^3/m^2. In recent times, there has been an increased application of slow sand filters for Giardia control in the drinking water for communities of fewer than 10 000 persons (Sims and Slezak, 1991). Table 1-1 below lists the details of the process parameters for slow and rapid gravity filtration.

Table 1-1. Comparison between slow and rapid flow sand filters.

Operational Performance	Slow Sand filters	Rapid Sand Filters
Filtration rate	0.1 m/h	10 m/h
Water level above surface	1.5 m	1.5 m
Retention time above sand	15 h	9 min
# Retention time in bed	3.2 h	2 min
Bed depth	500 – 900 mm	500 – 1500 mm
Grain size (d)	0.15 - 0.3 mm	0.35 - 2 mm
Under drain system	Perforated plastic pipe or open tile	Nozzles, porous tiles or plates
Cycle length	1 - 6 months	1 - 4 days

Note: (1) rapid sand filters are cleaned by backwashing after each cycle, whereas slow filters are scraped after each cycle to remove the clogged surface layer of 10 to 20 mm depth.
(2) Coagulants are never used in slow sand filtration.
Based on pore volume, using a porosity of 0.4 (Sims and Slezak, 1991).

1.8.4 Rapid gravity filters

Particulate matter is collected and retained in the inter-granular spaces throughout the upper part of the bed and in some cases throughout the entire bed. This retention results in increasing resistance to flow (Degrémont, 1991). Optimum operation of a filter is achieved when the maximum available head is reached just before solid break through occurs. Rapid gravity filters can be operated at a constant flowrate (increasing head) or at a constant head (declining flowrate) both as deep bed and shallow bed filters. Most plants in New Zealand operate in the constant flowrate mode, regulated by a valve at the outlet. Constant flowrate operation simplifies the overall hydraulic control of a water treatment plant. In declining flowrate filtration, a header tank provides a constant head. During this type of operation, no attempt is made to keep the flowrate constant. Flow is greatest at the commencement of a filter run. The use of declining flowrates can lead to cost savings due to overall higher filtration rates, provided there are a number of filters operating at different ripening stages and the filter medium is coarse. The control of dosing pumps, using declining flowrates, requires increased effort in staff training (La Roche, 1989).

Other flow regimes include pumping and pressure vessels to generate faster flowrates, longer filter runs or flow in horizontal or upwards direction.

1.9 *Granular filter media*

This section provides an understanding of existing filter media and a comparison to titanomagnetite (TM) and the silicon sponge (SS) medium, which is proposed for this study.

It also introduces existing literature on the use of pumice as filtration medium and on TM sands in general.

Materials used as filter media must possess good hydraulic qualities and have good filtration characteristics. They must be hard and durable, free of impurities and insoluble in water. Naturally occurring minerals and rocks are graded to a desirable size range and used without any further processing or they can be chemically, thermally or magnetically modified.

The medium most commonly used in municipal water filtration is silica sand. It is economical and typically used at a grainsize between 0.5 - 0.8 mm. Anthracite may be used to replace the sand as a filter medium or it can be placed on top of sand. The size of anthracite can be twice as large as sand. Separation after backwashing of the media is facilitated by anthracite's lower density, allowing sand to settle in place before the anthracite. The lower headloss and reduced usage of wash water for anthracite has to be considered in relation to the higher loss of medium due to abrasion (Salvato, 1992). Instead of anthracite, coal can also be used as a top layer. As is the case for anthracite, the low density of coal and its high friability requires more careful backwashing procedures to prevent losses (Cheremisinoff, 1998 and 1995).

Modification processes to the media are described in Appendix 16, which also lists details on specialised media types and their application.

1.9.1 Single dual and multi media filters

This section introduces the possibility to place different media types in layers. It provides the background knowledge to alter media configurations for TM and the SS.

Rapid gravity filters can be designed to use a single medium or a combination of two or more media. For single homogeneous medium filters, no special hydraulic expansion is necessary during backwashing. In single medium filters with heterogeneous filter material, a screening effect of the filter material occurs during hydraulic expansion. Separation into the coarsest grains at the bottom and finer grains at the top occurs, leading to shorter filtration cycles. Dual or multi media filters have been designed to avoid these screening effects. In multi media filters, media are arranged in layers with decreasing grain size and increasing density from top to the bottom. The main advantage of dual or multi media filters is longer filter runs and greater solids holding capacity. Also mudballs formed in the filter remain above the interface to the second layer where they are subject to auxiliary scrubbing action (Cheremisinoff, 1995). No significant difference between single and dual media filters was found when solids removal efficiency was compared (Dahab and Young, 1977). Materials used as top layers in multi media filters include anthracite or coal and the lower layers typically contain finer material such as silica sand or garnet. Grain size (d), uniformity coefficient (UC) and density (ρ_g) have to be carefully chosen to allow a similar expansion of each medium during the backwashing operation. For multi media filters, UC between 1.5 - 2 have been recommended (AWWA, 1972). Common media configurations include tri-media comprising 450 mm anthracite (d = 0.9 to 1.1 mm) over 200 mm silica sand (d = 0.45 mm to 0.55 mm) and 100 mm garnet (d = 0.2 to 0.4 mm) at the bottom (Kenner, 1987; Salvato, 1992; Ongerth and Pecoraro, 1995; Tchobanoglous and Burton, 1998). For filtration of flocculated tertiary treated effluent, a tri-media filter containing 700 mm pumice (d = 25 - 35 mm) over 500 mm anthracite and 200 mm sand (d = 0.5 - 1.2 mm) was used (Boller, 1980).

1.9.2 Novel New Zealand filtration media

Information relating to the source of both TM sands and river pumice, which is modified into SS, are the key elements of this section. Understanding the local abundance and uniqueness of the material further underlines the potential for their application as promising filtration media.

River pumice

River pumice is highly weathered pumice mined from the lower reaches of the Waikato River.

Pumice is the common name for a natural vesicular aluminosilicate glass formed as a result of volcanic activity. The vesicles occur in a great variety of shapes and sizes. The most common shape is ellipsoidal, with the vesicles in the direction of elongation. Minerals at, or near the surface of the flat plane of these grains are susceptible to abrasion (Tilly, 1987). New Zealand pumice originates from the Taupo volcanic zone, which is one of the world's most active silicic volcanic systems (Briggs et.al. 1993). The most recent eruptions occurred between 70 AD and 180 AD. These eruptions exploded some 25 km^3 pumice – ranging from ash (4 - 0.25 mm), lapilly (32 - 4 mm) and block (> 32 mm) sizes. Deposits extended North and South over an area of 160 km^2 (see map of New Zealand in Figure 1-1). Most of the ejecta became deposited in river valleys, forming dams, leading to temporary lakes. Abrasion processes occurred following the breaking of these dams. Processes of fluviatile and lacustrine attrition then followed, transforming Taupo air-fall pumice into alluvial pumice with granule and sand sizes of 0.06 - 4 mm and a more robust surface. Crushing and liquid separation were used to identify minerals from the pumice. These minerals were phenocrysts, plagioclase (usually andesine), quartz, hypersthene, magnetite, rhyolite, obsidian, claystone, argillite, ignimbrite and greywacke (Tilly, 1987).

There is little literature on the use of pumice as a filter medium, in many cases, medium specifications include a requirement that the medium be pumice free because of pumice's reputation of being prone to abrasion and the tendency of pumice fines to block filters (Boller, 1996).

Some applications include the use of pumice as oil absorbent, or as substrate for biofilm: Eyraud (1988) reported on the use of white pumice for removing oil contamination. An oil absorption capacity of 40 - 41 mL oil/100g pumice was found.

This capacity was attributed to its large surface area and the vesicular texture. Pumice was used in oil contamination control (Elsenhaus, 1982) involving a 200 mm deep bed, pre-treated with active coal and aluminium sulphate. Patents by Masato, (1981) and Thoemel, (1980) described the use of pumice medium as a substrate for biofilm in trickling filters. Growth of biofilm on pumice was also utilised in a moving bed to remove algae and other organic matter from lake water under aerobic and anaerobic conditions (Muccioli, 1998). Deep bed pumice filters were used for treatment of industrial wastewater, such as pectin manufacturing and in swimming pools (Mörgeli, 1979; Tanaka et.al. 1989; Fujimatsu, 2001).

Works Filter Systems (WFS) of Hamilton New Zealand have taken advantage of the relatively hard characteristics of abraded New Zealand river pumice to develop the proprietary filtration medium Silicon Sponge (SS). River pumice is thermally modified and has been successfully used in a porous ceramic dual media (PCDM) water filtration technology developed by WFS (see section 1.13). It is of interest to test the suitability of the SS medium as substrate for biofilm growth in view of possible further applications, similar to those reported above

Titanomagnetite

Ubiquitous to New Zealand, TM sands are "ironsands" belonging to a complex oxide class. Ironsands are intermediate members of the isomorphous series of solid solutions of magnetite (Fe_2O_4) and titanium oxide (TiO_2), with ultra-thin (about 1μm) lattice intergrowths. These intergrowths fall into the following groups: 1) ilmenite-geikielite, 2) spinel - magnetite and magnesian magnetite, 3) spinel-pleonaste and 4) spinel - hercynite (Krasnova and Krezer, 1995). The New Zealand titanomagnetite sands used in this study are members of the magnetite-ulvospinel series with minor elemental substitution of aluminium, magnesium and manganese within the cubic lattice (Graham and Watson, 1980). By separation of TM grains into magnetic and non-magnetic grains, Lawton and Hochstein (1980) found a larger proportion to be non-magnetic. Density of these sands correlated with magnetic susceptibility where higher density correlated with higher volume % of magnetite.

Dasler (1980) and Graham and Watson (1980) listed the following minerals as major and trace components of the Taharoa ironsands: ilmenite, augite, rutile, sphene, zircon, pyroxenes, amphiboles, vermiculite, biotite, quartz and feldspar. Lawson and Hochstein (1980) attributed high density to hornblende and augite. Most of these minerals were found both as free crystals and as inclusions in TM grains. TM grains can be half ilmenite, which is similar in appearance to TM but less magnetic (Prokhorov, 1980). It has been reported that the ilmenite tends to replace TM as the principal iron mineral northwards from Waikato heads (Hamill, 1985). Another trend was observed by Dasler (1980), who reported an increase in the moL percentage of FeO and TiO_2 and a decrease of Al_2O_3, relative to Fe_2O_3 with sample age.

The West Coast of New Zealand's central North Island has large reserves of ironsand, see Figure 1-4, containing between 5 - 40% (by weight) of TM minerals (Watson, 1979). An Australian mining company (BHP New Zealand Ltd.) is mining beach and dune sands to produce 1.2 million tonnes pa of TM concentrate, using a combination of both magnetic and gravity methods to separate TM from the bulk of the ironsand grains (Stokes et.al. 1989). Most of the TM is then used in the steel production at Glenbrook. Works Filter Systems of Hamilton, New Zealand uses this concentrate in the production of filter nozzles. It is also proposed to trial its effectiveness as a filtration medium.

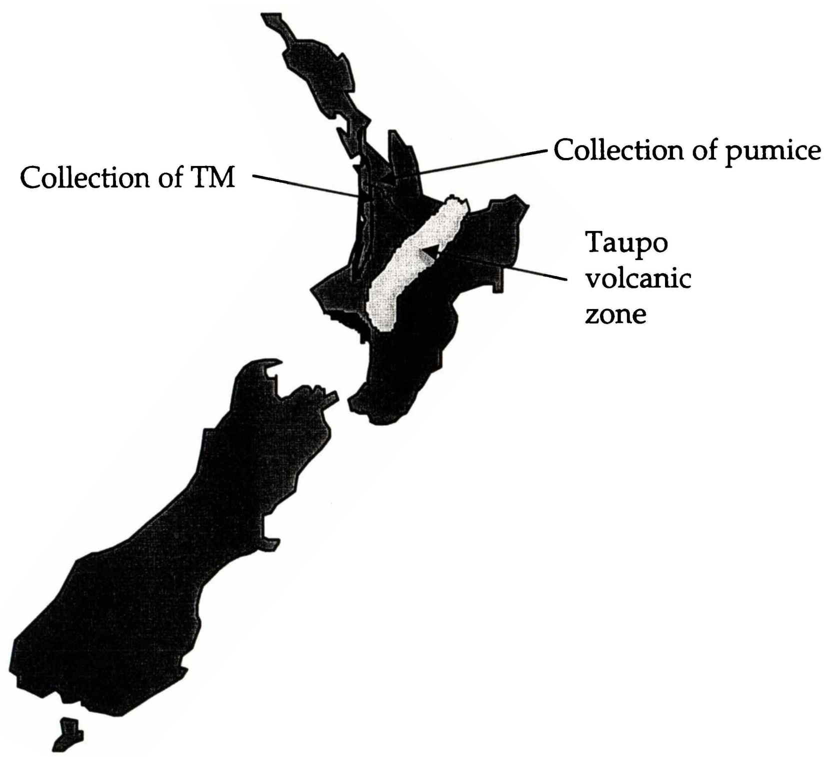


Figure 1-4. Map of New Zealand, indicating the source of river pumice and TM.

The general area of pumice is shown as □ and that of iron sand deposits as ■ (Stokes et.al. 1989; Briggs et.al. 1993).

There is no current literature on the use of TM sands as filtration media. An application in wastewater treatment of a similar magnetic mineral, magnetite (Fe_3O_4) can be found in the Sirofloc™ process, where it is used as a flocculant. Magnetite has also been used as an adsorption media for negatively charged particles, following chemical pre-treatment (Mc Rae and Evans, 1983 and 1984; Dixon; 1985, Mc Rae, 1986).

1.10 *Flocculants and filter aids*

Flocculation is often used in combination with tertiary treatment in wastewater applications. Where flocculants were added to primary treated effluent, removal of TSS and BOD improved by 30%. In addition, 90 to 99% of the bacteria and viruses and more than 90% of protozoa and phosphate can be removed with the settled sludge, following flocculation (Shao et.al. 1998). Following settling, flocculation - filtration, improves the efficiency of deep filters and can reduce residual phosphorus levels significantly (Boller, 1980). Flocculation is achieved by the addition of flocculants, filter aid or polyelectrolytes. Prior to each application, optimal dosage is experimentally determined.

1.10.1 Flocculants

A flocculant is generally an organic or inorganic polymer, which facilitates the agglomeration of destabilised suspended particles into micro-floc and later into bulky floccules, which subsequently settle as flocs. The most common flocculants and coagulants used are aluminium or iron salts as chlorides or sulphates. Application rate and timing are determined by flocculation or jar tests. Flocculation only occurs following coagulation by which neutralisation of the negatively charged surfaces of the suspended colloids occurs with cations from the inorganic coagulant. The use of polyelectrolyte flocculants facilitates the formation of a dense sludge, which can be more easily de-watered (Degrémont, 1991).

Less common flocculants include sodium aluminate, pulverised limestone, bentonite, clays, or cationic starch (Jacob et.al. 1996). Sodium silicate and polyelectrolytes are used in conjunction with aluminium salts to improve floc strength, resulting in less sludge and lower chemical dosages. Ozone has also been used for the same purpose, reducing the amount of aluminium salts required (Salvato, 1992).

1.10.2 Filter aids

Filter aids are finely divided solid materials, consisting of hard, strong particles that are incompressible and have the ability to interact with small particles by adsorption or flocculation. The most common filter aids are diatomaceous earth, expanded perlite, fly ash or carbon. In some instances, asbestos, cellulose or sawdust has been used. Filter-aids may be applied in two ways. The first method involves the use of a pre-coat filter aid, which can be applied as a thin layer over the filter bed prior to a filter run. This pre-coat prevents finely suspended particles from becoming entangled in the filter bed and facilitates the removal of the surface layer at the end of the run. The second method involves the incorporation of a certain amount of the material with the suspension before it enters the filter. In each case, the addition of filter aid increases the porosity of the sludge, decreases its compressibility and reduces resistance of the surface layer (Cheremisinoff, 1995).

Diatomite is the most important filter aid in terms of the volume used. It can be processed to different size grades. It has a high solid holding capacity and a low hydraulic resistance.

Perlite is the purest filter aid and it is therefore most commonly found in food processing applications.

Powdered carbon can be used in its activated and non-activated form. As non-activated powder it can be used in suspensions to help buffer aggressive liquids, in its activated form, it is often used for decolouring (Salvato, 1992).

Fly ash is primarily used to de-water sewage sludge by application of a 50 mm thick pre-coat on top of coarser filter media.

Cellulose and asbestos are both used to cover metallic cloth filters. Their high cost and environmental concerns (regarding asbestos) make them a less common filter aid (Cheremisinoff, 1995 and 1998).

The use of fine magnetite sand (Fe_3O_4) as a recyclable filter aid was the only published application of a material similar to TM in wastewater treatment. It was used as a flocculant, because of its faster settling times and reduced environmental impact

(Hencl et.al. 1995). Naturally occurring magnetite was ground to fine particles (d = 1 - 10 μm) and introduced by seeding it into the water. After separation, the magnetite was recovered and re-used in the process water (Kolarik, 1980; Kolarik et.al. 1994). This technology (Sirofloc™) is based on the attachment of non-magnetic pollutants to a magnetic carrier material. A positive charge appears on the surface of magnetite in acidic media at a pH below the IEP and as a result, negatively charged particles of impurities (eg. clay, silica, bacteria and viruses) contained in untreated water are attracted to the surface. At a later stage in the process, alkaline conditions above the IEP causes a negative charge on the magnetite surface and therefore the negatively charged impurities are repelled from where they can be collected and disposed as sludge (Anderson and Priestley, 1983; Dixon, 1982). A similar technology was used in combination with aluminium or with iron based coagulants for sewage treatment. Enhanced floc growth and settling rate of slurry lead to shorter plant residence time (Booker, 1994; Kolarik et.al. 1994; Kolarik, 1980; Terashima et. al. 1986; Spevakova, 1995). This study will explore the application of TM as filter aid.

1.11 Factors affecting filtration efficiency

Understanding of the key factors affecting the filtration efficiency is essential for the planning of filtration experiments.

The effective grain size and the uniformity coefficient (UC) are the principal characteristics that affect the filtration operation. The UC is defined as the 60 percent size d_{60} divided by the 10 percent size d_{10}. The effective size, $d_{10,}$ is defined as the sieve size that allows the smallest 10 percent by weight of the filter media to pass. Further variables affecting the efficiency of rapid gravity filters are bed depth, flow rate and the quality of the incoming effluent. The qualitative dependence of various parameters on filtrate quality and filter run time is summarised below in Table 1-2 (Degrémont, 1991).

Table 1-2. Effect of grain size, bed depth, flow rate and available headloss on filtrate quality.

Affected result	Grain size	Bed depth	Filtration rate	Available headloss
Quality of filtered Water	↘	↗	↘	=
Filter run	↗	↗	↘	↗
Bed loading	=	↗	=	↗

Note: ↗ Indication of the nature of the filtration response to the variable change.

1.11.1 Effect of bed depth:

The function of bed depth has to be considered in relation to the grain size, the filtration rate and the size of the particles filtered. For grain size with d < 1 mm, filtration performance improves with bed depth for those bed depths up to 900 mm (Tchobanoglous and Eliassen, 1970; Dahab and Young, 1977). When solid particle break-off occurs in the top layers, recapturing relies on filter bed depth availability (Lawler et.al. 1993).

Optimal bed depth for coarser media has been expressed as the ratio of bed depth (L) to grain size (d). L/d ratio of 1000 was recommended for rapid sand filters with a grain size of d < 1 mm. This ratio increased with the size of filter grains to L/d of 1300 for coarser media (d = 1.2 - 1.4 mm). For media grains with d > 1.5 mm, the space between the grains becomes larger compared with the void space in regular filter beds. For these media types, the L/d ratio should be used as an estimate only (Kawamura, 1999).

Dahab and Young, (1977) did not find any effect on filtration performance when bed depth increased from 457 to 1375 mm. The filter media were silica sand (d = 1.5 mm) and coal (d = 1.3 mm) with flowrates of 5 m/h and of 10 m/h. Deeper beds were only beneficial when flows exceeded 20 m/h. Solid break-through occurred in shallow filter beds of < 450 mm depth for both media types

1.11.2 Effect of flowrate

A negative effect on filter performance has been observed when flowrates exceeded 10 m/h. At lower flows, there was little effect on filter performance (Stevenson, 1997; Ching - lin et.al. 1996). This is consistent with earlier work by Tchobanoglous and Eliassen (1970) who, using secondary treated effluent, found that filtration performance was not greatly affected by increasing flowrates from 5 to 10 m/h. The effect of flowrate became more pronounced at flowrates > 10m/h. For smaller filter grains (d = 0.5 mm), increasing the flowrate led to less TSS removal, eg. 40% TSS removal at 5m/h or 30% at 20 m/h. For larger grains (d > 1 mm) increasing flows reduced TSS removal from 20% for 5 m/h to 10% for 10 m/h. In the case of flocculated, low turbidity river water, Stevenson (1997) reported that efficiency was unaffected by the flowrate up to 7 m/h for d = 0.6 to 1.18 mm. However once the limiting shear was reached, the efficiency became flow dependent and break-through of solid particles occurred earlier.

Brown and Wistrom (1999) tested the effect of increasing flowrates for primary effluent filtration. Three different filter types were used. They were a 750 mm deep bed of porous spherical synthetic balls (d= 40 mm) in up-flow mode (Fuzzy filter); a 250 mm shallow bed of silica sand (d= 0.45 mm) in down-flow mode and a 2 m deep coarse sand bed in continuous up-flow mode (Dynaflow). Flowrates tested, were: 50 to 93 m/h for the Fuzzy filter, 5 to 14 m/h for the shallow bed and 3.4 to 23 m/h for the Dynaflow filter. In each case, increasing the flowrate reduced the performance with TSS removal from 70 to 40%. The negative response in filtration behaviour, including early solids break-through was confirmed by Matsumoto et.al. (1982) with a shallow bed and fine filter grains.

Conflicting with these results is the work by England et.al. (1994), using a Dynaflow filter (1770 mm deep bed of silica sand, d = 0.9 mm, UC = 1.8). They found, that for primary treated effluent, increasing flowrates from 5 m/h to 15 m/h had no effect on filtration efficiency. From the initial 30 mg/L TSS, 70% were removed and from the original 42 NTU turbidity, 46% were removed.

Removal efficiencies were independent of flowrates, after the flowrate was normalised, by relating the contact time to the media surface to a decreasing flowrate during a filter run. Samples were taken at a greater depth when flowrates were fast and compared with shallower depths when flowrates were slow (Kau and Lawler, 1995).

In the US, regulatory agencies restrict the maximum flowrates for specific filter types. The maximum rate for sand filters is 7.5 m/h and for dual media filters it is 10 – 12.5 m/h. Higher rates may be permitted where design engineers present satisfactory results from pilot plant studies. No such limitations are currently in place for New Zealand (Jamieson, 2002).

1.11.3 Effect of media type

Surface area, surface composition and grain shape are more important factors in determining filtration efficiency than is the media type. Also these factors most affect the filtering performance during the ripening process when adsorption mechanisms of fine particles are dominating (Ching - lin et.al. 1996). After this time the surface becomes coated with effluent particles, which provide their own capture mechanisms for any further solid removal.

In some instances, enhanced filtering performance was found when comparing activated carbon with anthracite. Spherical glass beads gave lower efficiency than angular sand grains. Grain surface area to filter bed volume ratio was a possible factor in initial capture efficiency (Kau and Lawler, 1995).

Dahab and Young (1977) found no difference when comparing coal with sand of the same grain size in single medium filters or arranged as a dual media filter. A 75% TSS removal was obtained for secondary treated effluent for each media type.

1.11.4 Effect of grain size

Generally higher filtration efficiency is achieved with smaller grain size. Large media captures particles throughout the filter bed relying on the total surface area available. Therefore, the effect of bed depth is evident. For smaller media (d = < 0.8 mm), most removal occurs over the top 40 mm (Clark et.al. 1992; Kau and Lawler, 1995).

Filtration efficiency was doubled by reducing grain size from 0.65 mm to 0.35 mm (Matsumoto et. al. 1982) or from 1 mm to 0.5 mm (Tchobanoglous and Eliassen, 1970) when filtering primary treated effluent with TSS ranging between 70 to 106 mg/L.

There has been little systematic study of the effect of grain size on the filtration of primary treated effluent through coarse grain deep filter beds. Grain sizes less than 1 mm had generally been used.

1.11.5 Effect of uniformity coefficient

The uniformity coefficient (UC) influences performance mainly in dual or multi layer filter beds. Intermixing between the different layers occurs during backwashing where the UC in each media is high. With lower UC, media separation can be maintained and controlled by differences in density. Where smaller filter grains accumulate towards the top of the filter bed, shorter filter run times and increased headloss result. It is therefore desirable to keep smaller, high-density filter grains at the lower layers of a bed. UC of 1.3 to 1.5 (Kawamura, 1999 and Matsumoto 1982) or < 1.8 (Degrémont, 1991) are usually specified.

1.11.6 Effect of charge

Filter media used at any pH below the IEP will be positively charged and will therefore attract negative particles, such as clays and negatively charged micro-organisms (Yang, 2001 and Shaw et.al., 2000).

1.11.7 The development and effect of biofilms

In view of the possible application of silicon sponge as substrate for biofilm, see Appendix 17, this section outlines how the formation of biofilms can change surface characteristics and filter bed porosity, hence affect the overall filtration performance and leading to biofouling (Ching - lin et.al. 1996; Schell, 1992; Characklis and Marshall, 1989; Cunningham et.al. 1989). Biofilm formation and deposits of flocculation chemicals can also alter the grain shape and can reduce the flow required for fluidisation during backwashing.

A biofilm consists of cells immobilised at a substratum and frequently embedded in an organic polymer matrix of microbial origin. This can result in a uniform or a patchy coverage from less than a monolayer of cells to a multi-layer of up to 30 - 40 mm thick in algal mats. Biofilm developments are manifest in more or less desirable forms; in controlled conditions, biofilms are used for the extraction and oxidation of organic and inorganic contaminants from water and wastewater (eg. rotating biological contactors, biologically aided carbon adsorption). Bio–hydro-metallurgy is an operation, whereby minerals and ores are extracted from mediated biofilms. Some bio-technological processes rely on immobilised biofilms for improved productivity and stability.

Less desirable effects can occur following biofilm accumulation in water distribution and wastewater treatment applications and in cooling towers. The detachment of micro-organisms from biofilms can lead to an increased health risk. In laboratory and field observations, maximum biofilm accumulation occurred immediately inside the column entrance and diminished rapidly within a short distance down stream, probably due to a decrease in substrate and oxygen (Raiders et.al. 1986). They concluded that the rate of biofilm growth was strongly influenced by transport characteristics, including pore velocity distribution, surface roughness and other variables that affect the delivery rate of substrate and nutrients to the growing cells.

By introducing 10 mg/L Cl_2 during a backwash cycle the formation of biofilm was reduced, but conflicting conclusions on the role of biofilm in porous media were reported by Miltner et.al (1995), who recognised the importance of a residual biofilm and warned against the use of chlorine.

During the measurement of biofilm growth, attachment and detachment mechanisms have to be considered in order to establish their influence as bioprocessors or as possible cause for downstream contamination or blockages. Kooji et.al (1995) developed biofilm measurements inside pipes, where Ohashi and Harada (1995) measured biofilm adhesion strength.

Biofilm formation in pipes was measured by van der Kooji et.al. (1995). They compared teflon with glass cylinders which were exposed to treated drinking water over a period of 125 days under dark and anaerobic conditions. Accumulated biomass was removed from the samples by immersion in an ultrasonic cleaning bath. Plate counts and colorimetric methods were used to evaluate the biofilm. Their work showed that material type (glass and teflon) and position in the streamline had minor or insignificant effects on biomass accumulation.

Ohashi and Harada (1995) measured biofilm adhesion strength. Polyvinyl-chloride plates were immersed into a recycling effluent model for a period of 99 days. The growth and adhesion of biofilm was monitored during that time. Conditions were anaerobic. Tensile adhesion was measured, using a centrifugation device, with the plates fixed at the outside of a rotating arm. The plates were secured to the arm in such a way that the biofilm was on the outside and was caught by an appropriately sized box upon centripetal acceleration. Creating an impact on down-hill collision tested shear adhesion. Biofilm thickness and volume was measured and photos taken to assess the impact of centrifugal and collision force. The tests showed that shear strength was consistently larger than tensile strength. Cavities (biofilm absent space) formed on the substrate interface towards the later stages of the experiment, resulting in a slight decrease of adhesion strength.

There is currently no literature available on the formation of biofilm on filtermedia. Quantitative assessment will form part of this study.

1.11.8 Filtration efficiencies for primary and secondary treated effluent

Filtration of un-flocculated primary treated effluent with initial TSS levels between 40 and 100 mg/L and turbidity between 40 and 80 NTU has been studied. Filtration efficiencies expressed as TSS removal ranged between 30 and 70% and expressed as turbidity removal ranged between 20 and 70%. Media used included silica sand, coal and porous latex spheres. The following grain sizes were used: sand or coal d = 1 - 2 mm (Dahab and Young, 1977); sand d = 0.35 - 0.65 mm (Matsumoto and Weber, 1988); sand d = 0.9 mm (England et.al. 1994) or sand d = 0.45 mm; porous latex spheres d = 40 mm (Brown and Wistrom, 1999); polyethylene cylinders with finns wheels d= 7 mm (Odegaard and Liao, 2002 and 2003).

In a trial with un-flocculated STE, a dual media filter bed, comprising 1400 mm deep pumice (d = 2.5 - 3.5 mm) over 400 mm silica sand (d = 0.7 - 1.25 mm) produced 85% removal of TSS (Hegemann et.al. 1997).

1.11.9 Ripening and break-through

A filter run can be divided into three stages. Immediately after backwashing the filter bed operates as a clean bed where particles are removed only by attaching to the surface of the media. After that stage, filter ripening occurs, until finally a net solid removal of close to zero leads to the termination of a filter run, whereby solid particles start to re-appear in the filtrate.

Trapped backwash solids present during the first stage of a filter run can have a negative effect by contributing to higher TSS in the initial filtrate (Herbert et.al. 1995). The duration of this first stage depends on the filters holding capacity and may last for up to 30 min (England et.al. 1994; Lawler et.al. 1993; Herbert et.al. 1995).

During the second stage (ripening phase) initially captured particles serve as additional collectors for incoming particles. Lawler et.al. (1993) and Kau and Lawler (1995) studied size discrimination of particles during the ripening stage and found that the larger sizes accumulated in the top layers of the filter bed and the smaller sizes throughout the bed. An increased removal of smaller particles occurred during a filter run and a decreased removal of larger particles. Sometimes larger particles (> 9 μm) detached themselves from the top 40 - 200 mm of the filter bed to be re-captured again at the bottom of the filter bed or to be washed out entirely. Filtration efficiency for smaller particles (0.9 - 2.7 μm) continued to increase with filter ripening. Through increased amounts of solids deposited, pore diameter was reduced, which led to improved attachment properties for the smaller particles (Boller, 1993).

The decrease in effluent quality with time signals the third stage, where break-through or detachment of previously captured particles occurs. It has been shown that velocity increases lead to earlier break through, especially for angular media and shallow beds (Dahab and Young, 1977; Matsumoto et.al. 1982; Kau and Lawler, 1995).

Break through and ripening are not necessarily separated in time; the removal of some sizes of particles can increase while other sizes decrease (Clark et.al. 1992; Lawton et.al. 1993).

When using a flow simulator, Buganos et.al. (2001) provided further explanations to transport and adhesion mechanisms in a granular filterbed.

The penetration of solids into slow sand filters was monitored by Ellis and Aydin (1995). For a 1200 mm deep bed, the highest levels of solids were measured on the surface with penetration to about 300 mm. Solid content was higher at 150 mm depth compared with 50 mm depth but decreased to a low constant level below a depth of 400 mm. Penetration was not affected by media grain size variations between 0.17 to 0.45 mm.

Continuous backwash filtration systems have been reported to perform without deterioration of removal efficiency and headloss with time (England et.al. 1994; Mc Gray, 2001).

1.12 *Filter bed regeneration and backwashing*

A filter is usually in service for 8 - 24 h between traditional cleaning cycles. Automated cleaning is very common; the pressure drop across the bed is monitored until it exceeds a preset value, the filter is then taken off-line and the cleaning cycle is initiated. It is also possible to control the cleaning cycle by monitoring filtrate quality, and to initiate cleaning when solids re appear in the filtrate.

Cleaning may be accomplished by scouring the clogged portion of the bed or by reversing the flow through the bed (backwashing). Backwashing operation causes fluidisation of a bed, which can allow open space for the trapped particles to be washed away. The fluid velocity required for the onset of fluidisation is called the point of incipient fluidisation or minimum fluidisation velocity q_{lmf}. The bed voidage expands to a porosity of $V\varepsilon_{mf}$ at this point (Fee, 1994). The Ergun equation is used to correlate the pressure drop through a porous bed with the superficial velocity Q_s and it is given below in Equation 1-9.

$$\frac{\Delta h}{L} = \frac{150\,(1 - V\varepsilon)^2}{V\varepsilon^3} \frac{q_{mf}}{(\alpha_p\,\mathrm{d})^2} + \frac{1.75\,(1 - V\varepsilon)}{V\varepsilon^3} \frac{\rho_f\, Q_s^{\,2}}{d\ \alpha_p} \tag{1-9}$$

where:

Δh = pressure drop across the fluidised bed

L = bed depth of expanded bed

ρ_f = density of fluid

μ = dynamic viscosity

Vε = bed porosity

α_p = particle shape factor

Cleasby et.al. (1975) worked out the porosity of the expanded bed by considering a bed which is fluidised from initial porosity ε at bed depth L to a porosity ε_{Le} and a height L_e. Since the volume of filter grains within the bed remains constant, for a bed of constant cross section A, Equation 1-10 applies.

$$\varepsilon_{Le} = \frac{L_e}{\varepsilon + (L_e - L)} \tag{1-10}$$

For multi media beds, expansion occurs successively for each different layer. The upper layer will be expanded at lower backwash rates than the lower layers. In addition, low flowrates led to larger zones of intermixing. To achieve bed stratification at the end of a backwash cycle, backwash flowrates were increased.

Cleasby and Woods (1975) suggested optimisation of backwashing rates by pilot plant trials. Hydrodynamic shear at the water filled grain interfaces was at the maximum at a filter bed porosity of 0.68 - 0.71. Cleasby et.al. (1975) reviewed optimum backwashing rates and concluded that cleaning with backwash water alone was an inherently weak cleaning method because particle collision does not occur in a fluidised bed and abrasion is negligible. Air introduced during backwash (air scour) increased abrasion and shear through turbulence and scouring. Further scouring occurred during the contraction of the filter bed in the beginning of a backwashing cycle when air was used in combination with water (Addicks, 1992).

Backwash water consumption is highly dependent on TSS concentration and on the nature of the solids. For drinking water filters, it ranges between 1 - 2% of the total filtrate for systems using air combined with water and between 3 - 5% for air followed by water (Degrémont, 1991). For wastewater filters backwash water consumption is between 10 - 14% for pulsed backwashing (Matsumoto et.al. 1982). Wastewater filters typically require greater backwash water volumes, due to high solids concentration and high level of grease and oil.

A maximum solid loading of 2 kg/m^3 media was recommended prior to backwashing for a media of d = 1.5 - 2.2 mm (Cooper - Smith and Rundle, 1998).

Recommended flowrates for adequate fluidisation during backwashing were summarised by: Hegemann et.al. (1997) and are shown in Table 1-3.

Table 1-3. Guidelines for flowrates used for backwashing common filter media

Filter media	Grain size (mm)	Bulk density (kg/m^3)	wet density (kg/m^3)	Backwashing flowrate (m/h)
Anthracite	1.4 - 2.5	1.4	1.4	55
	2.5 - 4.0	1.4	1.4	90
Pumice	2.5 - 3.5	2.3	1.3 - 1.5	55
Silicon Sand	0.7 - 1.25	2.5	2.5	55
	1.0 - 1.6	2.5	2.5	75

1.12.1 Traditional backwashing

Water is introduced through the bottom of the filter by either mechanical pumping or by siphoning effects, generated from back-pressure at the top of the filter bed. This often fluidises the media. Introducing air during low-rate backwash creates a vigorous scouring of particles, which become dislodged and are carried away by the backwash water. Previously filtered process water is usually the source of backwash water (Rushton et.al. 1996; Dahab and Young, 1977; Chipps et.al. 1995; Miltner et.al. 1995; O'Brien, 1997; Slowey; 1990; Elsenhaus, 1982).

High water flow at the end of a backwashing procedure was recommended to purge out air bubbles (Kozar, 1994).

1.12.2 Factors affecting backwashing

Considerable turbulence in the filter bed occurs during the initial 2 min of air scour, causing most of the solids to be removed. The bubbling effect observed can create a false impression of total bed agitation where in fact only the top 250 mm may be affected. Mudballs may fall below this zone (Kawamura and Gramith, 1997).

Mudballs can be formed when rolling fluidised granules come in contact with coated grains and begin to form aggregates. The aggregates continue to grow in size with each succeeding backwash cycle. After a short period of time the aggregates may have grown to pellet size and the efficiency of the filter begins to decrease, filter cycles get shorter until the bed becomes channelled and solidified. A fixed arm surface wash with high-pressure nozzles that are able to wash to a depth of 1200 mm was suggested to overcome this problem zone (Kawamura and Gramith, 1997). Kozar (1994) investigated the use of alternative media in combination with an internal agitator system to overcome the problem of mudball formation.

Annual media losses of 5 - 10% were reported by Kawamura et.al. (1997). Position of the backwash filter trough affected media losses in cases of insufficient space between the top of the filter bed and the top of the wash trough. Cleasby et.al. (1975) reported on a 3% media loss for anthracite after 2 weeks of continuous air backwash. Modification of the backwash trough by a baffle helped to overcome this problem (Kawamura and Gramith, 1997). Further modifications, including flap gates instead of backwash troughs lead to even less media losses (Dohmann et.al. 1996). Apart from backwashing, media is also lost through the underdrain system. This can have several causes, the main one being filter nozzles which have been torn lose of their bolts, cracked joints of filter slabs and faulty support layers. Installation of the support media should be carried out by hand in order to avoid damage to the filter nozzles and each nozzle should be checked after tightening it with a torque wrench (Dohmann et.al. 1996).

1.12.3 Pulsed backwashing:

Pulsing generally loosens, redistributes and removes some of the solids retained in the surface layer. Pulsing is used periodically and in conjunction with traditional backwashing. Typically shallow filter beds with very fine medium are used for the filtration of primary treated effluent (PTE), typically < 500 mm with grain size d = 0.45 mm. In addition, the application of pulsed backwashing can provide a means to collect sludge for further processing. Backwash water from pulsed operations contains high concentrations of easily settleable solids, which can be combined with the sludge from primary settling tanks. Recycling of a major portion of the backwash water can thus be eliminated. (Howard, 1995; Irwin, 1982; Garzonetti 1984; Matsumoto et.al. 1982). Either filtered effluent or the atmospheric air, trapped in the filter's compartmentalised under drain can be used for the pulsing. Using low-pressure air diffusion, Howard (1995) created a rolling motion, and observed lifting and suspending of the particles in the water over the sand surface.

Solid particles stored in a pulsed filter bed were significantly greater than in filters using traditional backwashing methods. Pulsing increased the total filter run time between full backwashing cycles by a factor of four. Pulsed backwash was used every 3 - 6 h, in addition to chemical cleaning of the top 35 mm at weekly intervals. During a pulsed bed operation, TSS reduction of 65 - 75% and BOD reduction of 50 - 60% was achieved. Filtrate from a pulsed bed was further clarified with the use of trickling filters (Rengel - Aviles, 1986).

Irwin (1982) compared the removal of similar amounts of BOD and TSS between pulsed bed operation and activated sludge processes and estimated a saving of 90% energy costs by using pulsed bed filtration of primary effluent.

1.12.4 Two-stage backwashing

Scherfig and Mosharraf (1989) described a two-stage backwashing system. During the first stage, primary effluent was distributed between the coarse top layer and the finer lower media and used to backwash the top layer. During the second stage, final effluent was used to backwash the entire bed. The backwash water from the first stage was returned to the activated sludge system and did not contribute to the overall hydraulic load but it did reduce the amount of backwash water required for the secondary stage.

1.13 Porous ceramic dual media (PCDM) filtration

Porous ceramic dual media (PCDM) filtration technology is a form of dual media filtration developed by Silicon Industries Ltd (Hamilton, New Zealand) in collaboration with the School of Science and Technology at the University of Waikato (New Zealand). Instead of conventional plastic nozzles attached to a false under floor plenum, the PCDM system employs plastic bolts holding down porous ceramic (PC) tiles or purpose designed PC nozzles (Langdon et. al. 1996). The dual media system is created by using modified pumice, Silicon Sponge™ (SS) (d = 1.2 - 2.2 mm, ρ_g = 1.25 g/cm^3), as an upper layer and fine sand (d = 0.3 - 0.6 mm, ρ_g= 2.15 g/cm^3), as the lower layer added directly over the PC tiles or nozzles (Liu, 1997). Because of the high solid holding capacity of SS and the absence of support graded gravel, the PCDM system offers the advantages of increased flow rate and longer run times (Liu, 1997; Langdon et.al. 1994 and 1995). The increased efficiency of this system may a drinking water technology applicable to effluent treatment.

1.14 *Porous ceramic (PC) nozzles*

PC nozzles eliminate the need for a gravel layer to support the sand. In addition, when used in conjunction with SS, the system can achieve double the filtration rate and a four-fold increase in filtered volumes before backwashing becomes necessary (Langdon et. al. 1996). During refurbishment of the existing filter tanks, traditional filter nozzles are replaced with PC nozzles. Details on PC nozzles are shown in Chapter 5.

Mixing crushed, graded window glass and TM with an alkali metal silicate solution (Silicon Supplies, 1991) produces the porous TM ceramic (TMC). This mixture is compressed into the shape of a filter nozzle. Brief exposure to carbon dioxide initiates reaction with the alkali silicate. This provides strength to the nozzle, which is further, strengthened upon firing at a temperature sufficient to decompose the carbonate phases and to sinter the glass and mineral constituents. The porosity of TMC nozzles ranged between 0.4 - 0.45 (McCabe, 1991).

1.14.1 PCDM media

Two layers of media are used in PCDM filtration. The bottom layer consists of silica sand and the top layer of SS (14/24), d = 0.6 – 1.1 mm or SS (7/14), d = 1.5 – 2.2 mm (Yong 1997). To produce SS river pumice is graded and thermally modified by Works Filter Systems Ltd., Hamilton (WFS). Compared with the more commonly used anthracite, SS possesses greater resistance to abrasion and therefore allows for more vigorous backwashing procedures (Hill and Langdon, 1993). Shorter backwashing times and longer filter run times are therefore expected (Hill and Langdon, 1993; Liu, 1997). The lower density of SS allows fluidisation and settling in an upper layer in a dual media system. SS is less expensive than anthracite.

SS is negatively charged as a result of the large proportion of free silica sites at the grain surface which lead to a reaction with water to form hydroxyl groups (Hill and Langdon, 1993), see also Table 2.5. Positively charged flocs are therefore efficiently removed.

1.14.2 Applications of the PCDM system

The PCDM system is currently installed at several drinking water treatment plants throughout New Zealand. These filters have performed well, particularly with respect to flow characteristics, turbidity of the filtered water and residual aluminium (Liu, 1997; Hill and Langdon, 1993). The use of PCDM has also been successful in the beverage industry. The large solid holding capacity may make the PCDM system appropriate as a wastewater filtration technology and indeed PCDM filters have recently been installed in two small NZ communities to treat flocculated secondary treated effluent.

1.15 Objectives of the present investigation

The overall objective of the work is to obtain the data that will allow assessment of the possibility that filtration, particularly primary filtration without the use of chemicals, offers advantages as an effluent treatment strategy. An important part of this work will be to trial the PCDM system (with SS and TM as media) in the filtration of wastewater.

The superior performance of the PCDM system in water filtration, in terms of long filter runtimes and ease of backwashing, makes it the obvious filtration technology for these studies. The PCDM system uses dual media containing coarse modified SS upper layer and a lower layer of finer sand. It also incorporates nozzles made from a porous ceramic material. Characterisation of these materials will allow their optimum use. TM, an indigenous ironsand material, has properties that may make it useful as a filter medium. Characterisation of this material is also required.

Specific objectives to achieve the overall aim of the thesis are outlined below:

- **Media studies:** A necessary starting point for the research is the characterisation of the SS and TM materials to be used in the work. Determination of physical and chemical parameters as well as hydrodynamic properties such as bed porosity packing densities hydraulic conductivities and bed expansion and fluidisation effects during backwashing is needed. In addition, the attraction between the media surface and solid components from the effluent was also of interest, in terms of the potential of biofilm formation.

- **Bench trials of effluent filtration:** In order to gain an indication of how key parameters (run time, bed depth, grain size, flow rate, bed configuration, backwash regime) affect media performance during effluent filtration, a program of bench trials is required. It is proposed that a model effluent derived from reconstituted cow manure or primary treated municipal effluent in small glass columns of SS and TM be used in this work. A issue of increasing importance in water treatment as well as wastewater treatment, is the removal of micro-organisms such as giardia and cryptosporidium. Typically the cyst size of these organisms is in the vicinity of 5 •m. It is of interest to determine whether fine TM can remove particles of this size.

- **Backwash studies:** It was anticipated that filters using fine TM media would block after short run times. The usefulness of TM would therefore depend upon improved backwashing procedures. The majority of removed solids are likely to be deposited as a surface film. It may be possible to remove that film by a short backwash pulse. The conditions for fluidisation of SS and TM during conventional backwashing are also required.

- **Nozzle and bolt design and flow characteristics:** The ceramic nozzles used in the PCDM system have the major advantage of allowing the media, even fine TM, to be placed directly on top of the nozzle without the use of supporting layers of graded gravel. The ceramic material can be fabricated into a variety of shapes allowing novel nozzle designs. The flow characteristics of the PC nozzle-bolt combination need to be investigated. Of particular importance are configurations that allow maximum flow during backwashing.

- **Pilot plant trials:** Bench scale trials with small diameter tubes, while providing indicative trends, may suffer the disadvantages of wall effects and of necessity do not represent all the variations experienced in wastewater streams. Pilot plant trials are a necessary part of developing commercial treatment systems. In the current work pilot plant trials will be used to test the performance of TM and SS media under realistic filtration conditions, determine solids removal from both, primary and secondary effluent, allow optimisation of bed depth and investigate pulse backwashing.

Results obtained from these studies should allow the design and optimisation of treatment systems incorporating filtration as a technology for reducing suspended solids. At the primary stage such filtration will reduce suspended BOD, reducing the load to secondary treatment and may allow economic advantages through energy recovery and reduced CO_2 emissions.

The experimental work is divided into chapters 2 to 6. Chapter 2 describes methods and material that are used throughout the thesis. The construction and use of equipment designed for specific investigations are described in the chapters in which the results are reported. Filter media, effluents and the methods used to determine principal physical and chemical characteristics are introduced in chapter 2. The same chapter also describes the analytical methodology used during filtration experiments. Chapter 3, covers the characteristics of media and effluents used. In chapter 4, bench studies, involving the use of model effluents and bench-scale filter columns are described. Backwashing studies, using TM as a pulsed are also included in chapter 4. Specialised rigs and evaluation units used in pilot plant studies are described where appropriate in Chapters 5 and 6. Chapter 5 investigates flows through the filter nozzle-bolt combination. The knowledge gained from bench-scale filtration experiments was translated into a deep bed pilot plant study. This is described in chapter 6. Discussion of findings is generally included within the chapter the results are presented. Chapter 7 gives an overview of the work and investigates commercial aspects and examines potential energy-benefits.

Chapter 2

Methods used for characterising media and monitoring effluents

2.1 *Introduction*

The use of indigenous materials without chemicals as filtration medium required investigative work to characterise both, titanomagnetite (TM), Silicon Sponge (SS) and the PCDM nozzles. This chapter lists the methods used to measure the physical and chemical properties of these materials. Media characterisation in terms of its filtration efficiency includes the key factors such as filter grainsize and porosity. Methods used to monitor the performance during filtration experiments as well as health and safety issues are also part of this chapter.

SS and TM were the main filtration media used in this work. Other media types included silica sand and supporting gravel. The nozzles used were based upon a patented TM ceramic (Silicon Industries, 1991). Effluent was derived from three principal sources: farm dairy effluent, primary treated effluent and secondary treated effluent. The following sections will be principally concerned with the sources and methods of characterisation.

2.2 *Titanomagnetite*

TM concentrate was obtained from Works Filter Systems (WFS) in Hamilton, New Zealand. It was mined by BHP New Zealand Steel Limited and supplied as iron sand concentrate. A cleaning procedure at WFS used bed expansion and water flotation to remove the lighter and non-ferrous compounds. It was stored in 50 L bags and all the work was carried out with sub-samples from the same bag.

The sampled material contained residual clays and minor contamination by soil, wood and paper from storage bags. A range of different cleaning procedures was tested, including mixing with a sonic probe, rinsing in acid and caustic and the use of a mechanical shaker. Details are listed in Appendix A 1. The final method, adopted for

TM consisted of sieving, air-scouring in a solution of NaOH, followed by backwashing it in water. Grain size was limited to a range between > 75 µm - >180 µm. Details of this procedure are set out below.

Following the dry sieving, chemical washing of <600 g media was carried out inside a glass column of 50 mm ID and 500 mm height. A solution of NaOH (0.1 moL) was added to the medium inside the column to cover the material prior to air scour. Mixing was achieved with air, introduced from the base at a rate of about 40 m/h during 5 min and interrupted every 1.5 min by short backwash water pulses in order to break up the media plug. After 5 min of this treatment, 5 min of final backwashing with water at 100% bed expansion served to flush out all remaining loose material from the media. Turbidity of the final backwash water was <1 NTU.

Between experiments the medium was regenerated by backwashing, using the following steps: 2 min water at 150% bed expansion, 10 sec air-scouring and water again for 30 sec at 150% bed expansion.

2.3 *Silicon sponge*

SS is a modified pumice, marketed by WFS. It is available in two different grades: a coarse grade (SS 7/14) and a finer grade (SS 14/25). WFS obtained the unmodified raw pumice aggregate from Winstones Quarry in Puni, where it was mined from riverbeds. The hydraulic action of river transport tends to remove the softer material commonly found in pumice from land deposits, see also Figure 1-4. This distinguishes the material from other pumices that have been used in filtration. A proprietary procedure at WFS used thermal processing and size grading (Hill and Langdon, 1993). It was stored for use in large stockpiles. A representative sample for bench studies was obtained by shovelling about 50 L of the medium from various positions in the stock-pile.

After a 24 h period of tap water soaking the SS was vigorously backwashed to remove about 20% of pumice fines. About 10% of the pumice grains remain floating after the soaking period, this material was discarded. The remaining SS was backwashed with water at 100% bed expansion for about 10 min. Backwash water turbidity at this stage was < 1 NTU.

2.4 Silica Sand

WFS obtained silicon sand from a Winstones Quarry, in Puni, NZ Coal quarry and stored it in stock-piles. A representative sample was collected in a similar manner as described above for SS. No prior soaking was required for this medium. Backwashing procedures as outlined for SS were followed.

2.5 Research plan

The actual study began in April 1996. Preliminary pilot plant studies, using primary settled domestic effluent (PTE), were carried out between February and April 1995 by an employee of WFS. At that time the data was recorded only, the evaluation was still necessary. This formed a substantial part in the results section of the study, see Apendix A 9. Bench-scale experiments followed to seek explanations for suspected trends seen during the preliminary trials and to expand and introduce alternative variables. Deep bed trials followed as a conclusion to the bench-scale experiments. The final trial on secondary treated effluent resulted from a commercial opportunity. As with the preliminary pilot plant studies, data was collected by an employee of WFS during January 1999 and evaluated later as part of this study.

2.6 Methods for characterising filtration media

A variety of physical and chemical methods used to characterise the media and in some cases the TMC, are described below.

2.6.1 Hydraulic conductivity

In most situations, Darcy's law (Equation 1-2) can be used to estimate how much water flows through porous media. Hydraulic conductivity values for the media used during this study were determined.

Method:

A glass column of 35 mm diameter and 500 mm height was filled with 45 mm depth of filter media to a depth of 45 mm. In order to test the TMC nozzle material, a ceramic disc was fitted and sealed inside the glass column. Headloss was measured with manometer tubes at the top and bottom of the filter bed or immediately before and after the ceramic disc insert. Constant head was maintained during the experiment with an open glass tube inserted through a rubber bung from the top.

Out-flow of water was controlled in order to stabilise the difference in headloss before and after the filter bed.

2.6.2 Grain shape

Grain shape was determined by light microscope and by scanning electron microscope investigations. The light microscope was an Olympus BX60 and investigations were carried out at 20 – 50 x magnification. The scanning electron microscope was a Hitachi S 410 and sample preparation was required, prior to analysis. Sample preparation for dry and clean media grains involved slow drying in the oven and coating with platinum/palladium following photography. Sample preparation for samples requiring an assessment of the biofilm involved fixation in 2.5% w/v glutaldehyde, buffered to pH 7.2 with 0.1 moL sodium cacodylate for 24 h. Post fixation was in 1% w/v osmium tetroxide in the same buffer for 2 h. The fixed samples were rinsed twice with distilled water and dehydrated in an ethanol series. After drying by the critical point method, samples were coated and photographed as described above (Harris, 2001).

2.6.3 Surface area, void space and pore radius

Surface areas, void space and pore radius was determined by using low temperature gas adsorption analysis (Nova 1000 High Speed from Quantachrome Corporation). About 1 - 3 g of sample was placed inside the container provided with the analyser.

Principle

The sample is degassed by prolonged heating and by evacuation. It is then cooled to the temperature of liquid nitrogen by submergence in a liquid nitrogen bath. Finally it is exposed to gaseous nitrogen at increasing pressures. Some of the nitrogen gas molecules become physically adsorbed on the specimen surface. Exposure to increasing pressures results in increased adsorption until a more or less permanent layer of nitrogen a few molecules deep is acquired on all exposed areas, including the walls of the pores. At some point, depending on pore size, further increase in nitrogen gas pressure leads to the initiation of nitrogen condensation within the smallest pores. The condensation process extends to larger and larger pores with further pressure increases until all pores are filled with condensed liquid. The reverse process takes place upon pressure reduction. The liquid within the larger pores evaporates first. Finally the adsorbed molecules are removed. The amount of nitrogen gas adsorbed or condensed on a specimen can be determined by careful monitoring of gas pressure (P). The entire process is carried out under pressures ranging up to the saturation vapour pressure (P_o) of the nitrogen at the temperature of the experiment. Experimental data are plotted as curves of V_{STP} versus P/P_o where V_{STP} is the volume adsorbed at standard - temperature - pressure (STP). Alternative plots use the gas volume given as a percentage of the total gas uptake per unit weight of specimen. From this plot, surface area and pore size of the media specimen are calculated (Orr, 1980).

2.6.4 Bed void volume, wet media density, envelope and absolute density

Definitions and equations

Bed void volume ($V\varepsilon$) or porosity represents the amount of water, which continuously occupies the voids between the grains of a filter bed.

$$V\varepsilon = V_B - V_G \tag{2-1}$$

Wet media density (ρ_{wM}) represents the weight per unit volume of a filter bed with water in the void spaces and in the pores.

$$\rho_{wM} = \frac{(V_G + V\varepsilon + V_P)\, kg}{mL} \tag{2-2}$$

Envelope density (ρ_E) represents the weight per unit volume of the medium with pores in the grains filled with water but with air spaces between the grains.

$$\rho_E = \frac{(V_G + V_P - V\varepsilon)\, kg}{mL} \tag{2-3}$$

Absolute density (ρ_A) is obtained when the volume measured excludes the pores and void spaces between the particles.

$$\rho_A = \frac{(V_G - V_P - V\varepsilon)\, kg}{mL} \tag{2-4}$$

where:

V_B = Volume of filter bed

V_G = Volume of media grain

V_P = Volume of media surface pores

V_ε = Volume of void between grains or bed porosity

Methods for measuring void volume

About 100 mL of media was boiled in excess water for 30 min while stirring to remove the air bubbles. Excess water was decanted until it was level with the top layer of the medium. The combination of water and medium was weighed before and after drying at 110° C for 48 h. The weight and volume of the media (excluding the volume of pores) gives the absolute density ρ_A. Wet media density ρ_{wM} was determined by soaking about 200 mL media inside a 150 mm diameter glass dish for 48 h, draining all excess water to the level with the media surface, before the weight was taken. The amount of water evaporated from the pores of the wet media gives the media pore volume. The weight of water in the void filled system minus the weight of water in the pores gives void volume V_ε or porosity of the filter bed.

For the absolute density ρ_A a standard liquid pycnometer method (BSS 733:1952) adapted by University of Waikato, Dept. of Earth Sciences (1999) was used. Media characteristics determined by these measurements are summarised in Table 3-4.

2.6.5 Electrokinetic properties

Iso electric points (IEP) were measured using a: Mutek PCD 03 instrument (Mutek Analytic Inc. Germany). The instrument measures the potential caused by the velocity difference between the grain particles and the fluid.

Principle

The necessary shear flow is induced by periodic movement of a piston in a cylinder filled with a fluid and a suspension of the grains to be investigated. Acceleration and deceleration of the grain particles causes a measurable electrical potential in axial direction. Data of the observed potential (in millivolt (mV)) and the pH value of the suspension fluid allow the IEP of the particles to be determined (Yang et.al. 2000). Viscosity factors were ignored during these measurements, and water was used at constant room temperature of about 22 ^{0}C. SS and TM media samples were rinsed in distilled water and dried, prior to the measurement.

2.6.6 Hardness

SS was analysed by the School of Earth Science, Victoria University, Wellington, New Zealand. The hardness grade on the moh hardness scale ranged between 6 – 7. The large variation of untypical grains in SS, made it difficult to assess the exact hardness value, which was dependant on fragments from other minerals present on the grains (Jamieson, 2002). In another experiment by Langdon et. al. (1995) abrasion of anthracite was compared with SS by backwashing both media continuously for 22 h. Loss of media was measured after that time. Weight loss of anthracite was 13% and for SS it was 4%.

2.6.7 Grain size and particle measurement

Two different methods were used to determine the sizes. For coarse and medium sized media, dry sieving was used. For finer material, such as suspended solids from liquid effluent, laser diffraction analysis (Malvern particle sizer) was employed.

Dry sieving

Dry sieving was carried out on an automated shaker (Fritsch – shaker), using a method described by Barrett and Brooker (1989). Wire mesh sieves, graded to British Standard (BSS 410: 1962) were stacked on top of the shaker. Shaking time was made up of: 6 min on intermittent mode, 6 min on micro mode and again 6 min on intermittent mode. Sample volume was chosen, so that the amount did not exceed a depth of 20 mm on the top sieve. For TM, this allowed for batches between 200 g and 300 g and for SS, batches between 40 g to 60 g. Maximum volumes were determined previously to ensure consistency in results was maintained. Similar batch sizes for sieving was recommended by Cleasby and Woods (1975).

Laser diffraction analysis

Suspended particles in untreated and in flocculated effluent as well as some fine filter grains, were measured by using laser diffraction analysis. A flow cell (MS 17) was used for sample circulation and pre – mixing. Malvern Instruments supplied the instrument, Mastersizer 1000. Instrumental settings, together with a copy of a calibration sheet are attached in the Appendix A 2. For the purpose of determining particle removal efficiency, the surface area (d^3) and volume (d^4) of the particles was used to qualify particle size. Results were therefore expressed as the equivalent volume mean diameter D [4,3]. Equation 2-5 below illustrates an example of this calculation, based on three spheres of diameters 1, 2 and 3 units.

$$D[4,3] = \frac{1+2^4+3^4}{1^3+2^3+3^3} = 2.72 = \frac{\sum d^4}{\sum d^3} \tag{2-5}$$

Data was either saved on an Excel spreadsheet or copied manually on a Cricket graph spreadsheet prior to evaluation.

Chemical composition

For a detailed elemental analysis, media samples were milled to < 20 μm grainsize, using a ring mill (Rocklabs, NZ) at 7000 rpm for 30 sec. Samples were sent to the Geoscience laboratory, Sudburry, Ontario for analysis by wet digest and x – ray fluorescence (XRF). Iron oxides, present in TM sands as FeO, Fe2O3 and Fe3O4 (Briggs, 1997), were measured by titration (International Standard: ISO 9507: 1990 E), following BHP laboratory procedures (TM - 5000.070). This method detects Fe as m/m %, and by using a factor of 1.43, it can be converted into Fe_2O_3.

2.7 *Health and safety issues*

Sewer maintenance and wastewater treatment workers risk exposure to a variety of disease causing micro -organisms (Gerba and Enriquez, 1994). In order to minimise the risk during the testing required as part of this study, all laboratory experiments were carried out inside a designated area. All glassware was cleaned inside an autoclave at 130^0 C for 15 minutes. Where backwash water supply was connected to town water supply, a vacuum breaker should be installed to prevent cross contamination. Protective clothing and gloves were worn at all times, including a face mask during the pilot plant trials. Used media was sealed in a plastic bag before discarding it with other laboratory solid wastes. Used effluent was disposed through the sewerage system.

2.8 *Effluent systems*

The effluent studied included primary and secondary treated municipal effluent and farm dairy effluent. Primary treated effluent (PTE) was obtained from the Hamilton Wastewater Treatment Station (HWWTS). PTE was used for studies using small-scale model filters in laboratory trials. It was also the effluent of choice for most of the field trials using PCDM evaluation units. Secondary treated effluent (STE) was used during field trials involving a modified PCDM evaluation unit. It was obtained from the Morrinsville Wastewater Treatment Station (MWWTS). Farm dairy effluent (FDE) was collected as solid bovine manure from a dairy farm paddock in the Waikato. It was reconstituted in the laboratory to obtain 1% TSS, which is similar to the concentration commonly found in dairy shed waste after milking (Longhurst et.al. 2000; Nadarajah 1997). FDE was used for small-scale laboratory experiments. Attempts to prepare a synthetic material, which would represent the size range of suspended particles found in PTE and FDE were abandoned when it was found, that this material had a large proportion of particles with sizes > 10 μm, see Section 3.11.2.

2.8.1 Composition of municipal effluent at Hamilton Water treatment Station

The main contribution to HWWTS comes from domestic users, with industrial users making lesser contributions. The composition of the effluent at HWWTS is as follows:

- *35% Greywater*: waste from baths, showers, laundries and dish-washers.
- *25% Sanitary*: waste from toilets and urinals.
- *20% Commercial/Industrial*: the waste from this source includes meat works, hospitals and other industrial sources.
- *20% Infiltration*: there is continuing inflow; this occurs when pipelines are laid below the level of the water table. Inflow increases during heavy rain through undetected entry points, which during times of heavy rainfall increases sewage flows.

Daily loading rates vary, caused mainly by a higher proportion of the infiltration component during wet weather.

Hamilton municipal wastewater contaminants and volumes of flow

The main contaminants of Hamilton's municipal wastewater (HMWW) are summarised in Table 2-1. During a normal day, peak loading occurs in the early afternoon, between the hours of 1 - 3 PM.

Table 2-1. The principal contaminants and levels of Hamilton municipal wastewater.

Contaminant	Level (medium)
TSS (mg/L) inflow	360
TSS (mg/L) outflow	85
BOD (mg/L) inflow	260
BOD (mg/L) outflow	149

Volumes of flow vary; a monthly medium flow of 4.5×10^4 m^3/day includes extreme maximum flows of 1.5×10^5 m^3/day (Mc Graw, 1995).

Current regulations and resource consent conditions for the HWWTS

The permitted median TSS level in the outflow is 5000 kg/day. Under conditions of high flows, TSS concentration in the outflow should therefore not exceed 33 mg/L in the treated effluent. Under medium flows, 110 mg/L would still be permitted. These relatively high permitted discharge values are taking into account the already high TSS level in the Waikato River. At the point of discharge, a medium average flow of 1.3×10^7 m^3/day (156 m^3/sec) with TSS levels between 84 mg/L to 120 mg/L was been reported (Charles, 2001).

The high TSS levels in the Waikato river is caused by the rapid rate of tectonic up-lift which in turn add large volumes of sediment to the fluvial systems (Cooke, 2002). New Zealand is one of the globe's prominent sediment sources, accounting for about 2% of the annual suspended fluvial input to the oceans and about 9% of the annual input to the South West Pacific (Carter and Mc Cave, 1997).

2.8.2 Size of suspended particles in PTE

The particle size distribution of PTE was determined by laser diffraction analysis (see Section 2.5.7).

2.8.3 Composition of secondary treated effluent at MWWTS

Contributions to the MWWTS vary according to fluctuations in contributions from industries. These include: the Morrinsville stock saleyards, a hydrogen peroxide plant, a dairy processing plant and a clothing manufacturer.

Morrinsville municipal wastewater composition and volumes of flow

The main components of Morrinsville's municipal wastewater are summarised in Table 2-2 from data obtained by Opus International (1999).

Table 2 - 2. The principal contaminants and levels of Morrinsville municipal wastewater.

Contaminant	Levels (medium)
TSS (mg/L) inflow	25
TSS (mg/L) outflow	12
BOD (mg/L) inflow	20
BOD (mg/L) outflow	12

Volumes of flow vary from 4.4×10^3 to 1.1×10^4 m^3/day.

Current regulations and resource consent conditions for MWWTS.

In relation to the turbidity, the current resource consent condition, (Environment Waikato, 1995) specifies that "the turbidity in the receiving water shall not increase by > 20% at a point 50 m below the outflow into the Piako River, compared with a point at 20 m above the outflow". Existing turbidity levels in the Piako River vary between 20 - 80 NTU (Mc Graw, 2001).

2.8.4 Size of suspended particles in STE

Due to the relatively low concentration of TSS in STE, detection of the size, using laser diffraction analysis was not possible. Particle size distribution between 1 to 80 μm was reported by Boller (1980), using the Coulter counter method. This method was not available during this study. A similar size can be assumed for STE from MWWTS.

2.8.5 Composition of Farm Dairy Effluent

Data gathered on FDE in New Zealand found an average dry matter content of 0.9%. The amount of dry matter in FDE was related to nitrogen levels (Barton, 1994) with

Nitrogen levels up to 400 mg/L recorded. The dry matter content in FDE is made up of about 10% solids from excreta, 4% from teat washings, 86% from wash water plus other foreign matter (Gibson, 1995). Average values of Phosphorus and Potassium were 70 mg/L and 370 mg/L, respectively (Longhurst et.al. 2000). Goold (1980) reported Calcium, Magnesium and Sodium levels of 177 mg/L, 39 mg/L and 54 mg/L respectively. The reconstituted FDE used can be expected to have a similar composition.

2.8.6 Size of suspended particles in FDE

Following reconstitution of collected farm manure to a 1% suspension, particle size distribution of FDE was determined by laser diffraction analysis (see Section 2.5.7).

2.9 *Collection of effluent*

PTE was collected from (HWWTS). Collection point was after the primary settling tanks, immediately before the outlet drain and before chlorination. For laboratory trials, effluent was pumped into a 50 L drum which was used for return transport to the laboratory. Typical turbidity levels ranged between 70 to 100 NTU. Trials were carried out on the collection day and the following day.

FDE was collected on demand. Fresh material was gathered in a 10 L bucket from a farm paddock in Waihou. A model effluent, using fresh bovine faeces was prepared by reconstitution in water to a dry matter concentration of 1%.

Preparation of FDE

Dry matter of each fresh sample batch was determined, by drying 30 mL - 50 mL of fresh manure to a constant weight in a microwave oven. Dry matter values of the fresh manure ranged between 20 - 40%. Sub samples were stored at -10^{0}C in the laboratory freezer (Frigidaire by Fisher & Paykel). A model FDE with 1% dry matter concentration was prepared from the de-frosted samples. This suspension was allowed to settle overnight and the supernatant used for the experiments. The settling procedure removed about 98% of the turbidity, resulting in turbidity levels between 1200 - 2000 NTU and a suspension with about 0.1% solid content.

2.10 *Synthetic effluent*

An attempt was made to produce a synthetic effluent to represent PTE and FDE. The range of components used included: yeast, lactose, whey, meat extract, bentonite, kaolinite, blood and bone, $CaCO_3$ with Na-azide as stabiliser and Na-hexametaphosphate as anti coagulant.

Of the components listed above, it was found that the size of most particles in casein and in blood and bone was above the size representing PTE or FDE. This resulted in a final selection of the following components per 10 L water:

- 10 g meat extract "Bovril" (produced by Bestfoods, UK, distributed by Recket & Colman, Henderson, NZ),
- 2 g $CaCO_3$ (BDH) AnalaR
- 0.5 g Bentonite (Dept. of Earth Science, University of Waikato, Hamilton).
- 10 g Na-azide (BDH) AnalaR
- 50 g Na-hexametaphosphate (BDH) AnalaR.

2.11 *Measurement of filter performance*

Effluents are complex systems typically containing dissolved and suspended material contributing to total BOD, colour, turbidity, nutrient load and microbial population. Filtration has the capability of removing the suspended material. The preferred means of monitoring filtration efficiency is by determining total suspended solids (TSS), expressed as mg/L. Often TSS can be correlated with turbidity. Turbidity provides a much more convenient means of monitoring filter performance.

2.11.1 Turbidity

Turbidity was measured using a Hach Model 2100P instrument. Calibration procedures were followed according to the instructions supplied by the manufacturer. A stock solution of a 4000 NTU calibration standard was prepared as follows. Solutions A (1.000g of Hydrazine Sulfate $(NH_2)_2$ - H_2SO_4 dissolved into 100 mL distilled water) and solution B (10.00 g of Hexamethylenetetramine $(CH_2)_6N_4$ dissolved into 100 mL of distilled water) were combined to produce solution C and allowed to stand for 48 h at 25 ± 3^0 C. Dilutions were carried out to obtain calibration standards of 10, 50 and 100 NTU. The instrument was calibrated daily, using the three standards. Blank checks using distilled water were carried out before and after a run. Data were accepted if blank values were < 0.5 NTU. Sample cells were washed with Decon™ and rinsed with distilled water between runs.

Measurement of turbidity in liquid samples

Liquid effluent samples were collected directly before and directly after the filtration plant or the laboratory filter column. During laboratory trials, collection of the sample after the filter column was a composite of up to 30 min run time, depending on the flowrates. Analysis was usually within 1 min of the sample collection.

Measurement of turbidity released by solid samples

Wash water turbidity analysis was determined by sampling media from the centre of the filter bed with a soil core sampler (ID 50 mm, length 1 m). About 50 mL of this media was then mixed in 100 mL distilled water, which included a surfactant, (5% Decon™). A horizontal shaker was used (Miltern) at 300 rpm for 5 min. Following overnight storage, the wash water was separated from the media using a metal sieve (150 μm). A further 50 mL of wash water was used to rinse remaining solids from the media. Turbidity of the combined wash solutions was determined using appropriate dilution where necessary.

2.11.2 Total suspended solids

Effluent samples for TSS were collected before and after filtration and from the wash water obtained from the filter medium. The method used followed the standard method (2540 D) described by Clesceri et.al. (1998). Some of the samples were analysed at Waikato University, others under contract by Hill Laboratories, Hamilton. Agreement between the two methods was established by an inter-laboratory comparison listed in the Appendix A 3.

Method

The material included Millipore glass micro fibre filter (Catalogue number 1822), pore size was 0.22 μm and 0.45 μm, a Nalgene Filter Apparatus, with 250 mL capacity and a drying oven (Cutherm Digital Series).

Samples of fresh and filtered effluent were collected into a plastic container with 250 mL holding capacity. Sample size for the analysis was 100 mL of filtered effluent or between 40 mL - 100 mL of unfiltered effluent wash water from the filter media. Wash water from the filter media was obtained as outlined in Section 2.9.1. The clean filter media was then dried at 105°C over night to allow final calculation, relating the captured TSS to the amount of dry filter media.

2.11.3 Turbidity versus suspended solids.

Traditional solid analyses, usually completed by gravimetric methods, are time-consuming and technique sensitive. Generally, it takes from two to four hours to complete such an analysis. Faster evaluation, such as turbidity measurements was desired in the course of the experiments for this study, as a convenient means for monitoring trends. The reliability of the data obtained relies on a reasonably predictable correlation between turbidity and TSS of the samples. This correlation was tested for three different effluent types: farm dairy effluent (FDE), primary treated municipal effluent (PTE) and secondary treated municipal effluent (STE). A fourth sample was the wash water from used media. Figure 2-1 shows the correlation between turbidity and TSS for the effluent samples used in this study. A least square relationship was established and the curve fitted by the power method. A correlation of R^2 = 0.9 or greater indicates a workable relationship between turbidity and TSS (Sadar, 1998).

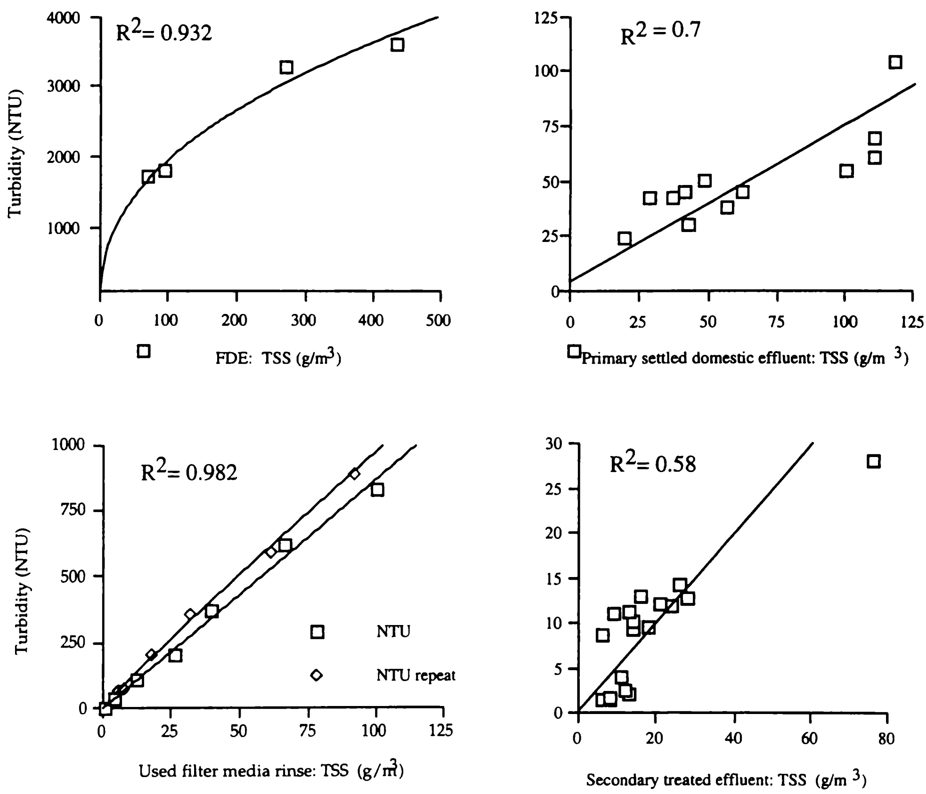


Figure 2-1. Relationship between effluent turbidity and TSS for primary and secondary treated municipal effluent and FDE.

The poor correlation between turbidity and TSS, shown in Figure 2-1, was realised, especially in the lower turbidity range, applicable to secondary treated effluent. However for the majority of the work, turbidity provided an adequate method for monitoring TSS removal. Where it was critical, during pilot plant studies using secondary treated effluent, samples were also tested for TSS and BOD.

2.11.4 Jar testing

Principle

A jar test simulates mixing and settling conditions found in a clarification plant. It was used to monitor the performance of coagulation chemicals. The jar tester (Boltac, from Boltac Industries Ltd, Hamilton, NZ) used, allowed for up to six individual tests to be run simultaneously with a variable speed motor to control the mixing energy in individual jars.

Two types of coagulants were used. The first coagulant was Poly Iron Sulfate (PFS); $Fe_2(SO_4)_3$ with a Fe^{3+} concentration of 12% min., marketed as PFS Solution and supplied via WFS by Aluminates (Morvell) Pty Ltd., Melbourne, Australia. Local costs: NZ$ 1400/m^3 (Jamieson, 2000). The second coagulant was Poly Aluminium Chloride (PAC); $[Al_2(OH)_5Cl_{2.6}(SO_4)_{0.2}]_n$, with a Al_2O_3 concentration of 10% min., manufactured and marketed as Liquipac from Fernz Chemicals Ltd., Morrinsville, New Zealand. Local costs: NZ$ 630/m^3 (Kennel, 2000). During this study experiments were carried out with the more expensive PFS, for commercial reasons.

Jar testing method

Effluent samples were distributed between each jar, resulting in 1 L / jar. For the PFS coagulation-dilution series, the coagulant was made up from a 4% stock solution and added at a high mixing speed. The final concentration of PFS ranged between 20 - 300 mg/L. For the PAC coagulation-dilution series, the coagulant was diluted directly from the container to obtain a concentration range of 200 to 300 mg/L PAC. The duration of high speed mixing was <1 min, this was followed by slower mixing for a further 5 min. In the slow-mix period, floc building advanced. The actual mixing time was refined as the test regime proceeded. A settling time of 5 - 10 min followed. Floc settling rate and turbidity of the settled liquid were recorded.

2.11.5 Refractive index

Effluent was measured before and after filtration. Instrument used was: Refractometer (Brix ref. 113), supplied by Bell Technology, Epsom, Auckland. When altering colloid concentrations in suspensions, the refractory index can provide an indication of colloid composition, which may affect the light scattering during turbidity analysis (Sadar, 1998).

Based on two FDE samples and two PTE samples, it was found that filtration did not affect the refractory index.

2.11.6 Biological oxygen demand (BOD_5) and carbonaceous biological oxygen demand $CBOD_5$.

Both these values were determined, by comparing dissolved oxygen levels before and after 5 days storage at 20° C. The method used, followed the standard method (5210B) described by Clesceri et.al. (1998). Analysis was carried out within 2 h of collection.

Chapter 3

Characteristics of media and effluents used in the studies

Methods described in Chapter 2 were used to characterise the various materials and effluents used in the studies. Most of the work on TM, SS and TMC nozzles is new. Data for silicon sand is included for comparison.

Hydraulic conductivity was measured to allow better planning of the pilot plant experiments. Together with the results on grain size, void space and pore radius, it was needed to estimate expected flowrates through the filterbed. The grain shape and the surface area of each media type influence its role as a substrate for biofilm formation and particle attachment. Wet media density and absolute density determine the media's behaviour during backwashing. Elektrokinetic properties can provide an indication of preference in the attachment of anionic or cationic particles. Filter media is often graded to uniform size and density in order to prevent unnecessary media loss and to prevent intermixing of the different layers of a dual or multi media filter bed. Media heterogenity and uniformity coefficient was a necessary starting point for this information. Chemical composition was of interest to compare the silicon sponge and the TM media with other sources of pumice and ironsands.

3.1 *Hydraulic conductivity*

This study relates the hydraulic conductivity to the grain size of the various sizes of TM, and the other filter media of this study.

Data for hydraulic conductivities of the various media used in the study are summarised in Table 3-1.

Table 3-1. Hydraulic conductivity of filter media.

Media type and size	Hydraulic conductivity (10^{-3} m/s)
TM, 125 - 150 μm #	0.30
TM, 63 - 106 μm	0.18
TM, 106 - 125 μm	0.09
TM, 150 - 180 μm	0.30
TM, 75 - 180 μm	0.27
TM, 180 - 212 μm	0.38
TM, 30 - 220 μm #	0.20
TM, 30 - 220 μm	0.25
SS 14/25 #	13.80
SS 14/25	13.10
SS 7/14	25.4
Silica sand 30/60 #	2.1
Silica sand 30/60	2
TMC nozzles	1.5

Yang, 2001.

Values of hydraulic conductivity are generally inversely related to particle size of the filter media, as is expected from application of the Carmen Kozeny Equation (Equation 1-4). In the case of TM, removal of the fines resulted in a small increase of hydraulic conductivity.

3.2 *Grain shape*

3.2.1 Microscopic studies of clean filter media

Surface characteristics of media particles were investigated at high magnification using SEM (Hitachi S 410). Sample preparation for dry and clean media grains, comprised slow oven drying and coating with platinum/palladium.

Data from light microscopic investigations are summarised in Table 3-2. Images from SEM studies are presented in Figures 3-1 to 3-3.

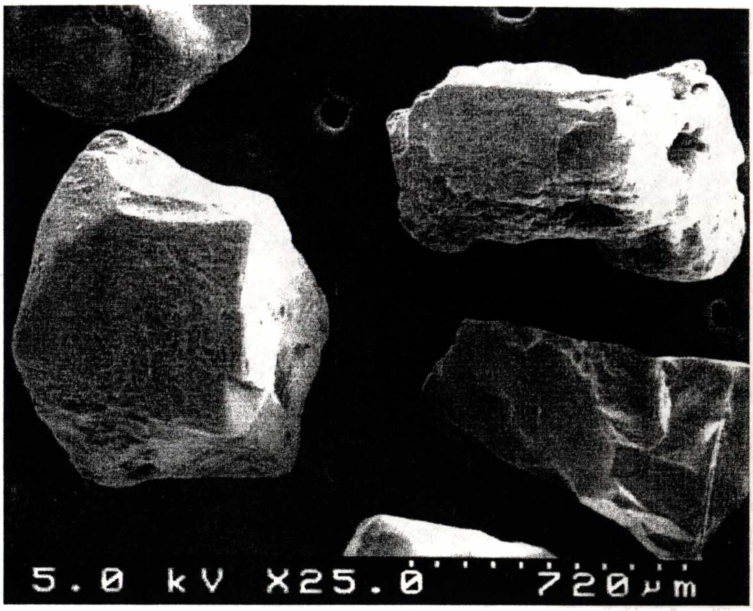


Figure 3-1. SEM image of silica sand grains.

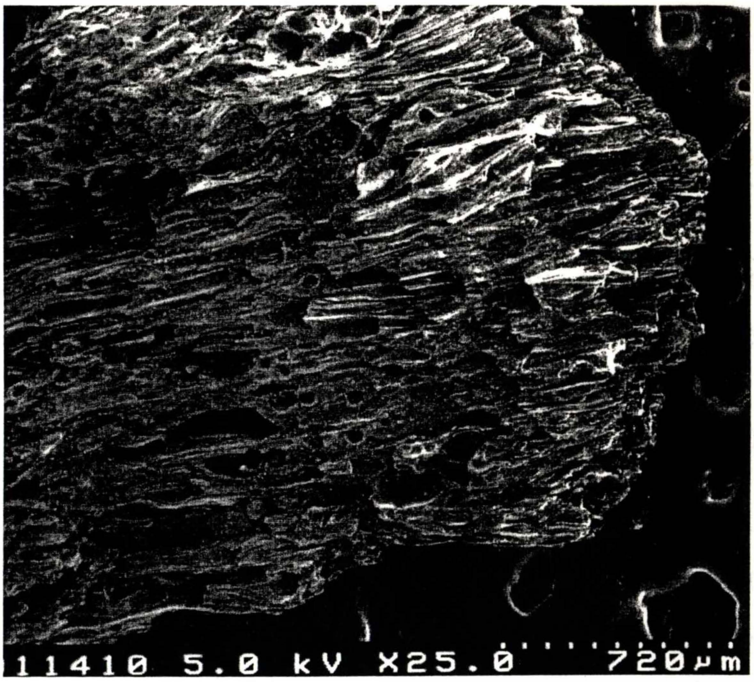


Figure 3-2. SEM image of SS grains.

Figure 3-3. SEM image of TM grains.

Table 3-2. Summary of light microscopic studies.

Medium type	Grain shape	Surface structure
Titanomagnetite	granular	black, smooth edges with fractured layers
Silicon Sponge	angular	grey-white to brown-grey, sponge or froth like, vesiculated
Silica Sand	granular	corny - grey, some visibly crushed with very small exposed pores

The main contrast amongst the media studied was the degree of surface roughness. For the SS media, the surface was roughened by its versicular nature, whereas the TM had a smooth surface, which was presumably the result of extensive weathering in a marine environment.

Images from SEM highlighted the porous surface of the SS grains even further. Silica sand had also fine pores visible on its surface and the grain shape was roughened from fractures, which had not weathered. TM grains were rounder, and there were no pores on the surface. The curved surface showed some evidence of fractures. According to Cooke, (2002) this could be explained by subchoncoidal fracture, a breakage along curved concentric surfaces ie, the TM grain has broken along a curved surface. This might account for the rounding also.

3.3 *Surface area, void space and pore radius*

Data for surface areas, void spaces and pore radii are summarised in Table 3-3.

Table 3-3. Surface areas determined by calculation and BET determination (m^2/g).

Media type and particle size	Calculated exterior area	BET area
TM	1.13×10^{-2}	2.1×10^{-2}
SS 14/25	2.2×10^{-3}	2.2×10^{-1}
SS 7/14	1.1×10^{-3}	2.1×10^{-1}
Silica Sand	2.7×10^{-3}	1.1×10^{-1}

Calculated exterior area values were obtained by assuming spherical particles of diameter equal to average grain size (d) listed in Table 2-6.

The difference between the calculated and the measured values for the SS samples is consistent with the large number of cavities seen on SEM photographs (see Figures 3-1 to 3-3). The internal surface area of these cavities account for a large fraction of the total effective surface area of SS.

3.4 *Bed void volume, wet media density and absolute density*

Bed void volumes, wet media density and absolute density are below in Table 3-4.

Table 3-4. Absolute and wet media density including bed porosity.

Media type	Error	ρ_{wM} kg/m^3	ρ_A kg/m^3	$V\varepsilon$ %
TM (bulk)	± 0.5	3100	4200	11
TM (sieved and washed)	± 0.5	3000	3900	13
SS 14/25	± 0.3	1300	1800	48
SS 7/14	± 0.4	1300	1800	49
Silica Sand 30/60	± 0.3	2100	2400	24

The bed porosity value $V\varepsilon$ of 24, for silica sand obtained in this work is lower than the value of 40 listed for this medium elsewhere (Boller, 1980). It is possible that a larger fraction of smaller grains and the weathered nature of the grains accounted for this difference.

Published porosity values of pumice sands point to average porosities of 60% for thermally altered pumice (Karapiti) and 70 to 75% for unaltered fresh pumice at Tauhara with particle densities of 1975 and 1680 kg/m^3 respectively (Bromley and Hochstein, 2001).

A feature of TM is its high density. Densities (ρA) ranged between 3200 to 4800 kg/m^3 and the variation could be attributed to large amounts of feldspars, quartz and volcanic glass indicated by higher levels of Ca^{2+}, Fe^{2+}, and Mg^{2+} levels (see also Table 3-10 and Table 3-11).

3.5 *Electrokinetic properties*

Results for electrokinetic studies performed by Yang, 2001, are listed in the Table below.

Table 3-5. Isoelectric point (IEP) (Yang, 2001).

Filter media type	IEP	Error
SS	4.1	± 0.5
TM	3.75	± 0.2

Both media have IEP values in the acid pH range indicating that they will be negatively charged at normal effluent pHs.

3.6 *Grain size measurement*

Grain sizes were determined by the dry sieving method described in Section 2-3. Results for SS, TM, silica sand and supporting gravel are summarised in Table 3-6.

Table 3-6. Size range and UC of TM (bulk), TM (sieved and washed), SS 14/25, SS 7/14, silica sand 30/60, and supporting gravel.

Filter Media type	Size range (mm)	UC
TM (bulk)	0.01 - 0.35	1.9
TM (sieved and washed)	0.08 - 0.18	1.3
SS 14/25	0.6 - 1	1.3
SS 7/14	0.4 - 2	2.2
Silica sand 30/60	0.2 - 0.6	2.2
Supporting gravel	0.8 - 8.0	1.4

Plots of particle sizes ranges allowing determination of uniformity coefficients are give in Figure 3-4.

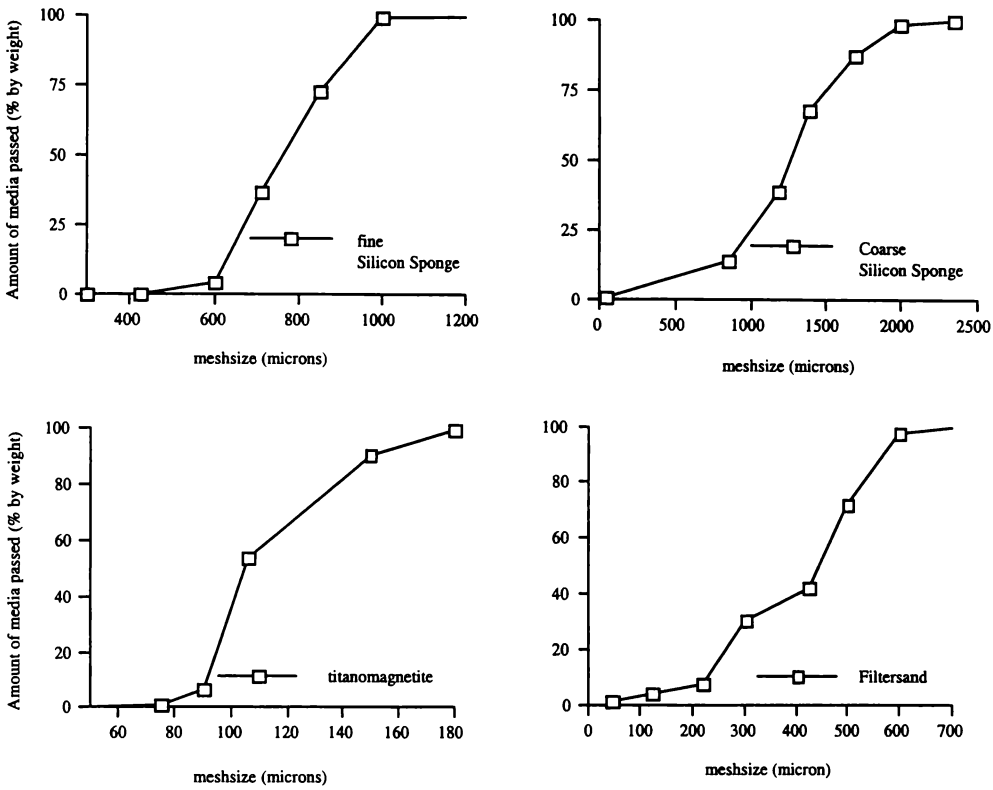


Figure 3-4. Media grain size distributions for TM, SS 14/25, SS 7/14 and silica sand 30/60.

The coarse SS 7/14 gave a larger uniformity coefficient than the value normally recommended (less than 2) (Degremont, 1992). The other materials were within the recommended range.

3.7 *Media heterogeneity including size of atypical material*

Dry sieving was used to study the content and size of atypical material present in the filter media. Visual examination was performed using a light microscope. At least 5 replicate analyses were carried out. Following size fractionation by dry sieving, approximately 5 mL sub-samples from each size fraction were counted and sorted into either typical or atypical grains.

Heterogeneity of the material was assessed following separation by fluidisation. This allowed a comparison of the principal grains, described earlier in Table 3-2 and atypical grains, which often appeared in different colours. A 100 mm deep media bed was expanded by 100% inside a glass column (50 mm ID, 500 mm long) at the appropriate backwash flowrate. In cases of media separation, the height of the settled, atypical material was measured. Data showing media heterogeneity, following filter bed fluidisation are summarised in Table 3-7.

Table 3-7. Proportion and description of atypical grains.

Particle type	Proportion of untypical grains after fluidisation	Description
Silica sand	0.5 -1%, no separation took place using this method	Fine black grains are present throughout the medium
TM	5.7%, no separation, using this method	Shiny crystals, fine grey clay, Silicon sand and red grains are present throughout the medium
SS 7/14	19%, separation visible with impurities at the base	Fine grey and large black grains

Note: A description of the chief minerals expected in New Zealand TM and pumice grains was given in Section 1.3.3.

Data showing the proportion of atypical grains in each size category are shown in Tables 3-8 and 3-9.

Table 3-8. Size and proportion of atypical grains in SS 7/14.

Size category (mm)	Description	Proportion (%) atypical grains
< 0.85	light grey clay dust	50.0
0.85 - 1.18	Black grains	0.5
1.18 - 1.4	Black grains	2.0
1.4 - 1.7	Black grains	5.0
1.7 – 2	Black grains	17.5
2 - 2.36	Black grains	22.5
> 2.36	white grains	28.0

Atypical material in SS was mainly found in the larger sizes, whereas the bulk of the medium (d = 0.5 - 1.5 mm) contained < 5% of atypical grains.

Data showing the size distribution of atypical grains in TM medium are shown in Table 3-9.

Table 3-9. Size and proportion of atypical grains in TM

Size range (mm)	Description	Proportion (%) atypical grains
> 0.063	grey clays	7.0
0.063 - 0.106	grey or yellow	5.8
0.106 - 0.125	yellow-grey	6.8
0.125 - 0.150	yellow-grey	16.0
0.15 - 0.180	yellow-grey	19.6
0.18 - 0.212	yellow - grey	22.3
0.212 - 0.25	yellow - grey	34.7

For both media tested, atypical grains are predominantly larger than the bulk of the media. In the SS media, atypical grains amount to about 20% and contain traces of iron. These grains will eventually settle at the base of a filter bed, partly due to their greater density and partly due to their larger size. Black grains at the base of an operational PCDM filter have been reported previously (Liu, 1997). This process can lead to a lowering of the UC by reducing the amount of larger grains present in the bulk of the filter media. At the same time, the effective grain size in SS 7/14 will be reduced, with less media > 1.7 mm. The large proportion of finer grey material of < 0.8 mm in size is be expected to be lost and washed out by repetitive backwashing. For the TM media, the larger atypical grains look similar to SS.

3.8 *Chemical composition of media used*

All the media used in the study were analysed for total elemental content using XRF (Geoscience laboratory in Ontario).

3.8.1 Major constituents

Data for major constituents (given as corresponding oxides) are summarised in Table 3-10 and are compared with values cited in the literature. Analytical results of 7 pumice samples, collected from other regions in NZ are from a study by Briggs (1993).

Table 3-10. Principal components of TM, SS and silica sand and comparison with published data (all values in wt%).

Oxide wt%	TM Graham and Watson (1980)	TM Dasler (1980)	TM Stokes (1989)	TM This study	SS 7/14 This study	Pumice, Briggs (1997)	Silica sand This study
SiO_2	0.02 - 0.26	N/A	N/A	4.92	71.02	69 - 73	68.37
TiO_2	5.3 - 16.2	9 -14	9.1	8.04	0.27	0.3 – 0.5	0.27
Al_2O_3	1.2 - 8.1	6 - 16.9	3.2	4.09	12.83	13 -16	16.27
Total Fe	62	> 56	75.2	76.59	2.88	2.2 –3.6	2.42
FeO	30 - 43	53 -57		17.72	1.19		1.15
MnO	0.3 - 1.15		3.2	0.58	0.11	0.03 - 0.20	0.06
MgO	1.15 - 6.1		3.9	3.17	0.32	0.3 - 0.6	0.86
CaO				1.75	1.56	1.9 - 2.2	3.85
Na_2O				N/A	4.02	3.1 - 4.7	4.46
K_2O				0.11	2.67	2.8 - 4.2	1.07
P_2O_5				0.31	N/A	0.03 - 0.11	N/A
NiO	trace to 0.1						
volatile loss on ignition				N/A	3.45		1.31

Loss on ignition was obtained following heating of the grains at 1000^{0}C for 60 min. It is not clear how much water from inside the vesicles vaporises with this procedure. Vesicle space in pumice samples can occupy as much as 35 % of the total volume of the grain (Tilly, 1987; Briggs, 1997, Bromley and Hochstein, 2001). FeO was determined by titration.

This study used TM following magnetic separation. If TM sands are collected as beach sands, their magnetite concentration can be highly variable. Samples in column 1 and 2 might therefore not be comparable samples.

Data for trace elements are given in Table 3-11.

Table 3-11. Composition of trace elements in TM, SS and silica sand, all figures in mg/kg

Trace element (ppm)	Silica sand	SS	pumice (Briggs)	TM
Th	4	13	6 - 13	n.d
Nb	3	7	4 - 12	18
Zr	90	183	100 - 250	51
Y	12	29	19 - 38	15
Sr	309	137	180 - 230	25
Rb	42	137	62 -144	10
Ba	531	682	420 - 670	807
Cr	15	10	2 - 6	59
As	< 0.01	21	2 - 5	<0.01
Pb	12	17	10 - 20	8
Ga	17	15	16 - 18	47

Note: Arsenic measured in SS is hermetically sealed inside the vesicular spaces of the grains, therefore it should not be available and not leak out with water. (Hodder, 1997). A leachate test included soaking of SS in water for 7 days, following analysis by atomic absorption spectroscopy. No Arsenic was detected using this method.

There were no significant differences in any of the oxide levels observed for the materials used in the current work. Of the trace elements, SS showed slightly larger concentrations of As and Cr, than pumice.

The amount of iron varied between 55 - 75%. Watson (1979) found a linear relationship in Waipipi iron sands between ferrous and magnetic composition.

3.8.2 Discussion of physical properties

The low value of hydraulic conductivity for the small grain size of the TM medium will limit its use to filtration applications where high flows are not critical, such as intermittent filtration or pulsed bed operations. The porous ceramic used in TMC nozzles had a hydraulic conductivity similar to silica sand and would not be expected to make a contribution to total bed resistance.

The value for bed porosity of SS, is about 20% higher then silica sand, commonly used in granular media filtration. Even higher porosity values ($\varepsilon = 60\%$), have been reported by Bromley and Hochstein (2001) for unweathered pumice sands at the surface of the Tauhara Field and the Karapiti Field (Wairakei). Higher flowrates and longer filter run longer filter run times can therefore be expected. In the case of TM and silica sand, lower values are a consequence of smaller grain size and may be affected by clay fines, which are still mixed in with the media.

Surface roughness and high degree of vesicular spaces observed on SS grains have the potential to provide advantageous conditions for solid particle attachment and may allow the use of larger grain sizes and hence greater solids retention.

Values for wet media density and absolute density provide practical information for filter design and determine backwashing behaviour. The relative low density of the SS points to reduced backwashing energy requirements compared to other media of similar size and allows the use of SS as an upper layer in dual media filtration.

Due to their low iso electric points, both TM and SS can be expected to attract positively charged particles.

The high value for UC (>2) measured for the coarse SS medium and the silica sand may cause layering into smaller grains towards the top of the filterbed, following backwashing. Some large atypical grains may be a further cause for media separation due to density differences.

3.9 *The development of biofilms on filtration media*

3.9.1 Introduction to biofilm experiments

The application of untested media for the filtration of effluents could be complicated by the development of biofilms. While these may be advantageous in terms of improved fluidisation and increased removal of BOD and COD (Addicks, 1992; Masato, 1981), biofilms might also lead to increased resistance to flow through the bed (Characklis and Marshall, 1989; Cunningham et.al. 1989), and blockage of ceramic nozzles.

An experiment was designed to quantify biofilm growth on SS, silica sand, TM, crushed glass and TM ceramic tiles (TMC) immersed in wastewater under aerobic and anaerobic conditions. Partly digested effluent from a septic tank serving a small community (20 - 30 residents) was used. Contaminants in primary treated septic tank effluent have been reported to be between 50 - 100 mg/L TSS and 140 - 200 mg/L BOD (Salvato, 1992; Tchobanoglous, 1991).

3.9.2 Materials and methods

Conventional filtration sand was used as a control. TMC was used because it is the material from which the PCDM nozzles are fabricated. SS 7/14 was used because it is the filtration medium used in the PCDM system. Crushed window glass was used, because it is the main component of TMC nozzles (see Section 1.14). TM was used because it is a possible wastewater treatment media and it is a component of TMC nozzles.

The medium was contained inside white polyester fleece bags (fleece cloth supplied by Maccaferri, Auckland, New Zealand; mesh size 45 μm). For each medium, two bags were used to contain about 150 mL of media. The bags were closed with nylon strings, which were used to suspend the bags in the effluent.

Domestic effluent from the septic tank was allowed to overflow into a square hole (500 mm side length and 1000 mm depth), dug in the clay ground, before being discharged into the existing absorption field (see Figure 3-5). Sample bags were secured to an insulated lid covering the hole. One set of samples was secured approximately 300 mm from the bottom of the hole where the conditions were anaerobic. The other set of samples was secured at a position such that fluctuations of the effluent level in the drainage system resulted in periodic exposure to air. Samples were incubated for 150 days, during the New Zealand summer months from November 1996 until April 1997.

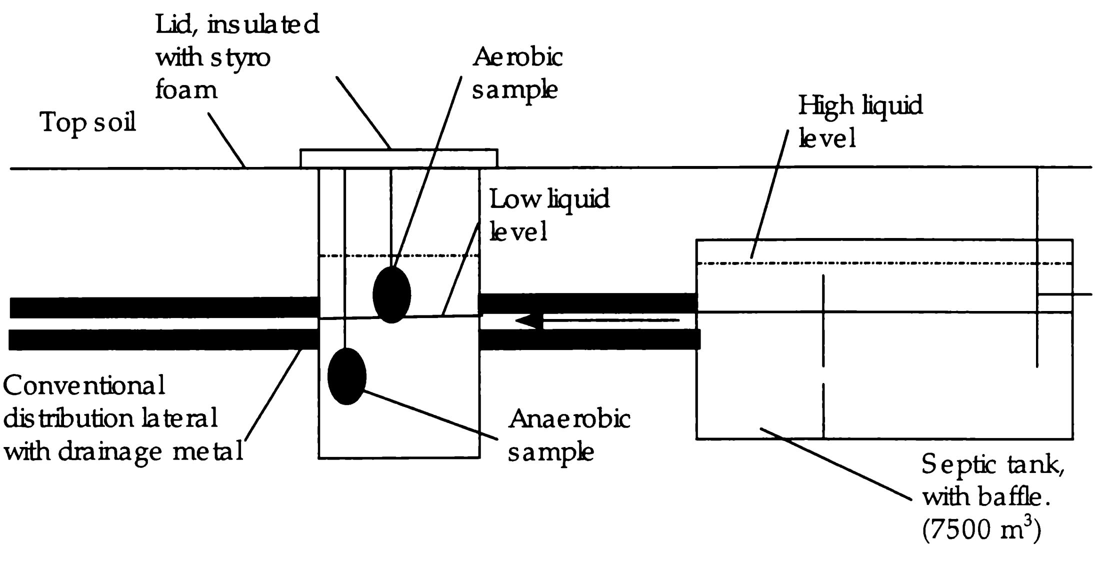


Figure 3-5. Diagram of system used for biofilm studies.

3.9.3 Assessment of biofilm

After incubation, samples of filter media were carefully removed for analysis. Biofilm from a sub-sample of the media was extracted into a phosphate buffer solution, using a sonic probe (Misonix ultrasonic processor xL, supplied by Alphatech, Auckland, New Zealand) at 2.6 ohm for 5 sec. This process was followed by 3 successive rinses with phosphate buffer solution. The protein content in the wash water was measured, following the Biuret method (Bollag and Edelstein, 1989) outlined in Appendix 4.

Another (unwashed) sub-sample was prepared for SEM analysis, following procedures outlined in sections 2.6 and 3-9.

3.9.4 Protein content

Results for protein accumulation in the biofilm are summarised in Figure 3-6 and Figure 3-7 and use surface area and density data summarised in section 3.4.

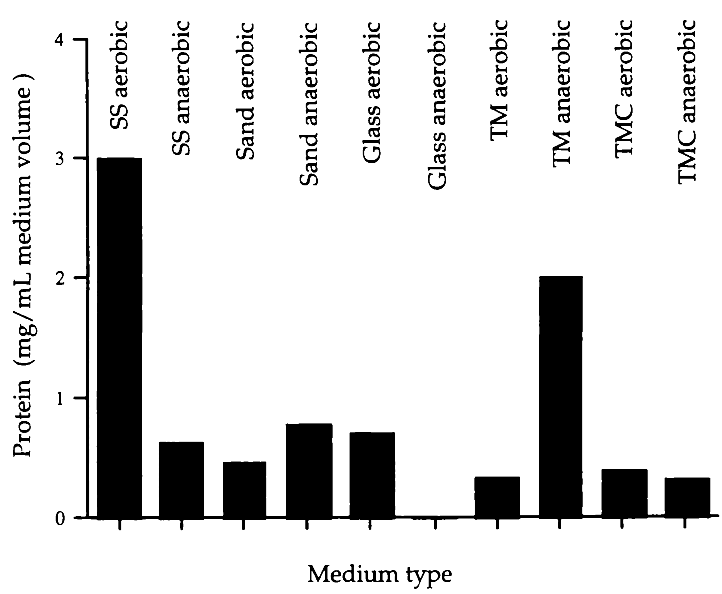


Figure 3-6. Protein uptake per unit volume (mg protein/mL) by various media types.

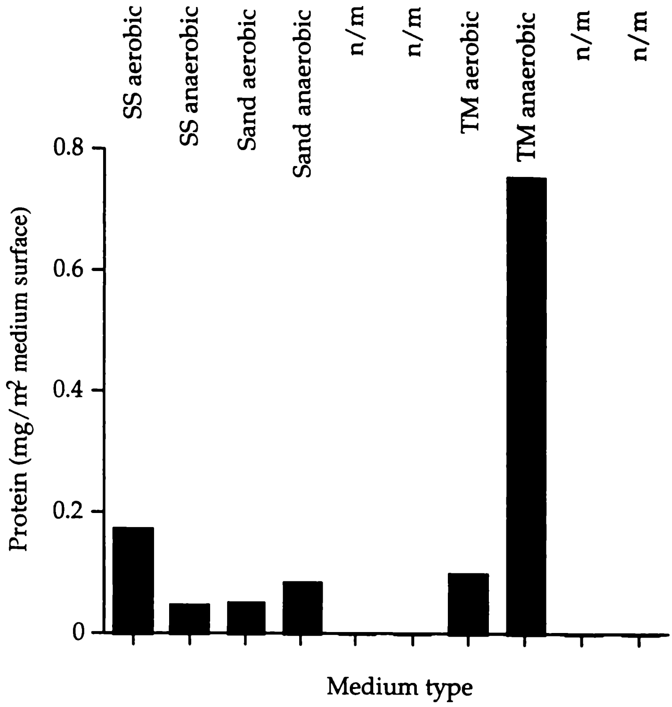


Figure 3-7. Protein uptake per unit surface (mg protein/m^2) by various media types.

Note: No data on specific surface area for TMC was available.

The results show that SS, under aerobic conditions, supported the greatest per volume protein content. Under anaerobic conditions the greatest per volume protein content was observed for TM.

When surface area of each media type is considered, it is apparent that the TM under anaerobic conditions showed the greatest protein content. Stains from a colloidal red material were observed on the polyester containment bags were observed presumed to be derived from Fe. It was not clear whether the Fe originated from the TM or the effluent (water supply contains 1 mg/L Fe). The fact that the other bags remained unstained is evidence of a role played by the TM. It is of interest to note that the TMC under anaerobic conditions produce films with much more protein than the crushed glass. No red staining was observed in the glass samples.

3.9.5 SEM investigation of adsorbed biological matter

Used media grains, following their exposure to effluent systems were examined by SEM and images given in Figures 3-8 and 3-9. Extended sample preparation was used, to preserve the biological matter attached to the grain surface as much as possible. The photos show the SS system after aerobic exposure and the TM system after anaerobic exposure.

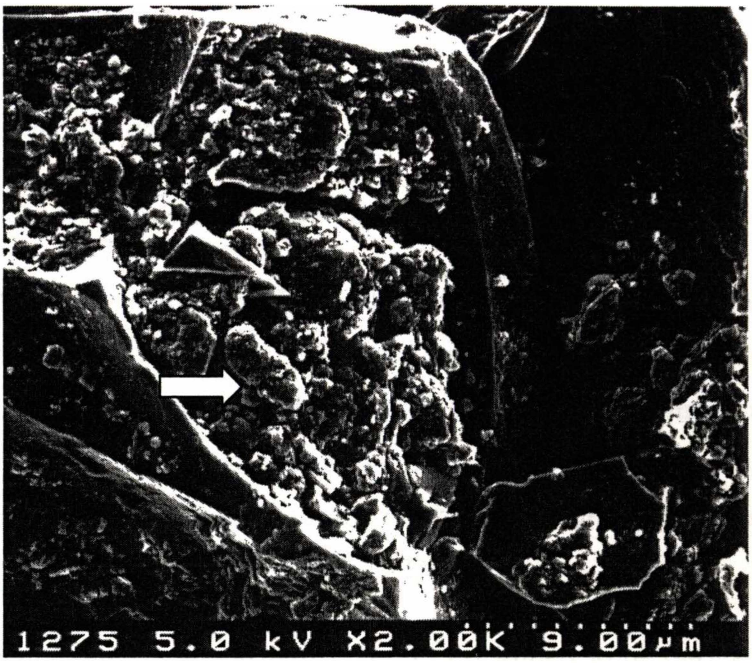


Figure 3-8. SEM of SS, following 90 days of partial immersion in primary treated effluent.

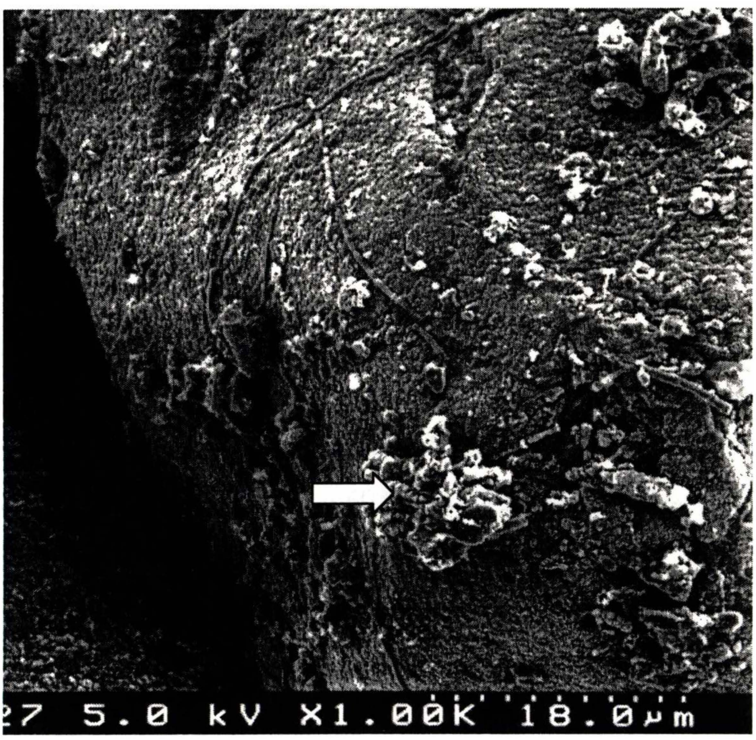


Figure 3-9. SEM image of TM, following exposure to primary treated effluent under anaerobic conditions for 90 days.

The vesicular nature seen on clean SS (Figure 3-2) seem to have disappeared by a layer of biological matter, some of which being identified by a white arrow as bacteria (Curson, 2002).

Despite the smooth surface of TM grains, some attachment of this bio matter occurred. This is also believed to contain bacteria.

3.9.6 Discussion on biofilm experiments

This work showed an affinity of TM and SS for growth of biological matter, with bacterial colonisation occurring on both these surfaces. SS favours biofilm formation under aerobic conditions whereas TM appears to give greater film formation under anaerobic conditions. Glass supports negligible film formation under anaerobic conditions and lower film formation under aerobic conditions. This result is consistent with influence of TM on biofilm formation in TMC.

During pilot plant trials, colloidal red material, presumed to be derived from Fe was observed outside some of the TMC nozzles.

A practical outcome of the work has been the identification of SS as a suitable substrate for biofilm growth. The material could find application in trickling filters or bio-reactors under aerobic conditions. Rapid biofilm formation on the surface of SS grains may offset the initial advantage of the media in particle capturing mechanisms relying on the surface roughness and vesicular spaces.

3.10 *Characterisation of effluents: PTE*

Solid effluent particles can be separated into organic and inorganic components and their individual contribution to the total mass will influence their behaviour during secondary treatment processes. As a general guideline, it was assumed that TSS concentration is made up of about 40 – 60% volatile suspended solids (VSS) and the balance of inorganic solids (Tchobanoglous and Burton, 1991).

Characterisation of effluents focused primarily on particle size of the suspended solids and their size distribution before and after filtration. Particle size was determined by laser diffraction analysis (see Section 2.6.7). Distribution of the particle size in % of the total volume and number on the y-axis is plotted against the relevant size on the x-axis.

Figure 3-10 shows that most of the particles in PTE are < 1 µm, however most of the volume consists of particles between 10 – 50 µm.

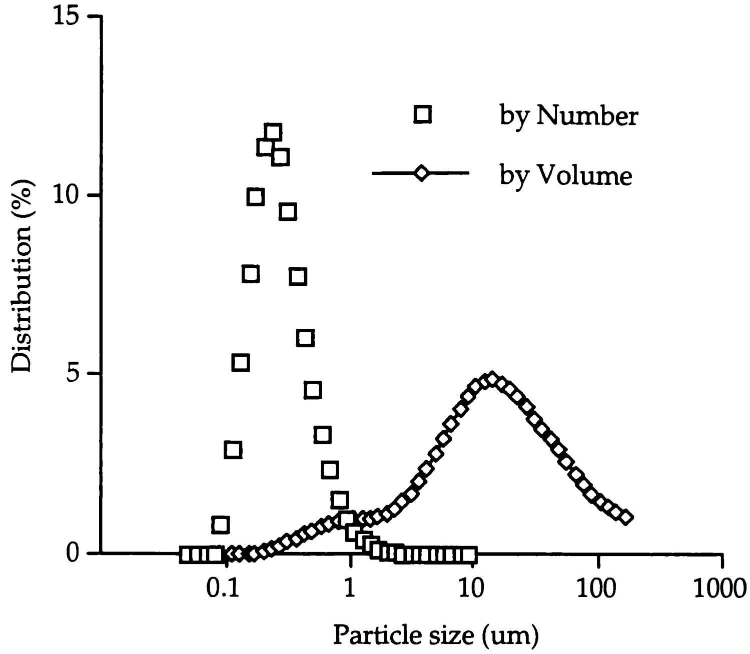


Figure 3-10. Size distribution of suspended particles in PTE. Comparison between size distribution by volume and by number.

The used method of particle size analysis (especially number distributions), are limited to sizes > 0.5 μm.

Particle size can be affected by storage. In the case of PTE, storage of the liquid at room temperature had the effect of increasing particle size distribution, this is shown in Figure 3-11 below.

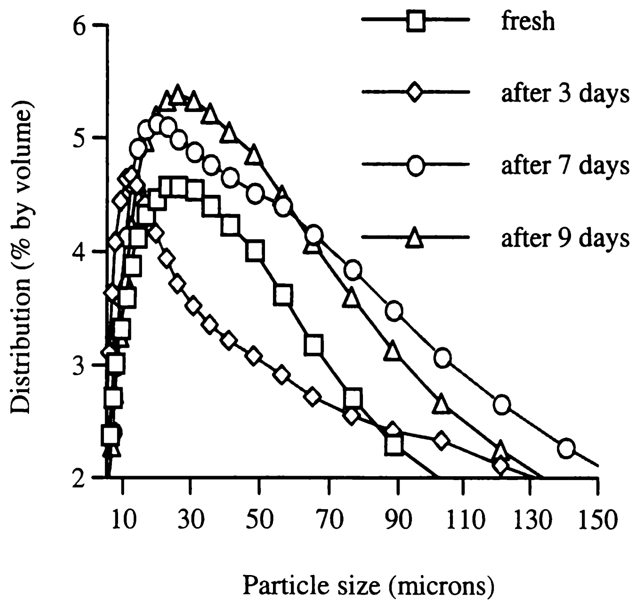


Figure 3-11. Effect of storage on particle size for PTE.

Most particles increased in size with storage. After 7 days, there was an increase of particles > 20 µm and some increase of very large particles (> 100 µm) after 3 days, but a decrease of particles 20 - 90 µm occurred with storage. Storage effects on particle size in PTE made it necessary to use it within 24 h of collection.

3.11 Characterisation of effluents: FDE

Following reconstitution of collected farm manure to a 1% suspension and 24 h settling, particle size distribution was measured by laser diffraction analysis. Results are shown below in Figure 3-12 alongside the size for PTE.

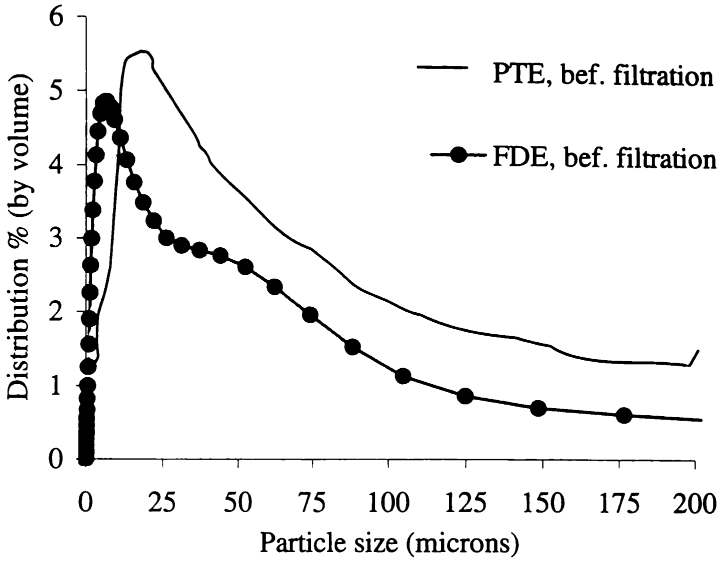


Figure 3-12. Particle size distribution in primary treated effluent (PTE) and farm dairy effluent (FDE).

It is clear the FDE has a larger percentage of smaller particles then PTE. This is likely to have implications in filtration experiments.

3.11.1 Effect of freezing on the size of suspended FDE particles

In order to be able to use the same effluent material it was desirable to prepare suspensions from a common stock material. To ensure freezing had no effect on particle size distribution, batches of fresh cow manure were stored in plastic bags and frozen. The effect of storage was determined by comparing the particle size distribution of the fresh material with samples that had been stored for one and two months. Data are summarised in Figure 3-13 below.

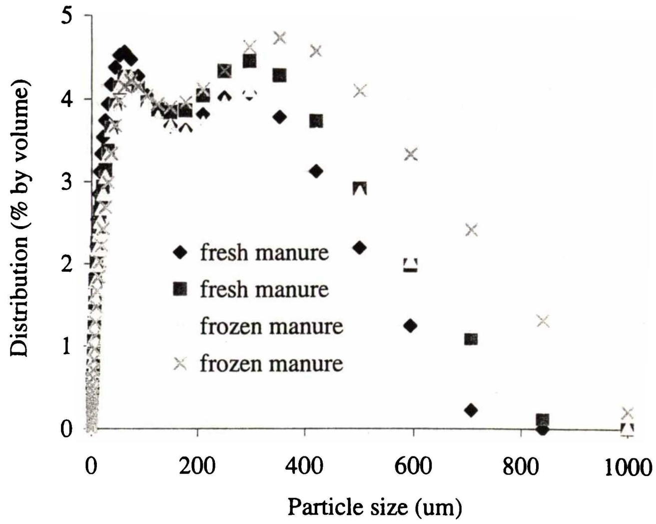


Figure 3-13. Effect of freezing on particle size for FDE.

While it appears that there was some tendency towards larger particle size upon storage, the effect was not large and probably fell within the variability of the experiment.

3.12 Characterisation of effluents: synthetic effluent

The synthetic effluent described in Section 2.8 was tested to determine its suitability for use as a model for PTE and FDE in filtration studies. Turbidity of the synthetic effluent described above was measured at 99 NTU.

Particle size (distribution by volume) was measured by laser diffraction analysis (Malvern Instruments, see section 2.5) and showed, that the majority of particles is between 20 - 50 μm. This is much larger than the particles in FDE, which shows sizes between 5 - 15 μm and it is larger than particles in PTE with sizes between 5 - 25 μm. When the particle size (distribution by number distribution) was compared between the effluent types, a much closer representation was found (see Figure 3-14).

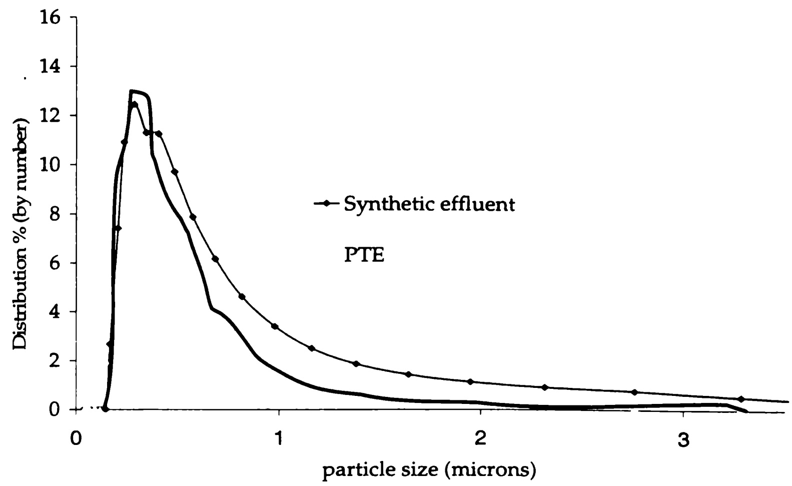


Figure 3-14. Distribution of particle size: comparison between synthetic effluent and PTE.

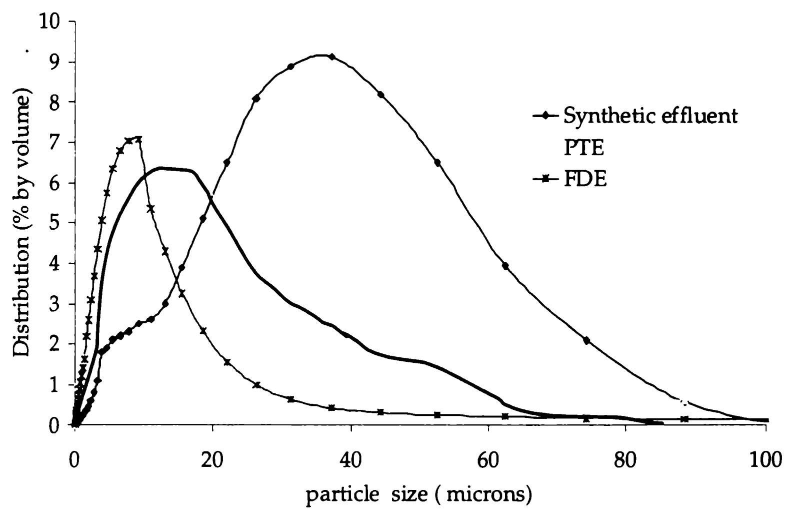


Figure 3-15. Comparison of particle size distribution between FDE, PTE and synthetic effluent by number and by volume.

For the purpose of assessing removal behaviour during filtration, a closer representation of the volume distribution was required then was achievable by synthetic effluent. It was therefore decided that synthetic effluent was not a good model for PTE or FDE and these effluents themselves would need to be used.

3.13 *Summary*

TM, with its small particle size and low void volume, may be useful for the removal of small particles, including pathogens. It usefulness may be limited by slow flow rates and short run times. Shallow beds with frequent backwashing may overcome these problems. On the other hand, SS with its larger particle size, rough surface and large void volume is likely to lend itself to large flow rates, high holding capacity and deep bed applications.

In the next chapter, these ideas will be tested using FDE and PTE. TM will be used in shallow beds. SS will be investigated over a range of bed depths including deep bed (greater than 1 m).

Chapter 4

Bench-scale filtration studies

4.1 *Introduction*

The behaviour of SS, silica sand and TM was compared in laboratory trials using small columns. The principal aim of this part of the work was, to identify filtration parameters (media type, media depth, grain size and media combinations) that would be useful in the design of pilot plant studies, using real effluent. The SS and TM media were of interest because they have been little used in filtration applications involving effluent.

In order to gain information on media behaviour during backwashing, bed expansion behaviour and fluidisation properties were investigated.

With the exception of hydrodynamic studies using water, studies with effluent were generally not repeated. While these experiments cannot be used to provide statistical confidence values they provide sufficient information to indicate trends.

Most of the experiments were carried out with the FDE and PTE characterised in Sections 3.10 and 3.11. In the case of TM, the fine grain size and the high density led us to believe that this medium could be advantageously used in conjunction with pulsed backwashing. This was investigated using high-solids FDE.

Further work with TM included filter bed modifications by blending it with finer material.

Performance measures during the filtration experiments were based on percent turbidity removal, flowrate and changes in particle size distribution. These parameters were discussed in Sections 2.8 and 2.11.

4.2 *Experimental*

4.2.1 Apparatus

For most of the work, glass columns with an ID of 50 mm, similar to those previously described by Martin et.al. (1992), were chosen (see Figure 4-1). Work by others has used 76 mm ID columns (Clark, 1992; Darby, 1991; Cleasby, 1975). For experiments where large effluent through put was expected, column diameter was reduced to compensate for the limited volume of effluent available. The configuration is given below:

Single glass column

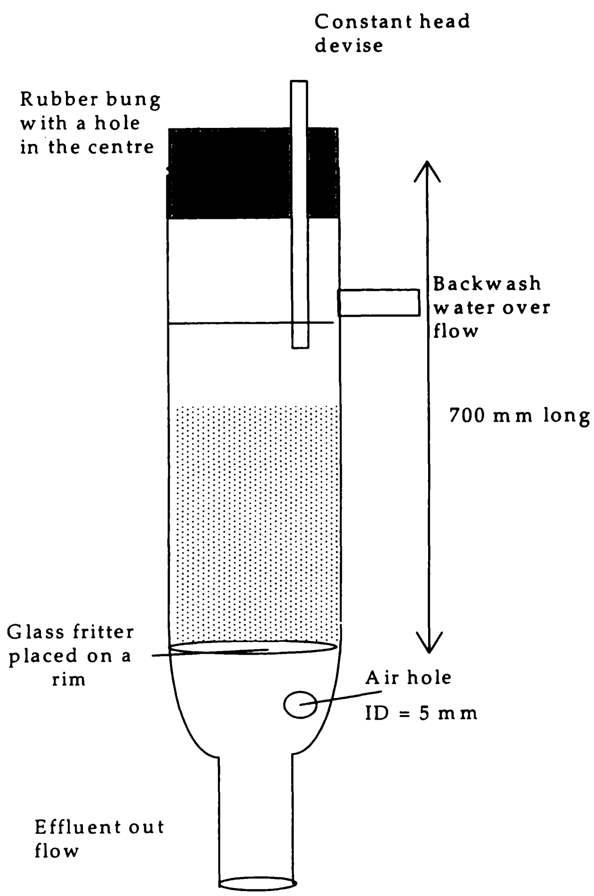


Figure 4-1. Single glass column ID = 50 mm with a porous glass frit at the base and a backwash water overflow on the side.

Effluent was supplied by frequent topping up from a container. During a filter run, the backwash outlet was sealed with a rubber bung. A constant head between 550 and 650 mm was maintained during the filtration process by access to atmospheric pressure through the open-ended glass tube. For bed regeneration, backwash water was supplied from the base through a rubber tube, connected to the water supply. Care was taken to avoid cross contamination to town water supply.

The air orifice at the base of the column, below the porous glass frit, prevented the formation of a plug of effluent, and it ensured atmospheric pressure was maintained at the base of the filter column.

Multiple glass columns

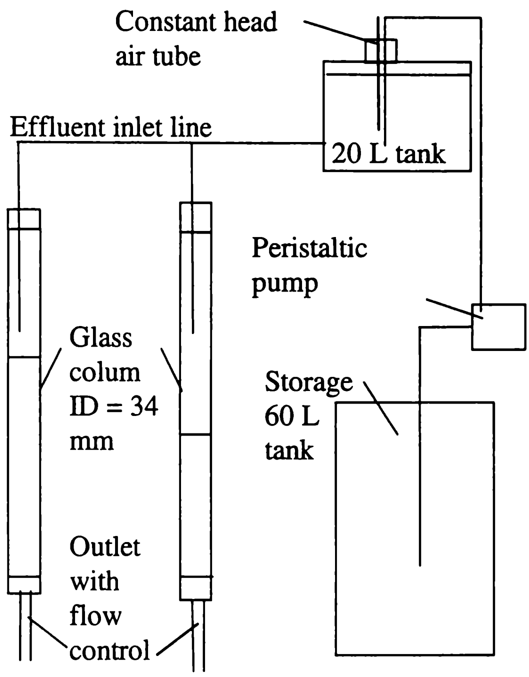


Figure 4-2. Experimental system for multiple glass columns.

The top of each glass column was sealed with a rubber bung and the effluent supplied by a glass tube inserted through the centre of the bung. In order to spread the impact of liquid on the filter surface, the end of the glass tube was bent up-wards. A rubber tube was connected to the base of the glass column, holding a flow-regulating valve.

When loading the long glass columns with the media, formation of airlocks was prevented, by pouring in dry media from the top, while carefully backwashing with water at the same time.

Effluent was supplied from a storage tank at about 2500 mm height and topped up when the levels dropped below the head regulating air tube. Mixing of effluent was maintained by incoming air from the head regulating tube.

An operating constant head of between 550 and 650 mm was used for the TM filters and a constant head of 2500 mm for the silica sand and SS filters. The lower headloss across the bed was achieved when the flowrate was controlled with a valve at the base of the filter column. Flowrate was measured manually with a stop watch during the run.

4.2.2 Effluents

These were described in Sections 2.6, 3.10 and 3.11. Turbidity of FDE ranged between 1200 - 2000 NTU. Turbidity of PTE ranged between 60 – 100 NTU.

4.2.3 Performance parameters

Rather than filter run time, bed void volumes passed filtered were plotted, to compensate for the varying flowrate during a run. A filter run was stopped when the bed blockage led to low flowrates. Turbidity removal efficiency was plotted as percentage of the initial turbidity.

4.2.4 Minimum fluidisation flowrates.

A set of experiments was designed to measure variation of bed expansion with backwash flowrate. As part of this study, the minium bed expansion for fluidisation of the bed was determined. Bed porosity values were calculated to allow correlation between porosity values and backwashing flowrates.

4.2.5 Measurement of the filter bed expansion rate during backwashing

The single glass column was used to measure the required backwash flowrate for a given bed expansion. During backwashing, side-ports not in use, together with the air-hole at the base of the column were sealed with rubber bungs. Backwash flowrate was measured by collecting the overflow through the backwash side-port. Bed height during expansion was determined using graduations on the glass column.

For all three media types, starting with a fixed bed of 110 mm depth, the bed was expanded by incremental increase of the flowrate. Water temperature ranged between 15 - 18^{0}C. In order to observe the onset of fluidisation some of the filter media grains were dyed before mixing with the rest of the medium. Onset of fluidisation was determined at the point where the indicator media moved freely through the bed. Figure 3-3 shows the relationship between flowrate and bed expansion, with the fluidisation point marked by the larger datum point. In the case of TM, two sizes of filter tube diameters with IDs of 34 mm and 50 mm were used.

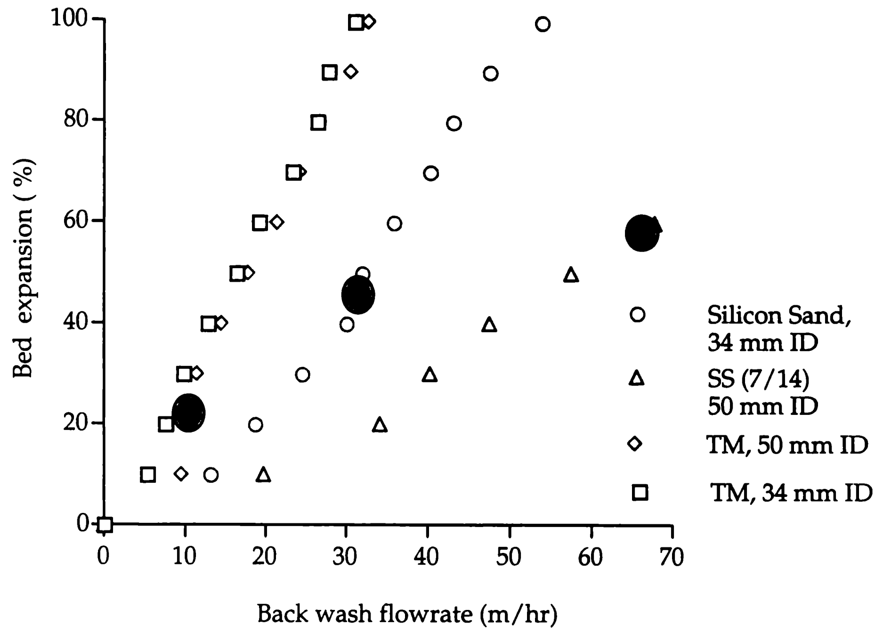


Figure 4-3. Effect of backwash flowrates on filter bed expansion. ● indicates the flowrates at which fluidisation was first observed.

The flowrate required to achieve a certain bed expansion increased with the grain size. The small grain TM bed expanded at very low flowrates and fluidisation point was reached at a very low bed expansion. Column size did not affect the expansion rate for TM. The minimum flowrates required for fluidisation of silica sand and the SS media were 45 m/h and 60 m/h respectively. While not representative of a full size filter, these flowrates are similar to those noted by Hegemann et.al. (1995).

Addicks (1992) developed a model for calculating the minimum bed expansion or incipient fluidisation point, incorporating the Archimedes principle. This gave a backwash water flow rate of 37 - 40 m/h when applied to silica sand of grain size between 0.9 and 1 mm. Lower flowrates would therefore be expected for the smaller silica sand grains (0.2 – 0.6 mm) used in this study.

4.2.6 Calculations of expanded bed porosity

The increase in volume caused by bed expansion was added to the original void volume of the fixed bed (see Section 3.3 for media porosity or void values of a fixed bed). Porosity as a function of flowrate is plotted in Figure 4-4 below.

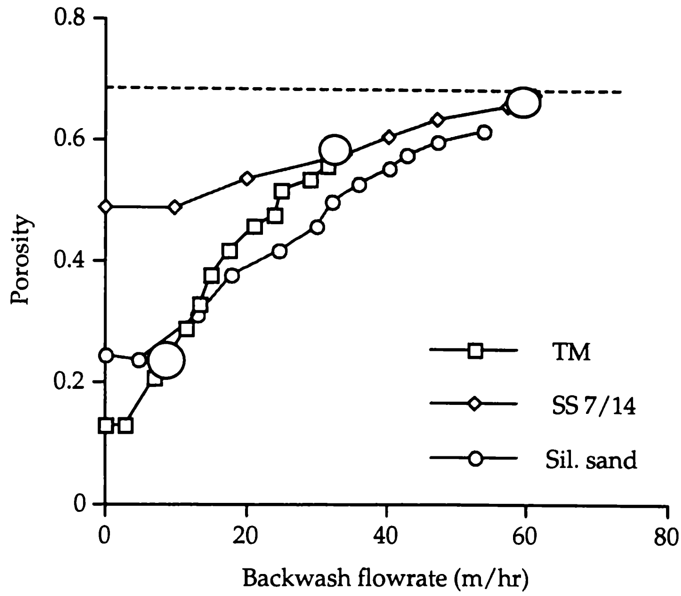


Figure 4-4. Relationship between the porosity value and the flowrate in an expanding filter bed.

O indicates the minimum bed expansion necessary for free movement of the indicator grains, measured during this study, – defines the value for maximum hydrodynamic shear, as reported by Cleasby (1975).

It is clear from the figure above, that both SS and silica sand fluidise at similar porosity but a lower flowrate is sufficient to achieve fluidisation of the finer silica sand. In contrast, the fine and dense TM fluidises at a lower porosity and requires a much lower flowrate.

Cleasby et.al. (1975) reported that the ideal hydrodynamic shear for backwashing common media with grain sizes between 0.4 - 0.7 mm occurred at a bed porosity between 0.68 - 0.71. This value corresponds to the onset of fluidisation observed for SS. However the value obtained for silica sand during this study is lower than the literature value quoted above. An even lower value was obtained for the fine TM.

When effective linear velocities were estimated by correcting for porosity (see Appendix 5), fluidisation of TM, SS and silica sand was found to occur at linear velocities of 61, 69 and 90 m/h respectively. The higher linear velocity measured for silica sand is a result of its lower fixed bed porosity compared with SS.

It appears TM will be useful in effluent filtration where frequent efficient backwashing is important. Reduced backwashing energy will is required. This idea is developed further in Section 4.5, where pulsed bed backwashing experiments are described.

4.3 *Filtration trials using farm dairy effluent (FDE)*

4.3.1 Filtration of FDE by TM

During an initial experiment, filtration of FDE through a single bed, using three different depths of TM was tested. A minimum bed depth of 100 mm was chosen in order to accommodate the TMC tube nozzles intended for use in evaluation unit trials. Additional bed depths of 200 mm and 300 mm were also studied. Results are plotted below in Figure 4-5.

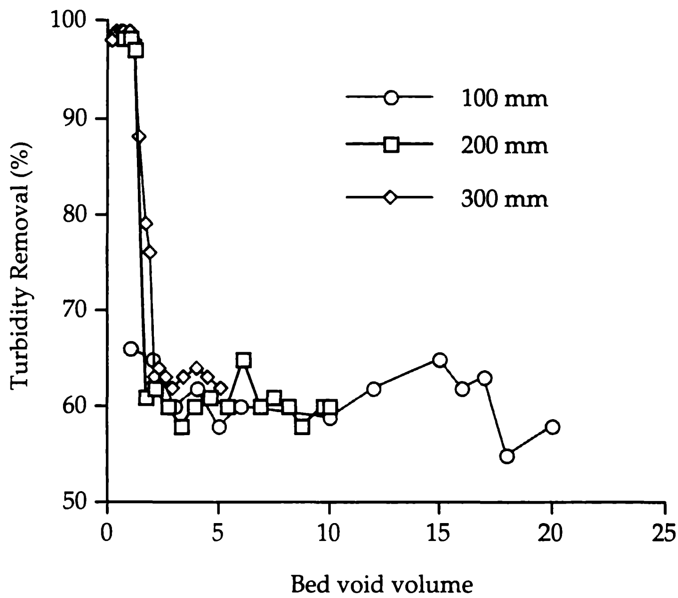


Figure 4-5. Effect of TM bed depth on FDE filtration efficiency.

Initial low turbidity levels resulted from a combination of the clean water trapped in the column voids and from sorption effects. Steady state filtration efficiency in the 100 mm bed of 60% - 65% turbidity removal was achieved after about 3 bed void volumes had passed. No increase in turbidity removal was achieved with deeper beds but decreasing amounts of FDE were filtered, before the flowrate dropped below 0.5 m/h. For the 100 mm deep bed, this occurred after 30 bed void volumes, for the 200 mm deep bed it occurred after 8 bed void volumes and for the 300 mm bed it occurred after 5 bed void volumes.

A comparison was made of turbidity removal by a 200 mm deep bed and the removal by two 100 mm beds in series. Results are shown below in Figure 4-6.

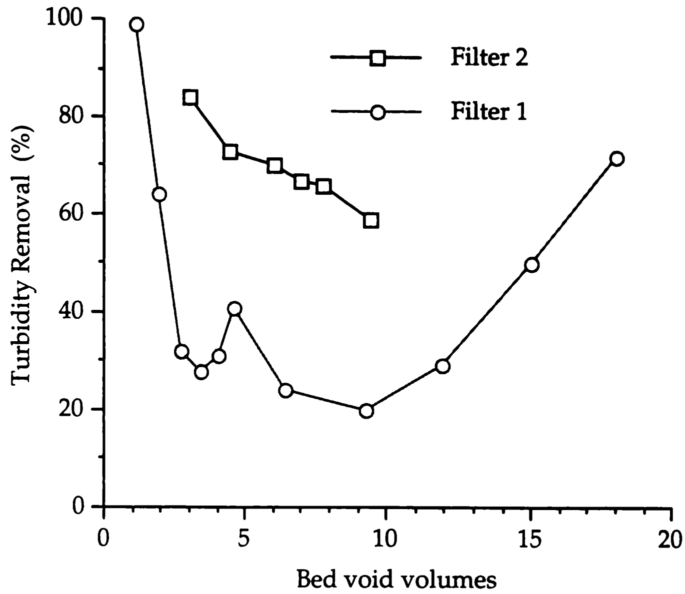


Figure 4-6. Effect on turbidity removal by passing FDE through two 100 mm TM beds in series.

By connecting two filter beds in series, the second TM bed presented an additional bed surface for removal of turbidity.

For completion, an experiment was carried out where effluent after the first pass through a 100 mm TM filter bed was filtered twice more, using the same bed. The results are summarised in Figure 4-7.

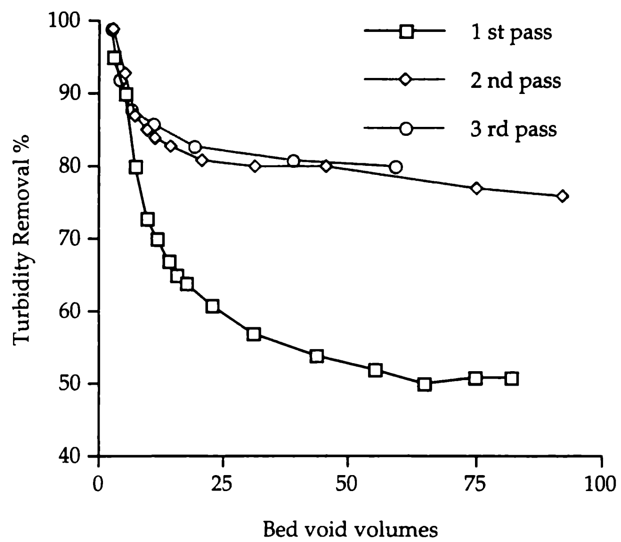


Figure 4-7. Cumulative effect on turbidity removal by repeated filtering of FDE through the same TM filter bed.

Passing effluent through the same filter a second time removed a further 20% of the original turbidity. The third pass did not lead to any further reduction. Straining caused by the conditioned bed probably contributed to reduction of turbidity between the first and second passage through the bed. A very low flowrate of 0.5 m/h was observed during the second and the third passes.

It was of interest to monitor the change of particle size distribution in FDE before and after the filtration through TM. Filtered effluent for this measurement was collected 20 min from start of the run, when a turbidity removal of 50% was indicated.

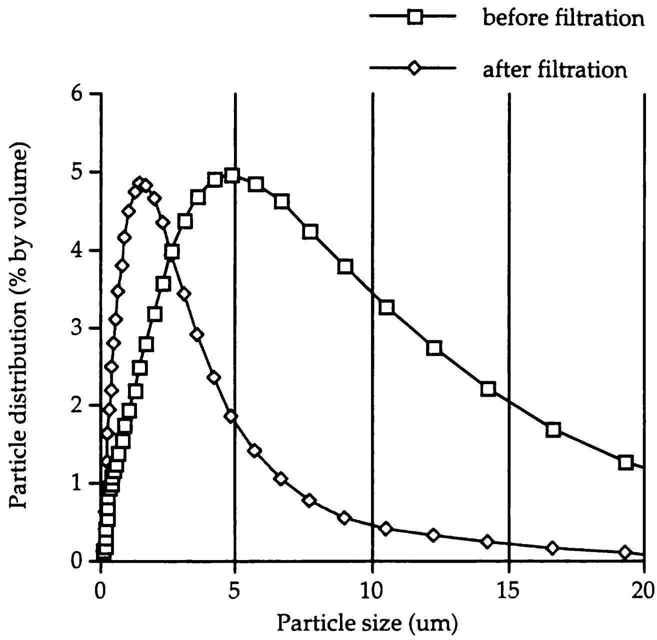


Figure 4-8. Effect of filtration through 100 mm TM on particle size distribution in FDE.

Particle size in FDE ranged between 5 – 150 μm, however only the smaller sizes are shown in Figure 4-8 to illustrate the effect of TM filtration. Most particles > 2 μm were removed. The reduction in the total volume of suspended particles was similar to the reduction of turbidity. The turbidity reduction was apparently due to the removal of the larger particles.

The effect of media modification on particle size distribution is discussed in Section 4.7.

4.3.2 Filtration of FDE by SS 7/14.

A single column of SS was prepared with a bed depth of 1700 mm. At the available head of 2500 mm and with unrestricted flow from the base of the column, the FDE passed through the bed at > 50 m/h, with virtually no reduction of turbidity occurred.

The experiment was therefore carried out under a controlled flowrate of 4 m/h. Results of filtration performance are shown in Figure 4-9.

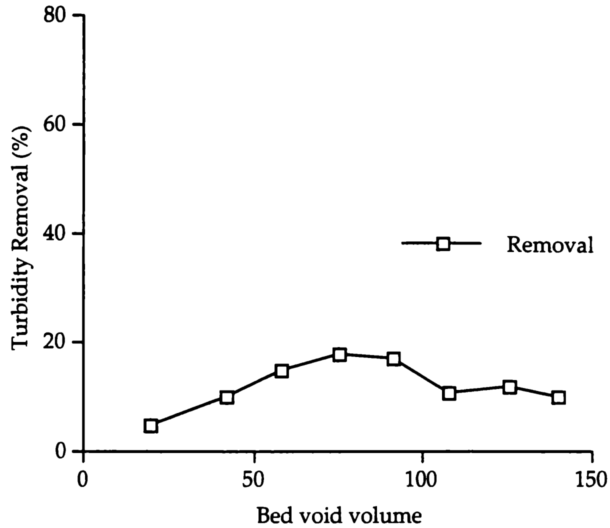


Figure 4-9. Filtration of FDE by SS 7/14 with a bed depth of 1700 mm.

Even by reducing the flowrates, turbidity reduction values of FDE remained low. A small increase in % of turbidity removal occurred over passage of the first 80 bed void volumes but subsequently decreased. There was no indication of bed blockage.

The effect of filtration by SS on particle size distribution in FDE was measured as described previously. Results are shown in Figure 4-10.

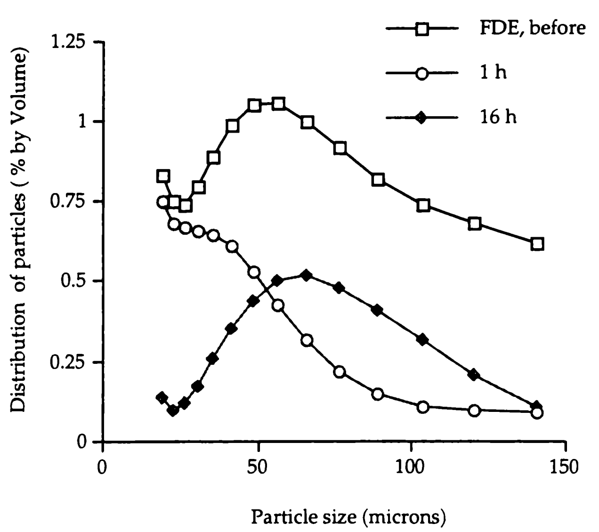


Figure 4-10. Effect of 1700 mm SS 7/14 filtration on particle size distribution in FDE.

During the initial stages of filtration, particles > 50 μm were removed but smaller particles were less affected. When particle size distribution was measured again after 16 h, the distribution of sizes was similar to that of the unfiltered effluent. This indicates break through of the larger sizes after that time.

4.3.3 Filtration of FDE by silica sand

A single column of silica sand was prepared with a bed depth of 400 mm. At a headloss of 2500 mm, and without restricted outlet, a flowrate of 24 m/h was observed. Results on % turbidity removal are shown in Figure 4-11.

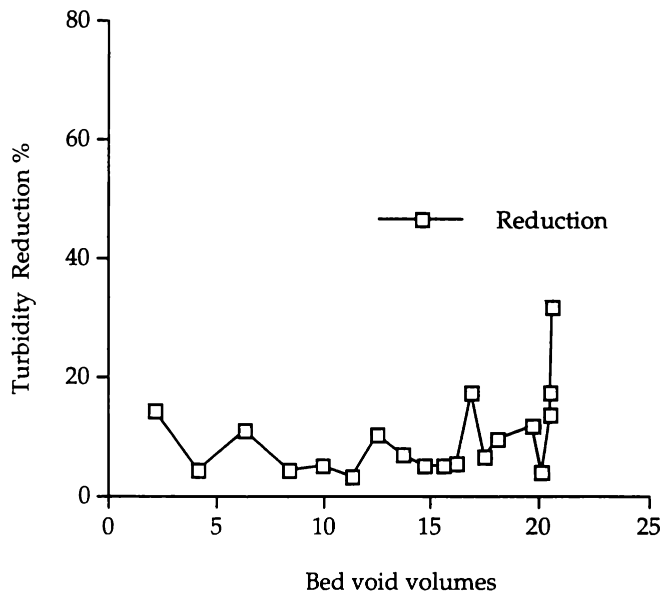


Figure 4-11. Effect of 400 mm silica sand on turbidity removal for FDE.

Turbidity removal fluctuated between 5 - 15% and flowrates dropped from 24 to 4 m/h within the first 30 min of the filter run, indicating bed blockage.

4.3.4 Dual medium experiment

A dual media experiment, using 50 mm TM in combination with 50 mm silica sand, was compared with single medium filters of 100 mm silica sand and 100 mm TM.

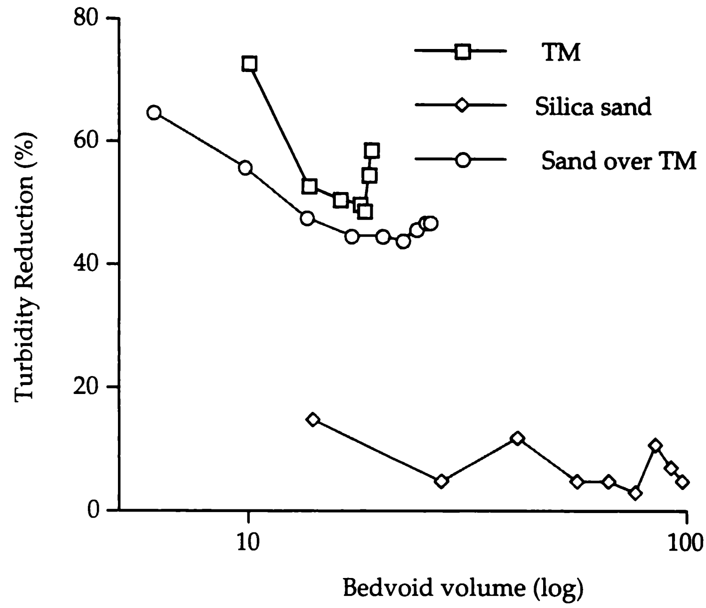


Figure 4-12. Filtration of FDE by 100 mm beds of single medium (individual TM and silica sand beds) and dual media (50 mm TM and 50 mm silica sand).

As was shown earlier in Figure 4-11, silica sand alone is not a satisfactory medium for filtration of FDE. TM alone produced superior turbidity removal, but the bed blocked after about 19 bed void volumes. There was an advantage in using half the bed depth as TM and the other half as silica sand. Filtration rates increased and run times were extended while turbidity removal was reduced only slightly.

4.3.5 Summary of FDE filtration and transport mechanism by various media

Turbidity of FDE was reduced by about 60% by a 100 mm bed of TM. Filtration rates of 1.5 m/h can be obtained for up to 30 min, after which time the bed blockage occurred. Increasing the bed depth beyond 100 mm did not increase turbidity removal. However a second 100 mm TM filter in series with the first did increase turbidity removal. A dual media configuration, using 50 mm silica sand over 50 mm TM resulted in a similar performance as TM alone with the added advantage of longer filter run times, indicating some bed depth effect. Much reduced performance in turbidity removal was recorded when coarser media, such as silica sand or SS were used. Neither of these media was considered suitable for FDE filtration, even with deeper beds of 400 or 1700 mm respectively.

TM medium was able to filter down to, particle sizes of approximately 3 μm. SS was effective for particles < 50 μm.

4.4 *Filtration trials using primary treated municipal effluent (PTE)*

4.4.1 Experimental

For studies with PTE, a system that offered more control over operating conditions was required. A multiple glass column system was set up (see Figure 4-2). Flowrates during a filter run were maintained between 6 - 10 m/h. Turbidity of unfiltered PTE ranged between 70 - 100 NTU. The filter run was stopped when either the flowrate dropped below 6 m/h or when the first 100 bed void volumes were filtered. Rather than filter run time, bed void volumes filtered were plotted.

4.4.2 Effect of media type and bed depth

TM, SS 7/14, SS 14/25 and silica sand media were used in these studies. For the TM medium, filter bed depths of 14; 25; 50; 135; 200 and 400 mm were used. In the case of SS 14/25, bed depths were 300; 560; 920; 1,470 and 1,830 mm. For the coarser grain SS 7/14 the bed depths were 300; 500; 700; 1,000; 1,300 and 1,600 mm and for silica sand, bed depths were 215; 400; 800 and 1740 mm. Flowrates for all media types were controlled between 9 and 12 m/h by adjusting the outlet valve by hand and measure the flow.

Performance of TM filtration: bed depth effect

Turbidity removal is shown for bed depths, ranging from 14 to 400 mm and bed void volume is plotted on a log scale in Figure 4-13.

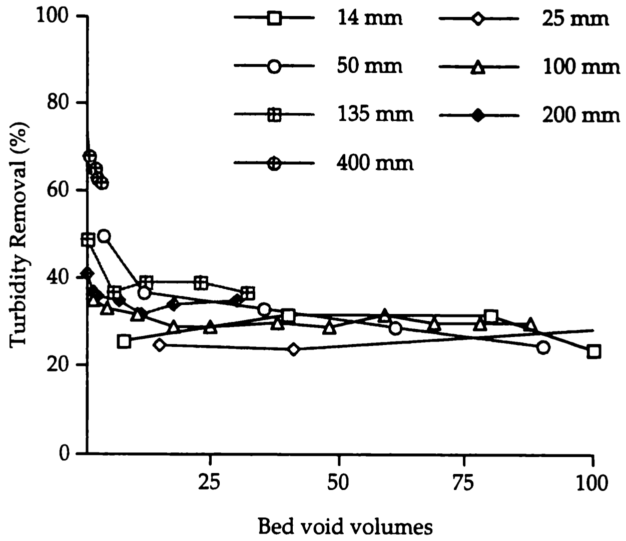


Figure 4-13. Effect of bed depth of TM on turbidity removal for PTE.

TM beds > 100 mm removed about 30% turbidity. At lower bed depths, removal was similar, but break through of turbidity occurred after about 70 bed void volumes. While the 135 mm deep bed removed an additional 10% in turbidity, the increase in depth led to an early blockage after 30 bed void volumes.

Performance of Silicon Sponge 7/14 filtration: bed depth effect

Turbidity removal is shown in Figure 4-14 for bed depths, ranging from 300 mm to 1600 mm. Bed void volumes passed are plotted on a log scale.

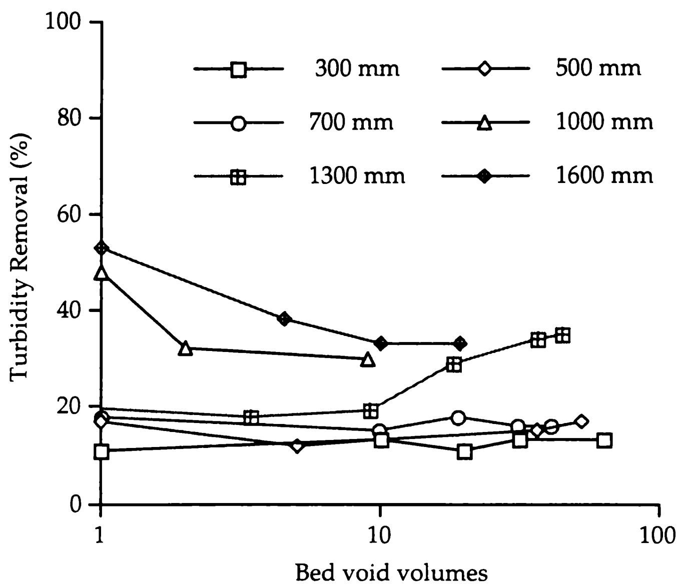


Figure 4-14. Effect of bed depth of SS 7/14 on turbidity removal of PTE.

Bed depths of 700 mm and less removed < 20% of the turbidity from PTE. As the bed depth was increased to 1600 mm the turbidity removal improved to about 35%.

Performance of Silicon Sponge 14/24 filtration: bed depth effect.

Turbidity removal is shown in Figure 4-15 for bed depths, ranging from 300 mm to 1830 mm. Bed void volumes passed are plotted on a log scale.

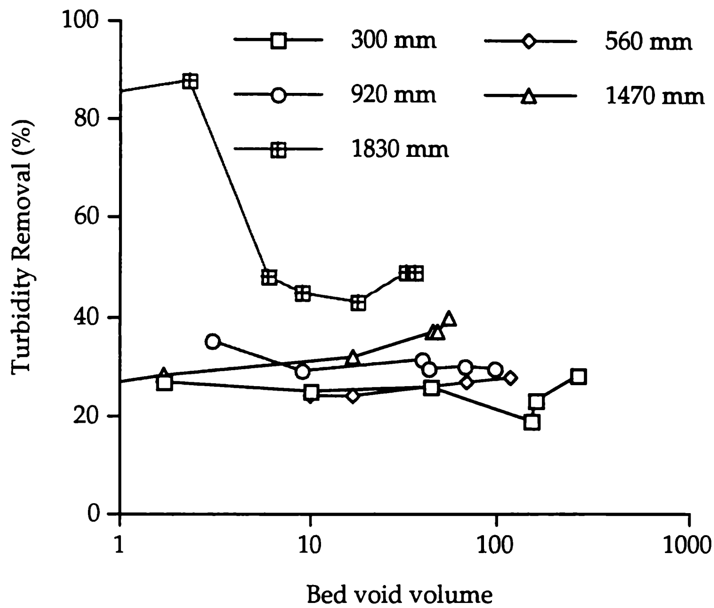


Figure 4-15. Effect of bed depth of SS 14/25 on turbidity removal of PTE.

For the finer SS, shallow beds of 300 mm and 560 mm removed about 25% turbidity. As bed depth was increased the turbidity removal also increased. Turbidity removal of up to 50% was observed for a 1800 mm deep bed.

Performance of silica sand filtration: bed depth effect

Bed depths, ranging from 215 mm to 1740 mm were used. Results are shown in Figure 4-16.

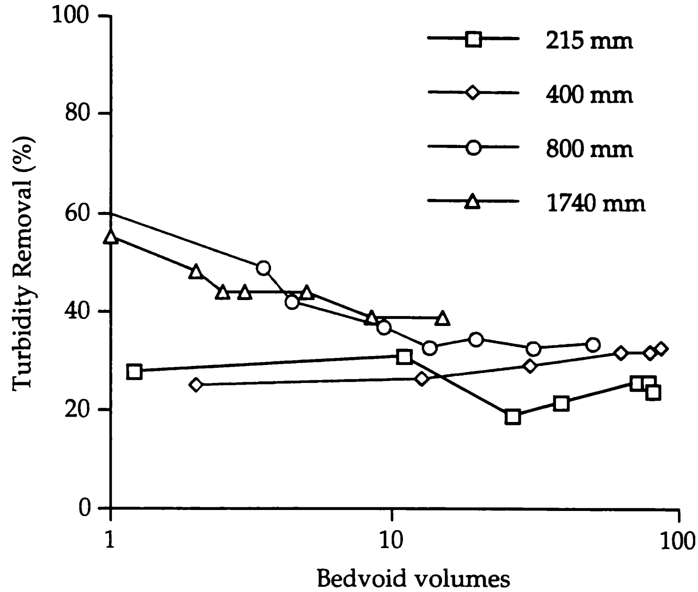


Figure 4-16. Effect of bed depth of silica sand on turbidity removal for PTE.

Approximately 40% turbidity removal was achieved by a 1700 mm deep bed.

4.4.3 Summary of PTE filtration by various media.

Turbidity removal after passage of 20 bed void volumes passed through the filter is plotted in Figure 4-17.

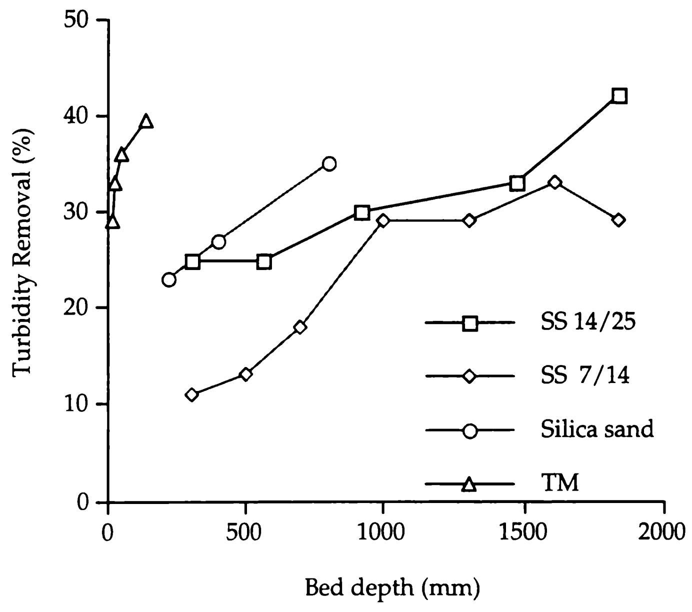


Figure 4-17. Comparison of turbidity removal by various media types after passage oft 20 bed void volumes.

Filtration efficiency increases with increased bed depth and decreased grain size of the media. Fine TM media removed about 30% turbidity even at shallow bed depths. However turbidity break through occurred at TM depths < 100mm.

The effect of bed depth on filtration efficiency for various media can be compared by the use of a parameter called the filtration coefficient -λ (see Equation 1-5 in Section 1.7). Filtration coefficients for the various media are plotted as a function of bed depth in Figure 4-18.

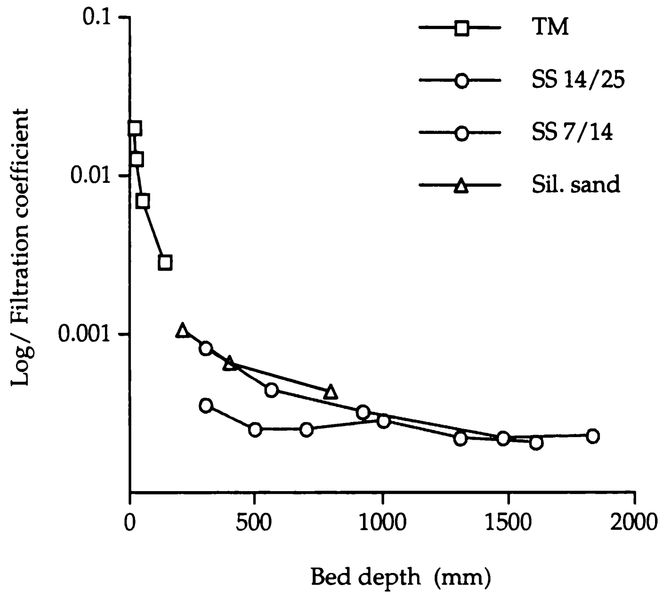


Figure 4-18. Filtration coefficient -λ, (Equation 1-5), after passage of 20 bed void volumes.

The filtration coefficient is largest for the fine TM. Increasing bed depths above 500 mm did not improve the filtration coefficient of the finer SS or the silica sand. At bed depths above 1000 mm, filtration coefficients for the two sizes of SS are similar.

Filter efficiency was calculated over the entire filterbed, resulting in the largest value recorded for the fine TM media. Deeper beds showed lower values then shallow beds, probably due to the smaller size of particles reaching lower regions, thus contributing less to the overall filtration efficiency.

During bench scale trials, the upper limit for turbidity removal from PTE was generally 30%. Deep beds of fine SS achieved approximately 40% turbidity removal.

4.5 *Investigation of pulsed backwashing using TM filtration beds.*

4.5.1 Introduction

TM, because of its high density, small particle size and the low flow velocity required for fluidisation, appeared to be a good candidate medium for use in pulsed backwashed systems. In order to test this possibility, it was of interest to determine the solids distribution through the TM bed and the effectiveness of pulsed backwashing.

4.5.2 Solid distribution in a TM filter bed

Custom designed filter columns were prepared and following filtration of FDE, the medium column was divided into slices of 10 mm thickness each. Trapped solids from FDE were washed from each medium slice and the TSS content determined, see section 2.11. In the first experiment, a single column was prepared from clear PVC (ID = 25 mm). The base was conical and glass wool held the medium in place. It was loaded with 120 mm deep TM. FDE was supplied through a clear tube attached to the top, providing a head of 550 mm. The used medium plug was removed with a plunger from the base, after the coned bottom was cut off. Figure 4-19 illustrates the distribution of FDE solids.

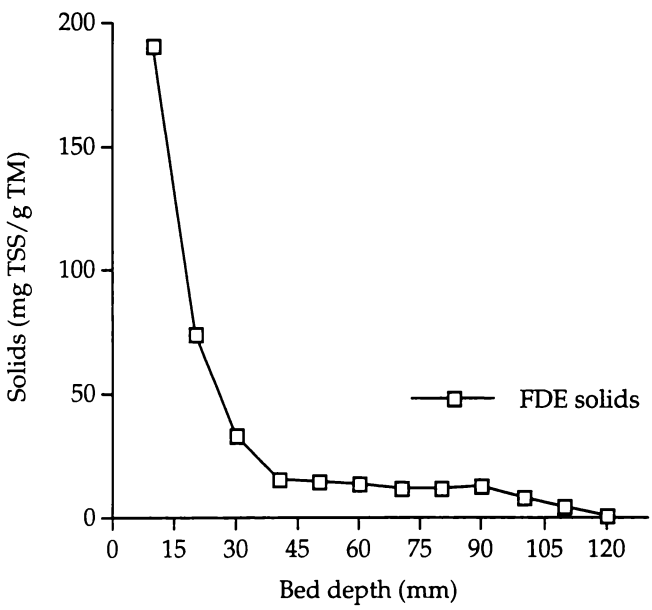


Figure 4-19. Distribution of FDE solids in a 120-mm deep TM filter bed.

Solid penetration decreased sharply over the top 30 mm of the filter bed. Below this depth the low solids content was evenly distributed. Total filtered volume of 370 mL at 1900 NTU supplied about 48 mg solids. Solid recovery, based on 60% turbidity removal of 28 mg was estimated. A solid loading rate in the filter bed of $2kg/m^3$ TM was calculated.

A larger diameter column (ID = 60 mm) was used in a second experiment. A sealed vertical split through the middle of the column allowed ease of media removal at the end of the run. At the base of the column, a rubber stopper held the media in place and a plastic tube through the centre with glass wool on top provided the outflow for the filtered effluent. Bed depth was 190 mm and headloss across the bed was 560 mm. A suspension of FDE with a turbidity of 1900 NTU was filtered until the bed blocked. Any unfiltered effluent was removed carefully from above the filter bed with a tube attached to a plastic syringe. The drained filter bed containing the used media was frozen for 3 h. To remove individual slices of the media, the sides of the PVC pipe were slit open and sections were cut with a custom shaped steel blade. Results for TSS content of each filter bed section are shown in Figure 4-20.

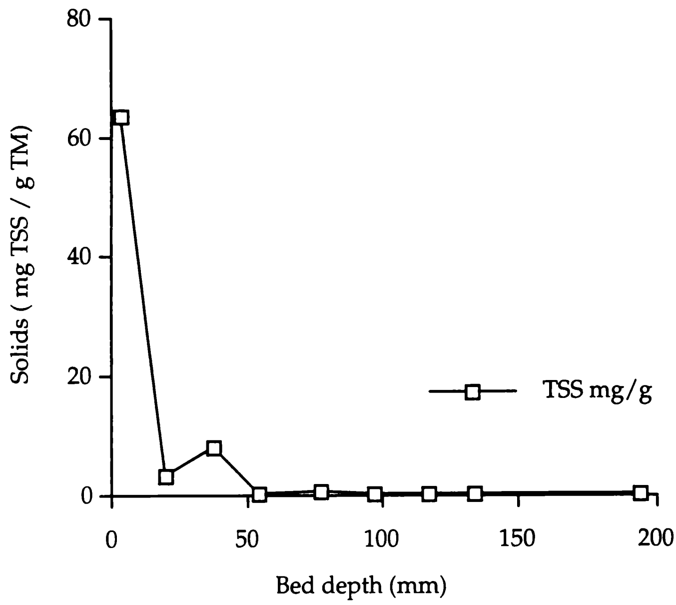


Figure 4-20. FDE solid distribution in a 190 mm deep filter bed of TM.

As with the smaller column, Figure 4-20 shows that the majority of solids accumulated on the filter surface and only a small proportion penetrated deeper than 20 mm.

Following this experiment, particle size distribution of the solids contained in each section of the filter bed was measured by laser diffraction. Results are displayed in Figure 4-21.

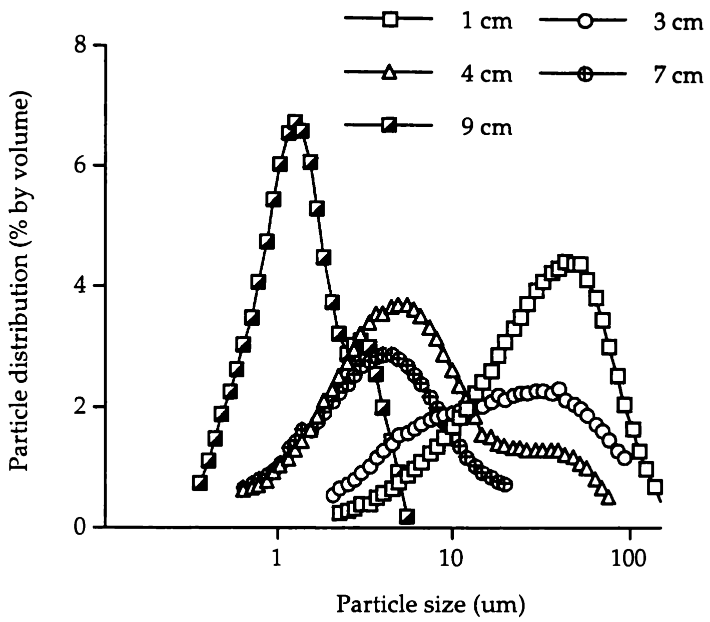


Figure 4-21. Distribution and sizes of solid particles from FDE in a used TM bed.

The majority of particles (by volume) from the top 30 mm of the bed were between 9 - 100 μm in size. Particle sizes decreased with depth. At a depth of 9 mm, virtually no particles > 3 μm were found. This confirms the 3 μm particle size cut–off point also shown in Figure 4-8.

4.5.3 Transport and adhesion in TM and SS beds

Distinction between transport and adhesion steps can be explained using results provided in Figures 4-19 and 4-20. The fine TM medium leads to an early formation of a surface cake where most of the particles are being strained by the small pores of this medium. This is shown in Figure 4–20 where trapped particles from the different unit bed elements were washed out and quantified. The top 10 mm retained mostly particles >20 μm, at 40 mm depth most particles >10 μm were removed and virtually no particle >3 μm appeared at 90 mm depth. The coarse SS medium implies different capture mechanisms. Particle accumulation occurs over the top 500 mm bed depth. The particles remained trapped in these top layers and did not rearrange during longer runtimes. Pore clogging may occur over this bed depth, preventing any further rearrangement of particle clusters. Ripening of the coarse media (SS 7/14) is an effect of gradual pore clogging over time. Perhaps coarser media would allow a more even distribution in a deeper bed, thus taking advantage of the greater capacity that a deeper bed has to offer.

The relatively poor filtration efficiency (up to 50% removal) of SS 7/14 medium may be a reflection of the amount of TSS occurring as aggregates large enough to be retained by the relatively coarse SS medium. The addition of flocculant to increase the percentage of aggregated particles could be expected to improve filtration efficiency.

4.5.4 Pulsed bed operation

Previous experiments (see Section 4.3) showed that TM is easily fluidised. Only a small amount of backwash water is required to achieve the incipient fluidisation point. Lifting of the "surface blanket" layer of solids formed during a filter run, should therefore be achievable with low volumes of backwash water.

Pulsed bed operation, described earlier in Section 1.12, was demonstrated using FDE (1800 NTU and 120 mg/L TSS, see Figure 2-6). Bed depth was 100 mm TM and a single glass column with an additional side-port at 5 mm above the media surface was used.

Method:

Filtration of FDE was continued, until flowrates reduced to < 0.2 m/h. This occurred after 800 mL (67 bed void volumes) containing about 96 mg solids had passed. Approximately 57 mg of the solids would be retained in the bed, (assuming 60% turbidity removal). Results from the filter bed slicing experiments (see Figure 4-19 and 4-20) show that most of the solids accumulate in and on the top 20 mm of the TM filter bed. Following draining of the partly blocked filter bed, to about 2 mm above the filter surface, the pulsing operation was started. The air hole below the glass frit of the filter column was closed. This allowed trapped air from the under drain compartment to form a gas bubble, which was forced up through the filter bed by the backwash water pulse. A 5 sec backwash water pulse at a flowrate of about 10 m/h provided 12 mL water to flush the surface blanket from the filter medium. This slurry mixed with the remaining 5 mL unfiltered effluent from the top was drained away through the side-port, as soon as the media had settled. This process resulted in a solid concentration of 0.3% in the backwash slurry. Higher backwash slurry concentrations could be achieved by better control of residual volumes. After each pulsing cycle, initial flowrates decreased, from 2.5 m/h for a clean bed to 1 m/h after the fifth pulsing cycle. A full backwash lasting 3 min was then necessary to re-establish initial flowrates.

4.6 *Strategies for improving filtration efficiency*

4.6.1 Effect of additions of finer material

Fine TM medium could possibly be used as filter aid, because of its similarities with magnetite sand (Fe_3O_4) which was used as a flocculant, because of its positive charge favouring adsorption of negatively charged particles of impurities, see more in section 1.9.

Improvements to the filtration efficiency were tested, using TM in combination with additions of finer material. The effect of flowrate and turbidity removal efficiency was measured, using FDE. Particle size of the retained solids was also of interest.

Effect of TM fines

Incremental additions of very fine TM were used to modify the TM filter bed. Experiments were carried out, using single columns.

Method: In a ring-mill (supplied by Rocklabs, New Zealand), TM medium was milled for 1 min at 7000 rpm. This reduced the mean grain size from 150 μm to about 30 μm. Between 1% and 10% (by mass) of the milled TM was blended with dry TM.

Turbidity removal was tested with FDE (1200 NTU). The filter run was terminated when the filter blocked. Turbidity removal % is shown in Figure 4-22 and flowrate is given in Figure 4-23.

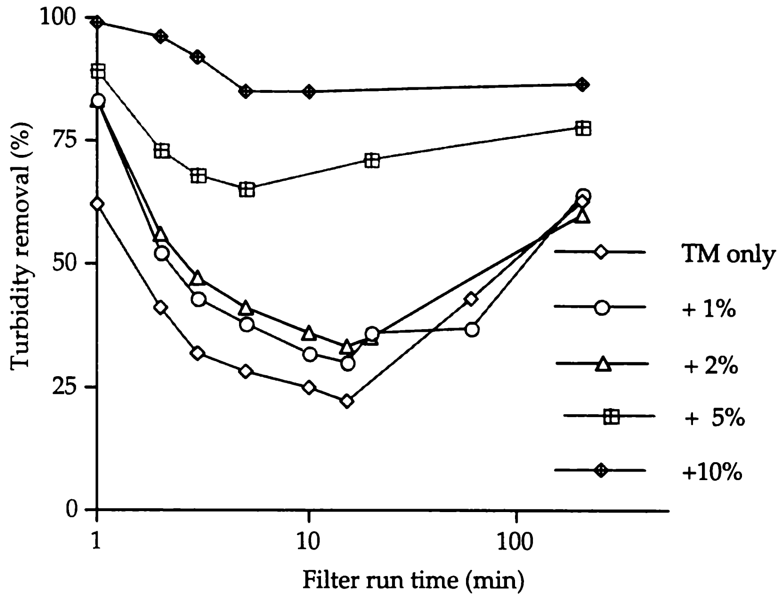


Figure 4-22. Bulk TM containing 1% - 10% milled TM. Effect on turbidity removal from FDE.

A small increase of turbidity removal was achieved, following the additions of 1 and 2% fines. Following the addition of 5% fines, the turbidity removal doubled from about 35% to 70%. Addition of fines decreased the flowrate as is shown in Figure 4-23.

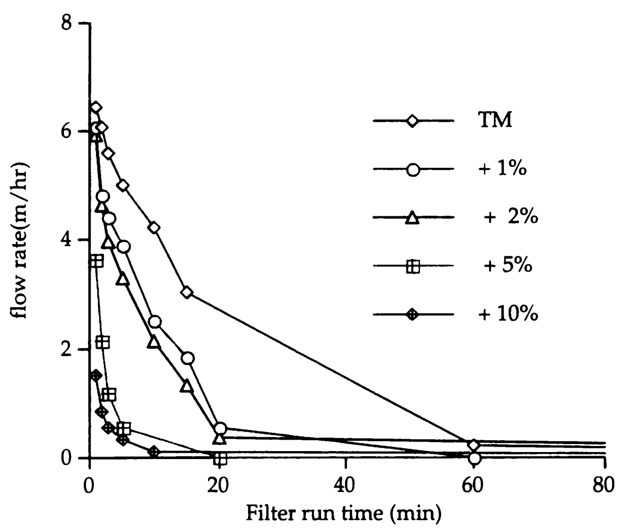


Figure 4-23. Effect of fines addition on the FDE flowrate through TM

The effect of TM fines on particle size retention is shown in Figure 4-24.

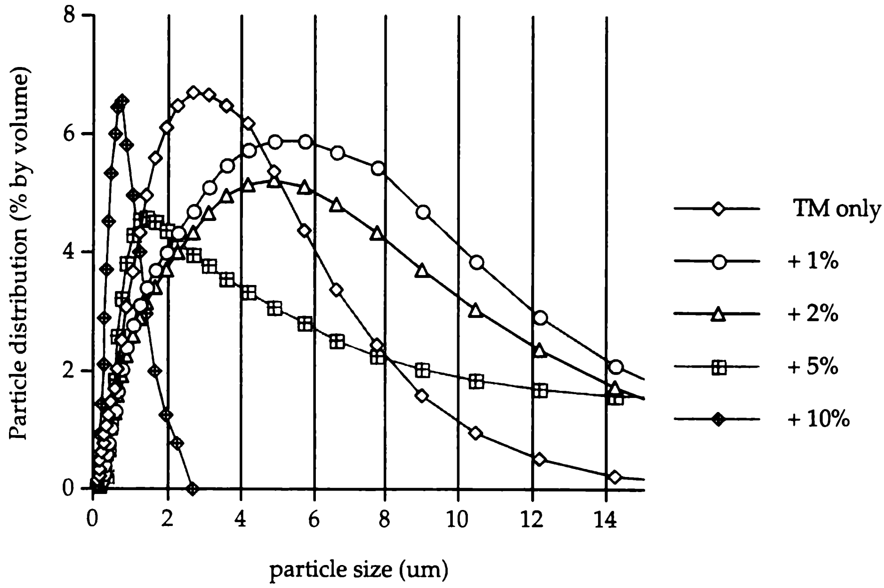


Figure 4-24. Comparison in particle size between filtered and unfiltered FDE and the effect of finer TM on the retention of larger particles.

Addition of 10% TM fines virtually eliminated particles of size greater than 2 µm, following the addition of 5% or more fines. However flowrates were too low for this system to be of practical use.

Flowrates through fine media have been improved by magnetic stabilisation. Work reported by Yang et.al. (2000) on TM showed an increase of hydraulic conductivity from about 0.3 to 0.7 x (10^{-3} m/s) in a bed expanded by 25% and vertically stabilised. The potential therefore exists to capture particles in the 1 – 2 µm size range, matching bacterial contaminants in wastewater streams.

The effect of calcium carbonate fines

Filtration of pond water using TM in a small pilot plant at WFS showed a very high turbidity removal of (up to 90%) from pond water (Hill, 1996). In order to find the contributing factor for this behaviour, the pond water was analysed by Atomic Adsorption Spectrometry (AAS). Unusually high levels of Ca were detected. The Ca was believed to be due to suspended $CaCO_3$. It was therefore decided to investigate the effect of $CaCO_3$ on filtration.

The particle size of $CaCO_3$ (from BDH, AnalaR) was in the range between 0.15 - 0.6 μm. It was combined with TM in two ways: added on top of the TM filter bed to a thickness of about 1 mm and blended with the media. The effect of $CaCO_3$ on the turbidity removal is shown in Figure 4-25 and the effect on flowrate is shown in Figure 4-25.

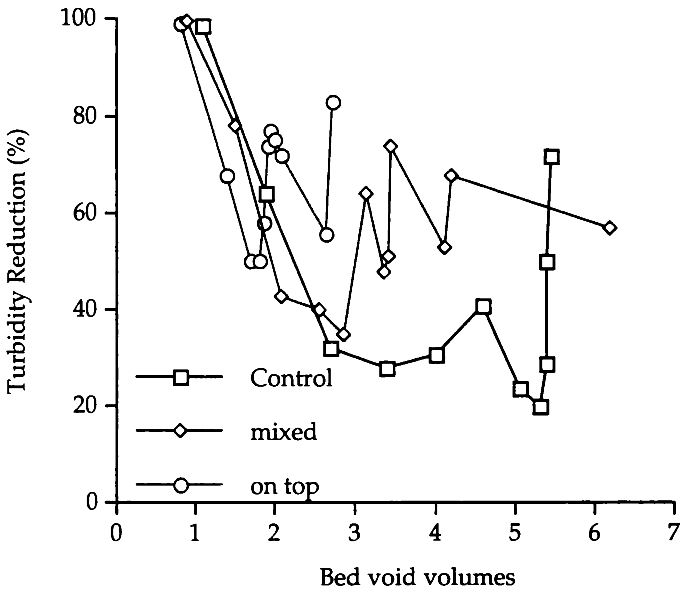


Figure 4-25. Effect of $CaCO_3$ on filter performance in conjunction with TM.

Improved turbidity removal, following the addition of a layer of $CaCO_3$ to the TM bed surface was observed. This could be the result of either a flocculation of suspended solids by the $CaCO_3$ Ca or increased straining by the fine $CaCO_3$ layer on the filter bed surface.

The effect on flowrate of the different configurations of $CaCO_3$ is shown below in Figure 4-26.

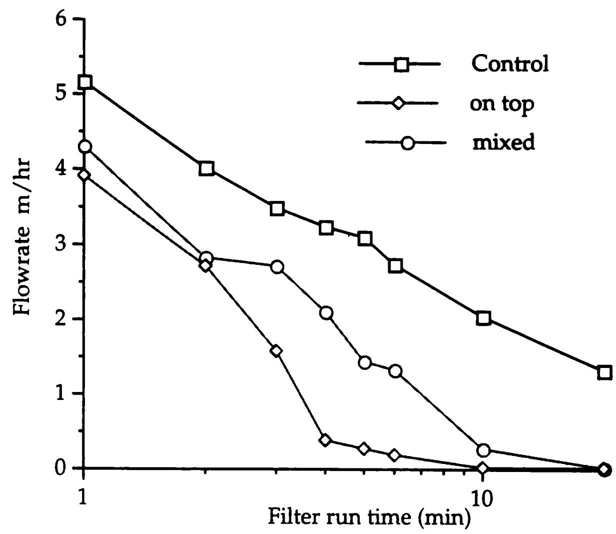


Figure 4-26. Effect of $CaCO_3$ fines on flowrate through a 100 mm TM bed.

A decrease of flowrate occurred when $CaCO_3$ was placed on top of the TM bed. Blending it with the media had less effect on the flowrate. It is apparent that the layer of fine $CaCO_3$ particles on top of the bed formed a filter mat causing improved filtration performance but reduced flows.

4.6.2 Flocculation experiments

While it is not envisaged that chemical flocculation alone should be considered as a treatment option, the following exercise served the purpose to obtain a figure on cost for daily amounts of coagulants required to reduce turbidity levels.

A jar test was performed to test the relative effectiveness of PAC and PFS flocculants. Data are summarised in figures 4-27 to 4-29. In both cases an initial increase in turbidity was followed by a decrease in turbidity at higher dose rates. This was interpreted to mean that even low additions of flocculant resulted in particle aggregation with an associated increase in turbidity. At higher additions particle settling and turbidity decrease was observed. Generally PFS performed better than PAC.

Odegaard and Liao (2002) have recognised that by adding coagulants before settling, cost savings can be achieved by reducing secondary treatment costs.

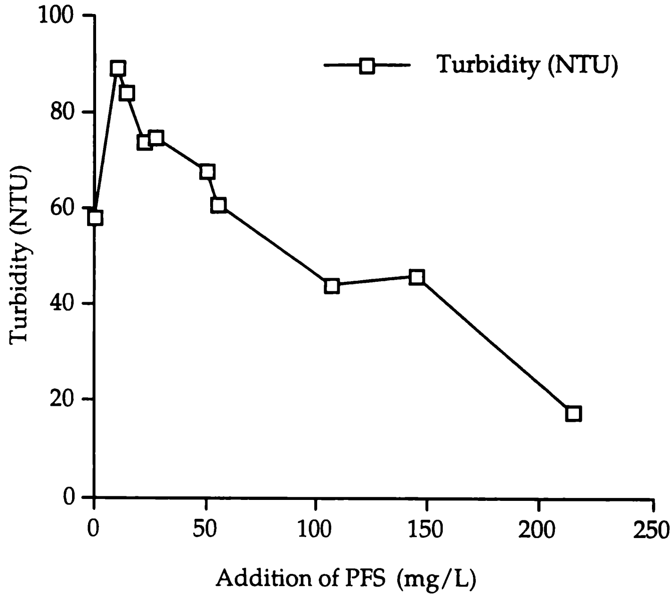


Figure 4-27. Jar test, using PFS (supplied by WFS, Hamilton) on PTE.

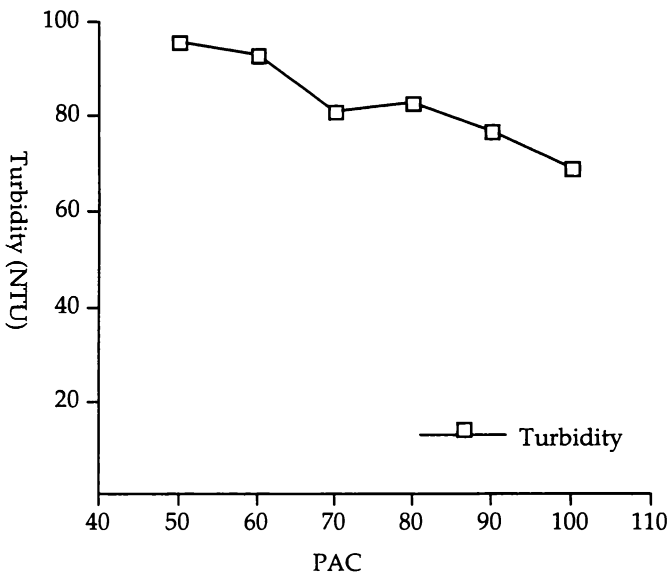


Figure 4-28. Flocculation of PTE (supplied by WFS, Hamilton), using PAC.

In some applications, coagulants are used as filter aids to increase the size of filterable particles (Boller et.al. 1981). The effect of increasing amounts of PFS on effluent particle size was studied. Figure 4-29 shows the results.

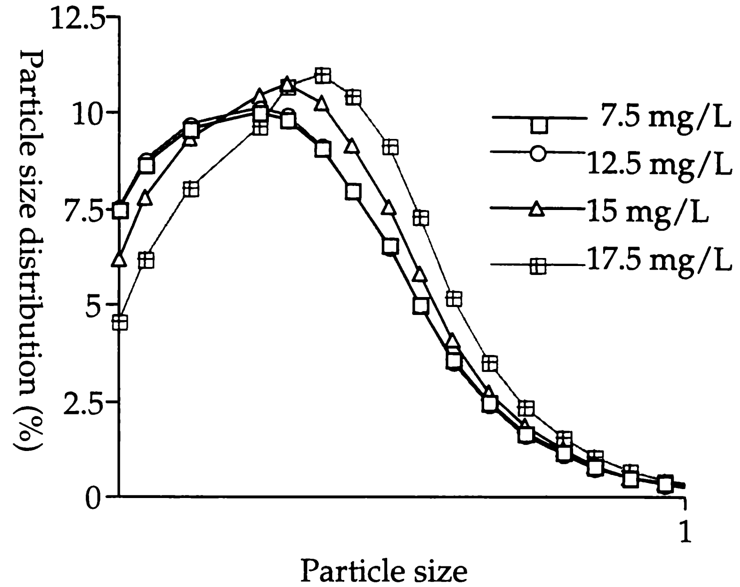


Figure 4-29. Effect of increasing concentration of PFS on particle size distribution (by number) in PTE.

Increasing particle size of PTE prior to filtration was achieved for doses above between 15 mg/L of PFS.

Distribution of the particle size shows a shift in size with flocculant addition at levels much less than the critical coagulation concentration. When using PFS flocculant as a filter aid at a concentration, sufficient to increase particle size, given a daily effluent volume of 4.5 x 10^4 m^3 (HWWTP), costs for chemicals would amount to \$290 for PFS.

4.6.3 Effect of pH on particle size and filtration by TM.

TM has an inherently low IEP, see Section 3.5 and it was of interest to check if electrostatic kinetic forces could be enhanced by changing surface charge. Initially the effect of pH on particle size was measured. Results for PTE at pH 3.7 (HCl) and pH 9.36 (NaOH) are shown below in Figures 4-30.

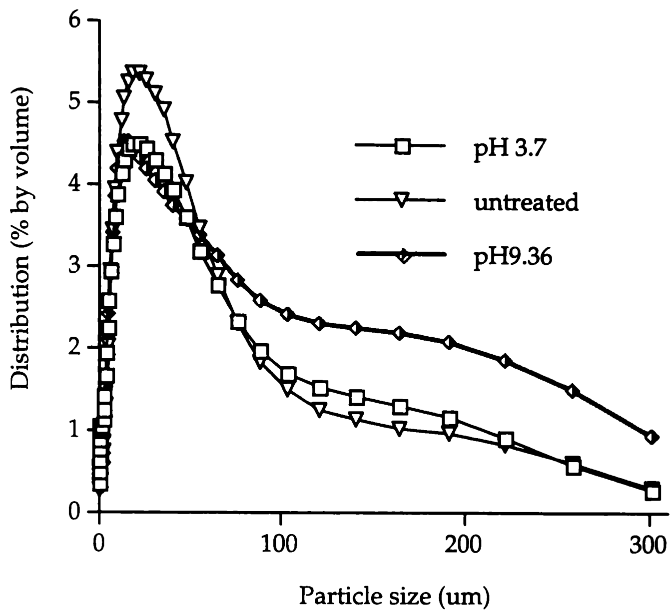


Figure 4-30. Effect of pH on the particle size in PTE.

There was a small increase in volume of particles of size > 100 μm for the high pH sample. The size of the majority of the particles in the acidified pH PTE was not affected.

The pH modified PTE solutions were filtered through a 100 mm deep TM filter. Figure 4-31 shows the flowrate data.

The lowest flow was recorded for the low pH sample. An accelerated blocking of the media was observed. The filter bed blocked after 30 void volumes, whereas filter runs lasted about twice that time for untreated and high pH PTE.

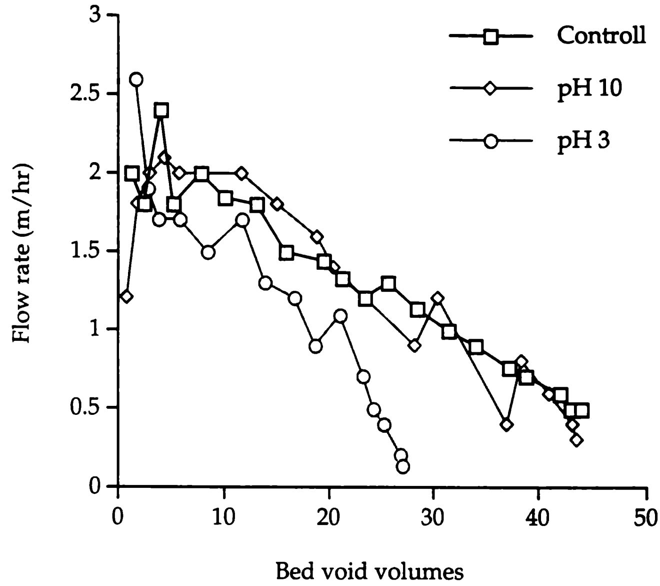


Figure 4-31. Effect of pH on PTE flowrate through 100 mm TM.

The results for turbidity removal are summarised in Figures 4-32 and Figure 4-33.

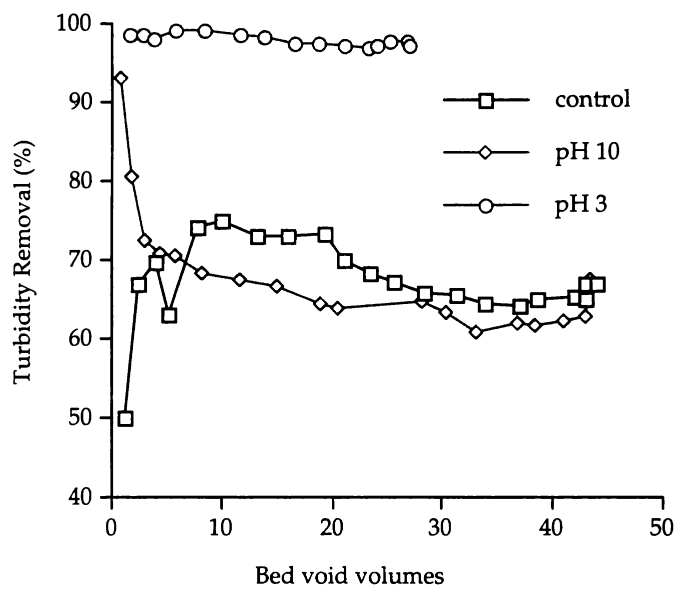


Figure 4-32. Effect of high and low pH on turbidity removal for, filtration ofering PTE through TM.

There was a significant improvement in the turbidity removal for the low pH PTE. Turbidity removal improved to > 90%.

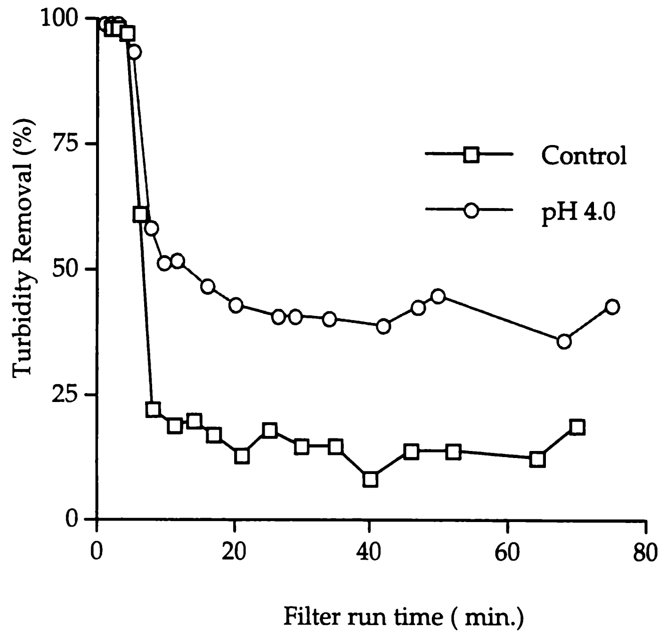


Figure 4-33. Effect of low pH on turbidity removal for filtering FDE through TM.

The effect of low pH enhanced filtration, was tested using FDE. Data are summarised in Figure 4-33 where enhanced turbidity removal is again indicated. The effect of lowering the pH can be explained by electrostatic attraction resulting when the pH is lowered below the 4.1 isoelectric point of TM (see Table 3-5).

4.7 *Summary of results from bench studies*

4.7.1 Filtration of FDE

Using a 100 mm deep TM filter bed achieved turbidity reductions of between 40 - 70%. Flowrate at a constant head between 550 and 650 mm started at between 1.5 - 2.5 m/h and dropped below 0.5 m/h after about 100 bed void volumes were filtered. The filter bed usually blocked after that time. Lower initial flows were recorded by increasing the bed depth but there was no improvement in turbidity removal with deeper beds and they blocked earlier. A bed depth of at least 100 mm is necessary to accommodate the tube filter nozzles designed to be used in conjunction with TM. Additional turbidity removal was obtained by adding a second TM bed in series with the first bed. When the filtrate collected after a first pass through a 100 mm deep TM bed was filtered again through the same filter bed, a further 20% turbidity was removed, but no additional improvement was achieved, following a third pass. Most FDE solids > 5 μm were removed by TM filtration, using a 100 mm deep bed. Additions of 5% TM fines, removed effluent particles > 2 μm were removed. A dual media of TM over silica sand achieved good turbidity removal and allowed faster flowrates. Filtration of FDE through SS and silica sand removed < 20% of the turbidity.

4.7.2 Filtration of PTE

TM removed about 30% of the turbidity from PTE. A bed depth of 100 mm performed best, with early turbidity break-through occurring within the shallower beds and early filter blockage occurring within the deeper beds. The coarse grain SS 7/14 reached its maximum performance at a bed depth of 1000 mm with 30% turbidity removal. The finer grade SS 14/24 performed similarly but increasing turbidity removal was achieved by increasing the bed depth from 1000 mm to 1800 mm. Silica sand showed a steady increase in performance with depth, reaching its maximum of 50% turbidity removal at a bed depth of 1700 mm.

Effluent particle sizes are irregular, the majority of numbers being < 0.5 μm. The majority of larger particles were retained on the surface of the Silicon Sponge bed, the smaller grain size in TM did not achieve significant additional removal, the threshold of particle size was too high.

4.7.3 Cost of flocculants when used as filter aids during PTE separation

Increasing particle size of PTE prior to was achieved by consuming between 10 - 15 mg/L of PFS or PAC. Given a daily effluent volume of 4.5×10^4 m^3 (HWWTP), daily costs for chemicals to reduce PTE turbidity by filtration would range between NZ$290 and NZ$160 respectively.

4.7.4 Pulsed filter bed using TM and FDE

Pulsed backwashing of the filter bed could lead to significant reduction in backwash water consumption and increased solids content in the pulse backwash water. This technology would be feasible where intermittent operation is not a problem, such as in dairy sheds.

4.7.5 Recommendations for pilot plant studies.

The effect of bed depth using SS should be studied further. The larger grade should be used, because it offers the advantage of a higher hydraulic conductivity and potentially longer filter run times, when compared with the finer media. In addition to ordinary backwashing, a pulsed bed backwashing operation should be tested. The effect of a single filter should be compared with two filters in series. Single media filters in combination with fine TM should be tested, possibly using TM as final polishing filter. Flocculation of effluent prior to filtration should be part of the pilot plant studies.

Chapter 5

Nozzle Flows

5.1 *Introduction: Flow characteristics of (TMC) nozzle and nozzle bolt combinations*

The function of filter nozzles and in particular TMC nozzles as part of the filter underdrain system has been covered earlier in sections 1.7.3 and 1.14. These TMC nozzles are installed by adopting the pipe lateral manifold system in conjunction with a false floor, commonly used in New Zealand. The resulting nozzle density comes to about 30 to 60 nozzles/m^2, which is sufficient based on their porous nature. Nozzles are being held onto the floor with a bolt; this is connecting them to the pipe system, see Figure 5-1.

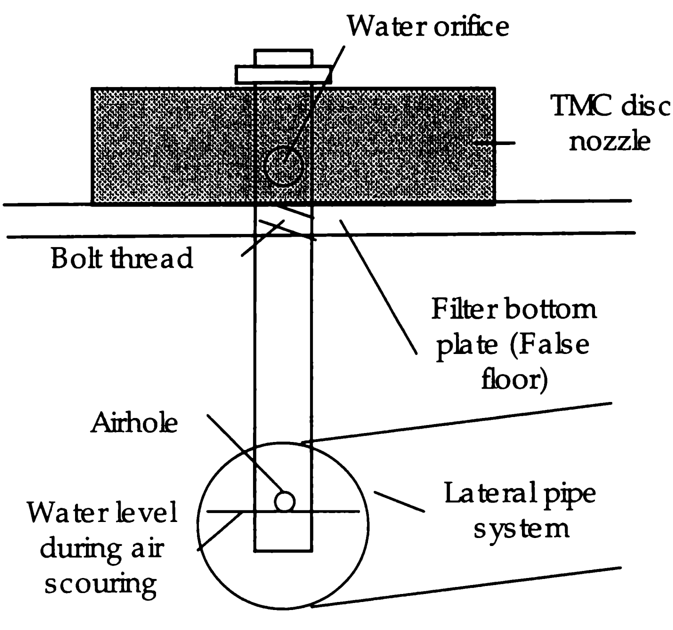


Figure 5-1. TMC nozzle secured to pipe lateral system on a false floor.

Bolt design and distribution is an integral part of a filter system. Sufficient resistance is required to achieve an even distribution of air and water through the underdrain. If resistance is too low, waves in the underdrain (plenum) can cause some bolts to receive intermittent flow because their ends are above the water surface (Geering, 1997). Under ideal conditions, the water surface remains relatively stable, due to the resistance caused by the nozzle system.

Water orifices in the bolt are crucial to ensuring appropriate flows in both directions during the operation of the filters (Lombard and Haarhoff, 1995). Theories exist for calculating the pressure drop along pipe lateral manifold systems (Chaudry and Reis, 1992). For individual nozzle–bolt systems, however, there is no reliable way of theoretically estimating the flow, so this has to be determined experimentally.

A range of TMC nozzles had been designed by WFS to replace the conventional plastic nozzle. Amongst them, a special type of TMC tube nozzle was developed to achieve fluidisation of fine TM sands. It was thus of interest to determine, whether these designs had any impact on flow characteristics. Initial work for conventional down-flow filtration was required to assess the performance of the new nozzles. The high headloss across the tube nozzles was recognised. These nozzles would not be used for other than TM medium. They are however a major source of resistance during backwashing, hence work was invested to establish flow dynamics in order to improve flows in upflow direction.

Further objectives included:

- Comparison between TMC disc and plastic rose nozzles.
- Effect on flow resistance of various nozzle shapes and bolt combinations in up-flow and down-flow.
- Effect on orifice area and orifice position flow resistance.

5.2 *Methods and material*

5.2.1 Testing rig

An experimental rig, see Figure 5-2 and Figure 5-3 was constructed from a 250 mm wide PVC tube with the top and base plates fixed by stainless steel bolts. Water supply was from a header tank. During the experiments, a head of about 3600 mm was available to create flow conditions, representative of actual flowrates between 15 – 60 m/h. For consistency, results for the various configurations were taken at a headloss of 2000 mm across the nozzles, representing a flowrate of about 30m/h in the case of the traditional nozzle-bolt combination.

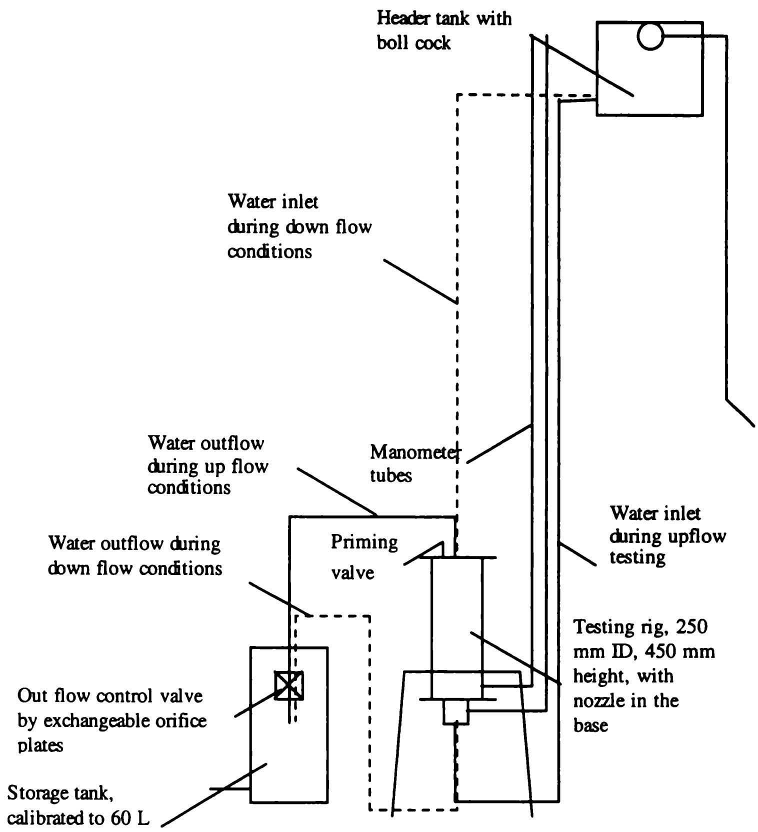


Figure 5-2. Experimental rig for testing of flows in nozzle-bolt combinations.

Flowrate through the rig was controlled at the outlet by inserting appropriate metal orifice plates, see Figure 5-3. Manometer tubes were fitted above and below the nozzle section of the rig to measure the headloss across the nozzle system, creating a range between 0.5 - 4 m.

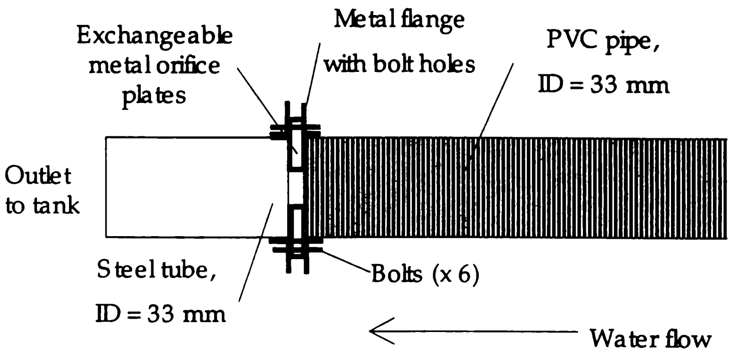


Figure 5-3. Flow control after the experimental rig, using a set of exchangeable orifices (ID range: 4.99, 7.45, 9.97, 14.98 and 25.55 mm).

Flowrate was determined by measuring the time required to fill a 66 L tank. From the plots of flowrate vs headloss, data for flowrate at constant headloss across the nozzle system was obtained by extrapolation.

Details of the nozzle installation inside the rig are shown below in Figure 5-4.

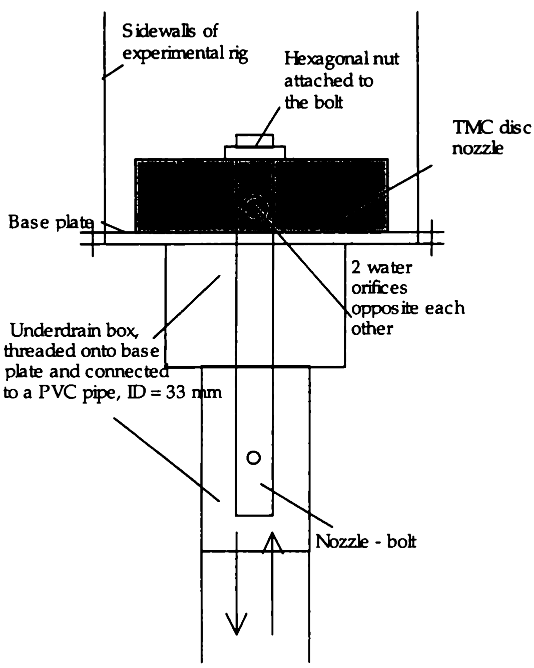


Figure 5-4. Details of filter nozzle inside the experimental rig.

The nozzle was fitted onto the base plate by its nozzle bolt. To swap nozzles, the base of the experimental rig was unscrewed, while the central tube remained supported on a metal-framed stand.

For experiments in up-flow mode, water was supplied from below the base plate of the rig. For down-flow experiments, water entry occurred through the top plate of the rig, it flowed down, passed the nozzle- bolt combination and as in up-flow mode, it left the system through the flow control at the outlet.

5.2.2 Filter nozzle-bolt design

Three nozzle designs were tested. They included TMC disc nozzles of three base thicknesses, a TMC tube nozzle and a plastic rose. These are illustrated in Figure 5-5 below. For comparison, the shape of a conventional plastic rose nozzle currently used throughout New Zealand is also included.

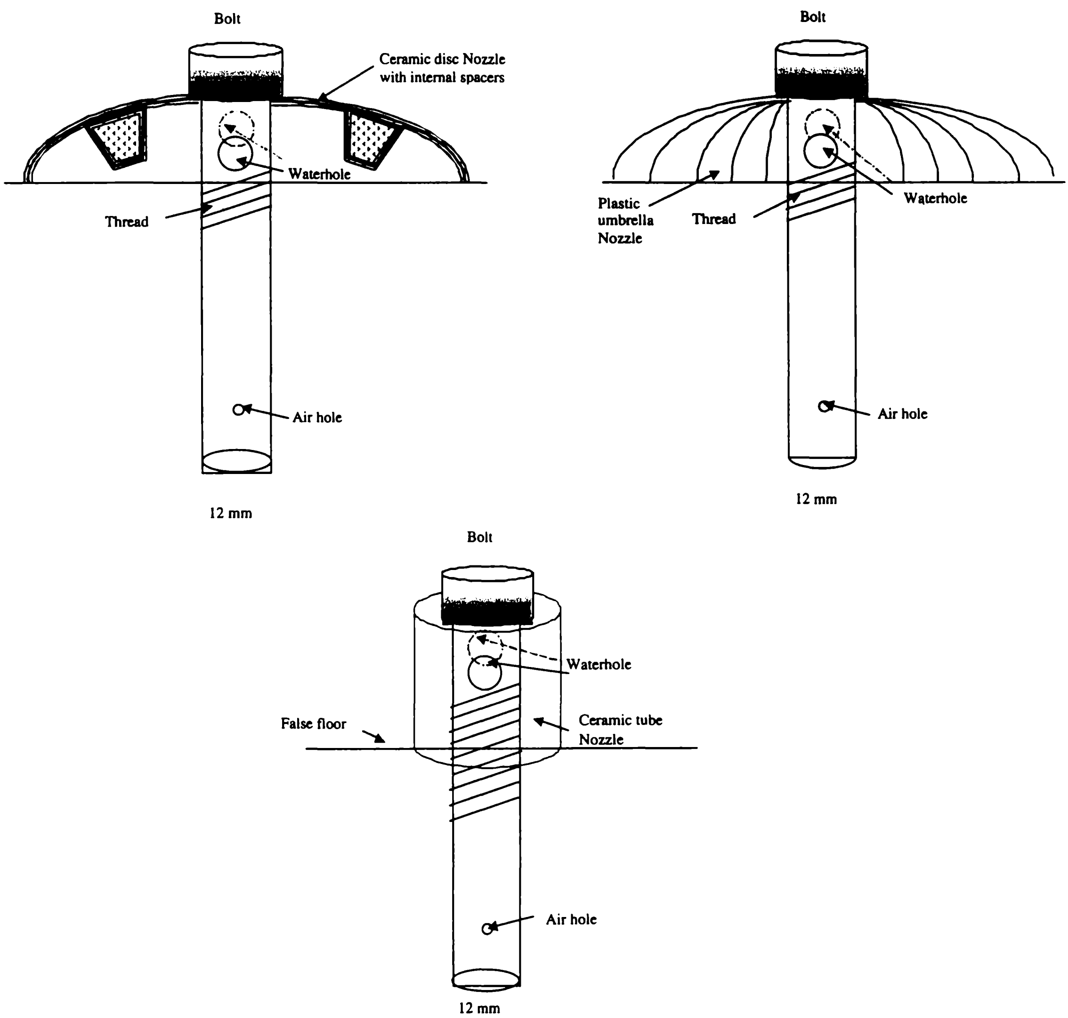
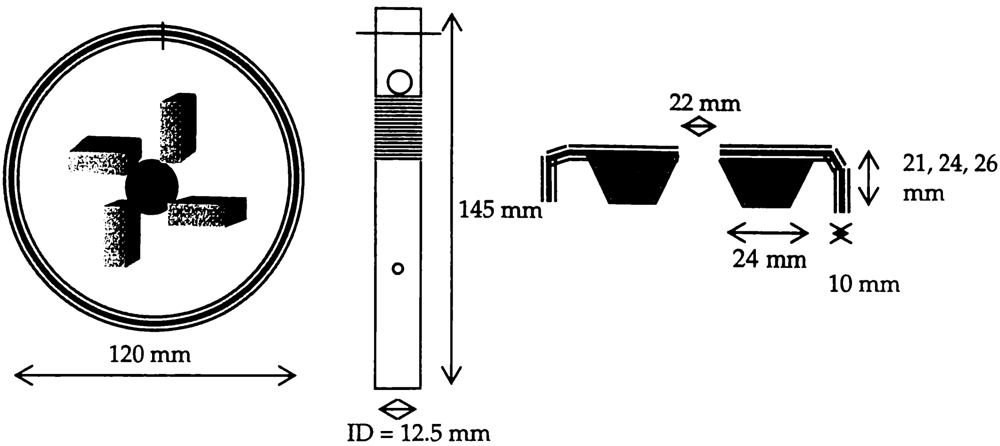


Figure 5-5. Three types of filter nozzles from top left to bottom centre: TMC disc, plastic umbrella, and TMC tube.

Further details of the TMC nozzles and the nozzle bolt are shown in Figures 5-6 and 5-7 and images from light microscopy examinations are shown in Figure 5-8.



TMC disc nozzle with bolt in the centre,
profile and side-view

TMC tube nozzle with bolt in the centre,
profile and side - view

Figure 5-6. Two types of TMC nozzles and a bolt. The bolt fits tightly through the centre of the nozzle and is used to fasten the nozzle-bolt to the under floor.

The two basic shapes of TMC nozzles are: disc nozzles, consisting of a circular porous ceramic plate with an opening in the centre for the bolt and tube nozzles, which are cylindrically shaped porous ceramic sleeves with their exterior surface fabricated with a spiralling corrugation of 2 mm depth. The purpose of the corrugation was to produce a variation of backwash flows to assist fluidisation; they were especially designed for the fine TM medium (Hill, 1996). A recent modification to the disc nozzle (not shown here) resulted in a double nozzle where two disc nozzles of 25 mm thickness each are cemented together at their base, with a longer bolt inserted through the centre hole for fastening.

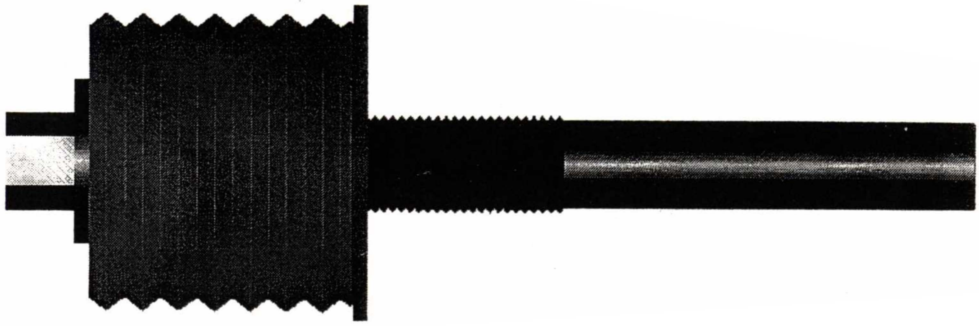


Figure 5-7. TMC tube nozzle and bolt.

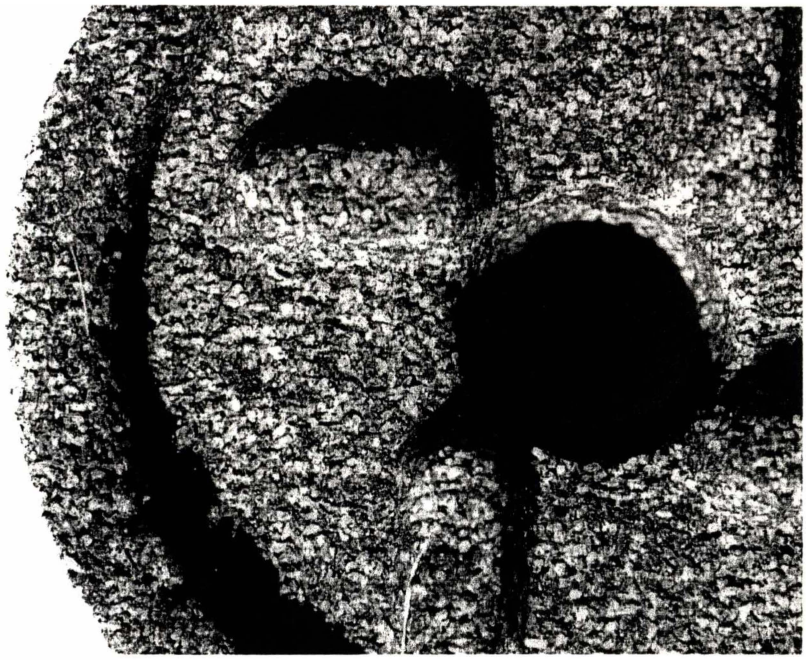


Figure 5-8. TMC disc nozzle, looking up from underneath.

Nozzle bolts are fitted through the centre of the nozzle and are tightened to the under floor by turning the hexagonal nut on top. Standard bolt design features are shown below in Figure 5-9.

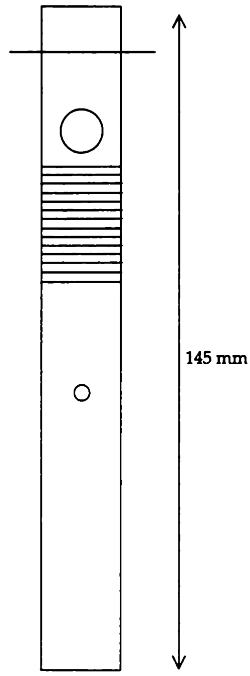


Figure 5-9. Standard bolt details.

The PCDM bolt consists of a 140 mm long PVC tube with an internal diameter (ID) of 12.5 mm, a hexagonal nut shaped on top and an air orifice of 3.5 mm diameter, positioned at 50 mm from the base of the bolt. In addition, two water orifices with ID of 9.5 - 9.8 mm opposite each other are positioned 100 mm from the base of the bolt, just above the threading. The airhole serves as an entry point for air during air scour, during which time the base of the bolt is submerged in water.

Longer bolt-tubes with larger water orifices were used, to accommodate the longer tube and the double disc nozzles. The bolt length was extended at the threaded area, resulting in a total length of 200 mm and the water orifice diameter was enlarged to an average ID of 13 mm.

5.3 *Nozzle-bolt performance characteristics*

5.3.1 Comparison between TMC disc and plastic rose nozzles

A single TMC disc nozzle and a plastic rose were compared. Both were installed with a standard bolt, which was also tested without the nozzle attached. Results are shown in Figure 5-10.

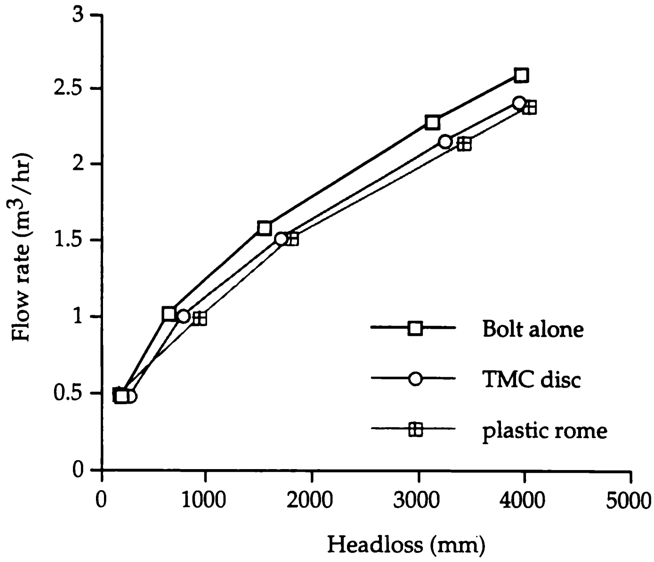


Figure 5-10. Comparison of down-flowrates between a bolt, a plastic and a TMC disc nozzle.

This experiment established that the TMC disc nozzle had flow characteristics very similarly to those of conventional plastic rose nozzles. Neither of them presented significant resistance to flow at rates passed by the conventional bolt system.

5.3.2 Effect of nozzle shape on flowrate

TMC tube and double disc nozzles were tested, both using the longer bolt system. For comparison, the standard TMC single disc nozzle of varying wall thicknesses (with a conventional bolt) was also included. Results in Figure 5-11 show their behaviour in down-flow conditions.

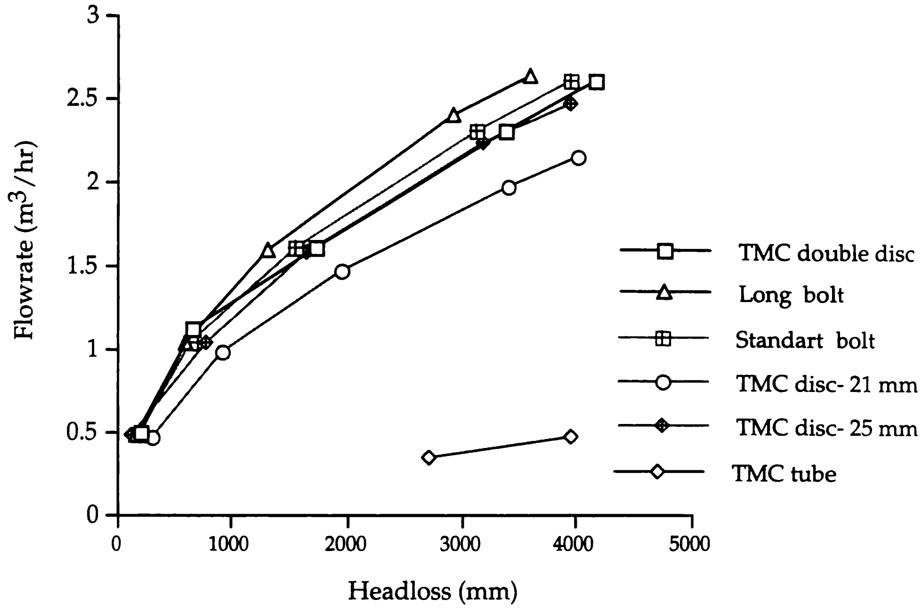


Figure 5-11. Effect of nozzle design on down-flow characteristics.

Different TMC disc nozzle designs did not contribute to significant differences in flow. However, in down-flow the tube nozzle, even when fitted with a high flow bolt showed approximately a 70% reduction in flow.

5.3.3 Comparison between up-flow and down-flowrates

To test the resistance presented by the ceramic nozzle material, a standard TMC disc nozzle-bolt combination and a bolt alone were installed and flows compared in both directions.

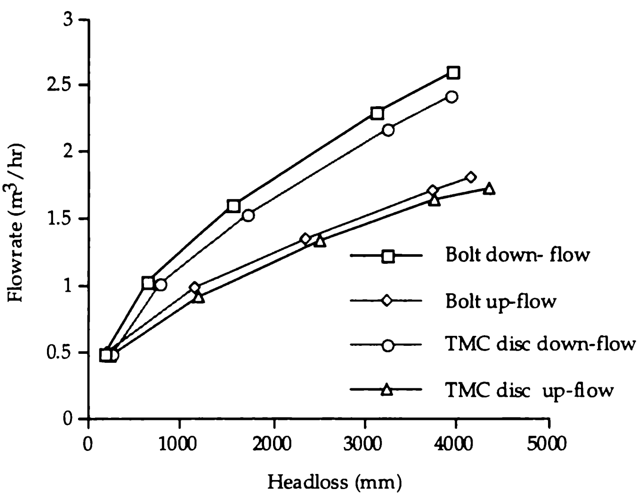


Figure 5-12. Comparison between up-flow and down-flow.

Figure 5-12 shows that the addition of a nozzle in the nozzle-bolt combination has a minimal effect on flows. As for the bolt alone, under 2 m headloss conditions, down-flow was 30% higher than up-flow with the standard disc nozzle attached. The controlling effect was therefore the bolt and its water orifices.

5.3.4 Effect of total water orifice area

Specific experiments were designed to alter water orifices in the bolt. A range of bolts was modified, resulting in a variety of the diameters for the water orifice between 1 and 14 mm each. For the standard bolt, the applicable water orifice diameter is: 9.5 mm and for the large–hole bolt, it is 13 mm. For the experiment, each bolt was installed individually to measure the effect on the flowrate under a range of headloss in both directions.

Figure 5-13 compares up-flow with down-flowrate for each size of combined orifice area by extrapolating the measured flowrate to a common headloss of 2 m. The values from Figure 5-13 were used to calculate velocities (m/sec) and the Reynold's number (Re), which are listed in Table 5-1. Standard deviation ($x\sigma n$) for down-flow at 2 m head was 0.0262 m/hr, as shown in A. 8.

Table 5-1. Mean velocity values and Reynold's number for a series of water orifice hole sizes for flows in both directions with a head of 2 m.

Hole diameter (m)	Down-flow (m/sec)	Up-flow (m/sec)	Re
1.7×10^{-3}	12.1	12.1	2×10^{4}
2.86×10^{-3}	8.6	8.6	2.5×10^{4}
4.87×10^{-3}	7.2	8.6	3.5×10^{4} / 4.2×10^{4}
5.98×10^{-3}	6.9	7.8	4.1×10^{4} / 4.7×10^{4}
6.9×10^{-3}	7.1	7.6	4.9×10^{4} / 5.2×10^{4}
7.98×10^{-3}	6.1	6.1	4.8×10^{4}
8.9×10^{-3}A′	7.4	5.6	6.6×10^{4}/ 5×10^{4}
9.6×10^{-3}	6.9	4.8	6.6×10^{4}/ 4.6×10^{4}
9.96×10^{-3}	6.8	4.4	6.8×10^{4} / 4.4×10^{4}
11.3×10^{-3}	5.4	4.94	6.1×10^{4}/ 5.6×10^{4}

Note: Where up – flow rate and down – flow rate vary, two Re were generated. The first number refers to down – flow and the second to up – flow.

The relatively high Reynold's number in both directions, indicates turbulent flow in all the tested configurations.

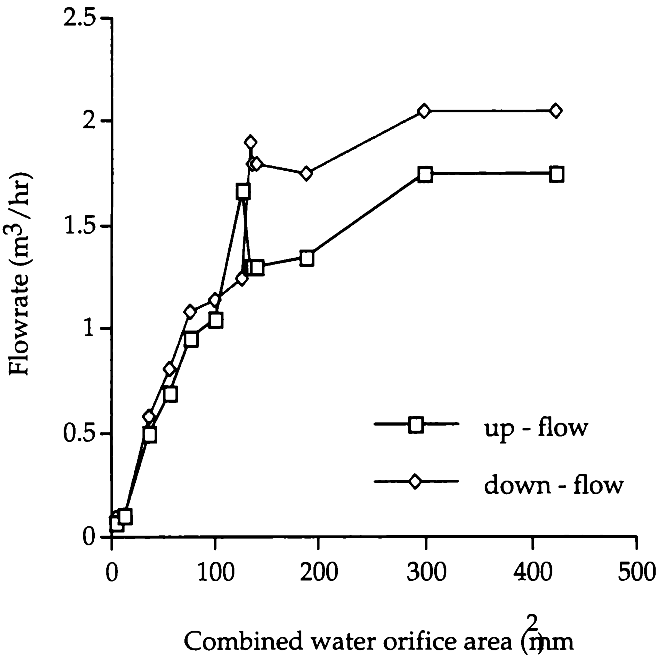


Figure 5-13. Effect of water orifice area in bolts on flowrates in both directions at 2 m headloss. Area of both water orifices combined is plotted.

At small orifice areas of less than 120 mm^2, down-flow rates were similar to up-flowrates. With larger areas between 120 - 140 mm^2, a reduction in up-flow of up to 30% occurred. When the orifice area increased further from 145 mm^2 to 300 mm^2, the earlier observed restriction in up-flow decreased from 30% to 15%. Standard bolts with an orifice area of about 145 mm^2, are therefore most affected by increased resistance in up-flow. This may be associated with a greater degree of turbulence in down-flow direction, indicated also by the high Reynold's number.

5.3.5 Flow characteristics of an open ended tube

In order to further establish the source of hydrodynamic restrictions in the bolt, additional modifications to the standard bolt were carried out. The bolt was modified in three ways: (i) nut was sawn off, (ii) water-orifices sealed, (iii) tip of the bolt head filled with cement, down to the opening of the water orifices. Results of the volume rate of flow (m^3/h) are shown in Table 5-2 and in Figure 5-14.

Table 5-2. Bolt design on directional volumetric flowrates at 2 m headloss.

Bolt type	Water orifice area (m^2)	up m^3/sec (Re)	Down m^3/sec (Re)
Standard bolt	1.45×10^{-4}	3.47×10^{-4}	5×10^{-4}
Nut-less bolt, orifices blocked	1.23×10^{-4}	4.78×10^{-4}	5.8×10^{-4}
Nut-less bolt, orifices open	1.23×10^{-4}	5.3×10^{-4}	5.7×10^{-10}
Standard bolt, dead-space filled	1.45×10^{-4}	3.3×10^{-4}	5×10^{-4}

This data shows that a 30% reduction occurs in up-flow when compared with down-flow. Flows through tube like systems (eg nut-less bolt) encountered less resistance in both directions. However the reduced flow in up-flow remained as is shown in Figure 5-14.

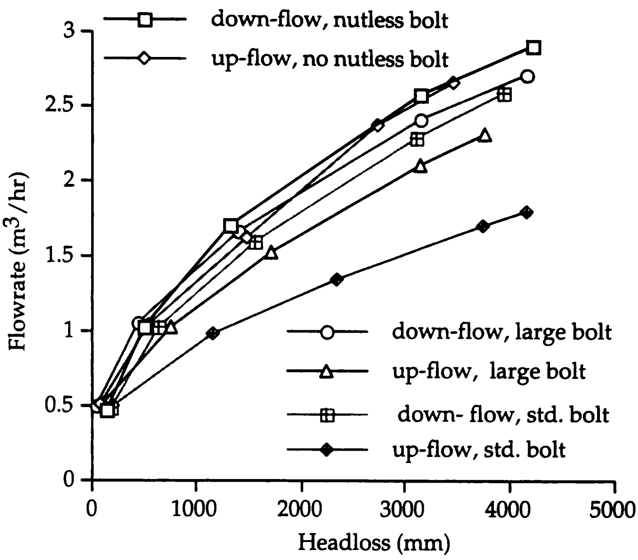


Figure 5-14. Effect of bolt modification on directional flowrates at various headlosses.

For the purpose of defining the hydrodynamic conditions, the Re was calculated from the mean velocity value (m/sec) through the water orifices. In the case of the nut-less bolt, the orifice area of the tube was chosen as the regulating flow area. Table 5-3 lists the results.

Table 5-3. Mean velocity (m/sec) and Reynold's number (Re) for various bolt designs.

Bolt type	Water orifice area (m^2)	up m^3/sec (Re)	down m^3/sec (Re)
Standard bolt	7.24×10^{-5}	4.8 (Re: 5.8×10^4)	6.9 (Re: 6.6×10^4)
Nut-less bolt, orifices blocked	1.23×10^{-4}	3.9 (Re: 5×10^4)	4.7 (Re: 5.8×10^4)
Nut-less bolt, orifices open	1.23×10^{-4}	4.3 (Re: 5.3×10^4)	4.6 (Re: 5.8×10^4)
Standard bolt, dead-space filled	7.24×10^{-5}	4.6 (Re: 4.4×10^4)	6.9 (Re: 6.3×10^4)

Note: In order to calculate the Reynold's number (Re) for the standard bolt and for the dead space filled bolt, d_o of the water orifice was chosen. For the nut less bolt systems, d = the bolt area.

Similar trends to the measured values shown in Figure 5-12 emerged; in general showing much reduced flow in up-flow. The nut-less tube with the water orifices in-tact (eg. open holes) registered the highest values in both directions and the smallest effect on flow direction, whereas flows through the tube-like bolt were restricted in up–flow.

Following modifications to the bolt, involving filling up of the dead-space above the water orifice holes, up-flowrate remained lower when compared with down-flowrate at the same headloss. The high Re number in both directions indicates turbulent flow in all situations.

5.3.6 Effect of number, size and distribution of the water orifices

Effect of multiple small holes

Multiple water orifices were drilled to give a combined water orifice area of the standard bolt (140 mm^2). Subsequent D_o ranged between 5.5 mm and 7.7 mm (see Table 5-4). The appropriate number (3 – 6) of these holes were arranged evenly spaced across the top section of the bolt. The effect of the distribution of multiple water orifices on the flowrate through smaller holes is shown in Figure 5-15.

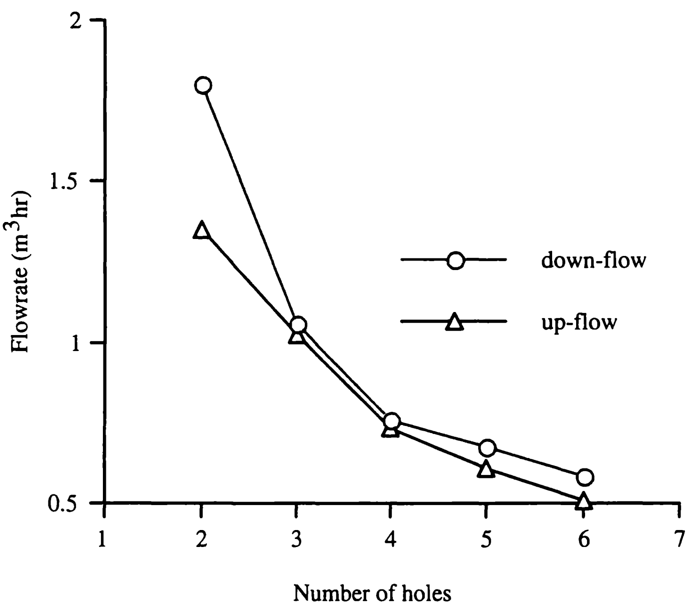


Figure 5-15. Effect on the flowrate after distribution of water flows through smaller water orifices in both directions (headloss: 2 m).

Figure 5-15 shows that as water orifice numbers increase from two orifices to three orifices, the resistance in the system also increased, total flows decreased and directional effects disappeared. Results from the multiple water orifice-bolt modification experiments, including Re are summarised below in Table 5-4.

Table 5-4. Summary of flows through multiple orifices, sizes and arrangements

Number of orifices	D_o (m)	up m^3/sec (Re)	down m^3/sec (Re)
2	9.6×10^{-3}	4.9 (4.7×10^4)	6.9 (6.6×10^4)
3	7.8×10^{-3}	8.4 (6.5×10^4)	8.7 (6.8×10^4)
4	6.8×10^{-3}	10.8 (7.4×10^4)	11.1 (7.5×10^4)
5	6.1×10^{-3}	14.3 (8.7×10^4)	15.7 (9.5×10^4)
6	5.5×10^{-3}	17.8 (9.8×10^4)	20 (1.1×10^5)

In all situations Re remained high, indicating turbulent conditions. As D_o decreased, Re increased.

The effect of hole position.

In the case of the three-orifice system, the positioning of holes was modified so that the orifices were either symmetrically or asymmetrically arranged. Resulting flow data are shown in Figure 5-16.

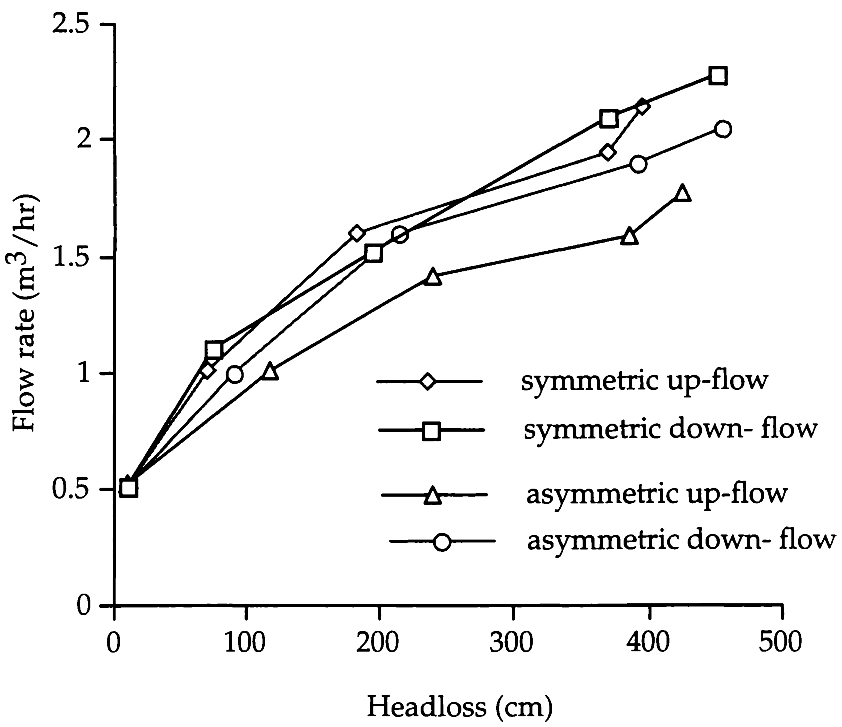


Figure 5-16. Comparison between a symmetric and an asymmetric arrangement of the water orifices.

The results show that up-flows with asymmetric arrangement of the water orifices were significantly lower than down-flows throughout the range of experimental conditions.

5.4 *Hydrodynamic considerations*

5.4.1 Flow through orifices

Consideration of Reynold's numbers for the various experiments described in Section 5.3 indicated that flow in both directions was turbulent. A consequence of turbulent flow through an orifice is the *vena contracta* effect (Munson et.al. 1998). The *vena contracta* effect occurs where streamlines continue to converge beyond an orifice until they become parallel, following converging. If the exit is not a smooth, well-contoured nozzle, but rather a flat plate, the diameter of the jet (sum of all the streamlines) A_j will be less than the diameter of the orifice hole A_o. The phenomenon is effectively a result of the inability of a fluid to follow the contours of sharp corners (see Figure 5-17). The area of *vena contracta* A_a is the section of the smallest area A_j after the orifice plate or after the entrance to the bolt.

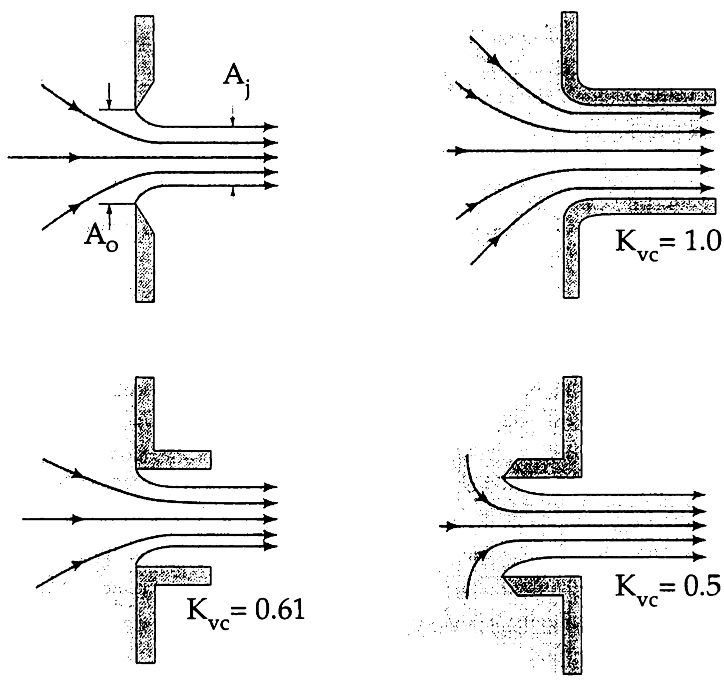


Figure 5-17. Schematics of vena contracta. A *vena contracta* region is often developed at the entrance of a pipe.

The contraction coefficient (K_{vc}) is defined as (A_j / A_O).

Well made, smooth orifices produce free jets with a K_{vc} value near 1. Where K_{vc} is large, the actual flowrate is large. The ratio of the contraction depends on the incoming flowrate, the thickness of the tube, and the smoothness of the flow, prior to the orifice plate. Small holes and smaller flows lead to a smaller K_{vc}.

A possible explanation of the hydrodynamic behaviour during flows through nozzle-bolts is drawn below in Figure 5-18.

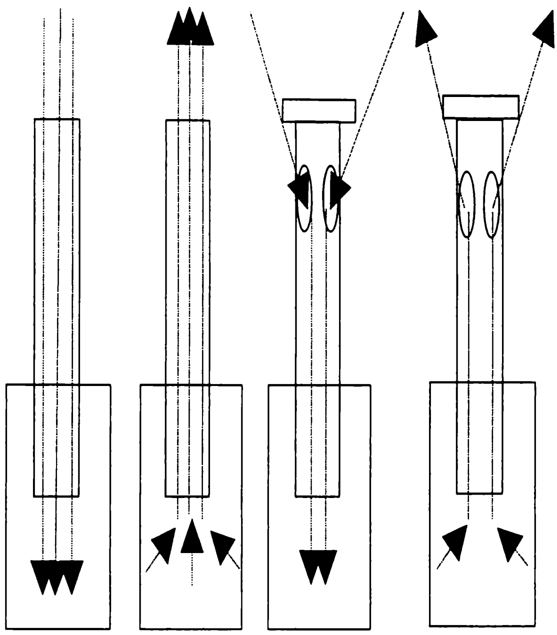


Figure 5-18. Model of streamlines. Open ended tube (nut-less bolt) on the left and standard bolt on the right.

If the effect of *vena contracta* is applied to flows in nozzle-bolt combinations, the streamlines bend around a sharp, 90^0 corner in up-flow. During down-flow, a similar situation occurs but the angle for the bend of streamlines will be smoother.

5.4.2 Experiment to test *vena contracta* effect at the base of the bolt

According to the model shown in Figure 5-18 for a standard nozzle-bolt, *the vena contracta* effect is experienced twice in up-flow and only once in down-flow conditions. In situations where a nut-less bolt is used (Table 5-3), it is assumed that the effect of *vena contracta* only occurs in up-flow conditions.

Consequently, an experiment was designed to test the effect of improving the hydrodynamics at the base of the bolt. A cone was attached to the base of the bolt to smoothen the streamlines. The testing unit is shown below in Figure 5-19.

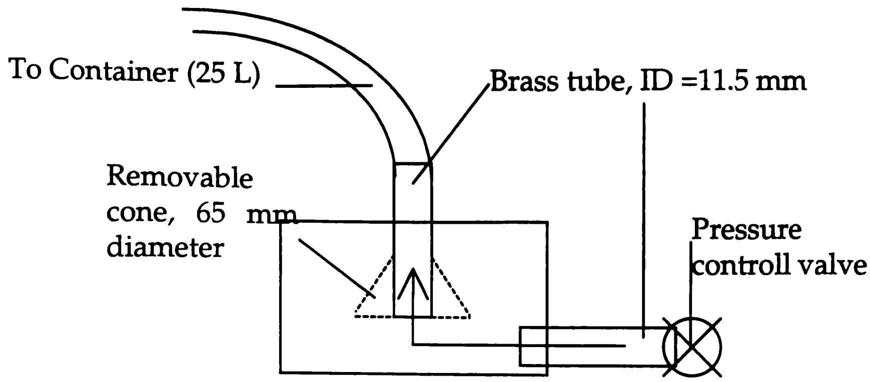


Figure 5-19. Testing unit to measure the effect of a cone on up-flow rate.

The testing unit was constructed from square PVC sheets, each 5 mm thick with 150 mm sides. A brass tube was inserted from the top and used to model the behaviour of a nut-less bolt. Flows in both directions were controlled at a pressure of 2 m across the pressure-regulating valve. A removable cone was attached at the bottom of the bolt, to allow comparison of flows with and without the cone.

Table 5-5. Effect of a cone on up-flow rates for a bolt.

Flow with cone (m^3/hr)	Flow without cone (m^3/hr)	% Increased flow with cone
0.856 (n=10)	0.832 (n=10)	0.024 % ± 0.013

A smooth entry provided by the cone shaped bottom end of the tube led to a slightly higher up-flowrate when compared with the straight tube. The increase was not large, being only about twice the standard deviation of the measurement. It appears that for bolts operated under conditions similar to those investigated in the experiment described, there is not much advantage to be had from improvements in the hydrodynamics at the base of the bolt.

More work is required to assess the potential for reducing resistance through the nozzle bolt system in up-flow conditions at high flowrates.

5.5 *Discussion and conclusions*

There was no difference in flow conditions when TMC disc nozzles were compared with the plastic rose. Both nozzle types performed well and did not add significant resistance to water flow.

The shape of the TMC disc nozzles did not affect the flowrate significantly. The small decrease in flow observed with a thinner disc could have been caused by lower hydraulic conductivity of the denser ceramic material used in this construction.

The TMC tube nozzle showed approximately a 70% reduction in flow. The greater resistance of the tube nozzle was probably due to thicker walls and finer material used in forming the ceramic.

A difference in up-flow and down-flowrates was observed. This difference and the total flow were sensitive to water orifice size and position in the bolt. There was no effect below a combined water orifice area of 120 mm^2. At larger sizes including those used for the standard bolt, up-flow rates were about 30% less than down-flow flowrates at a 2 m head. At larger orifice areas, the difference between up-flow and downflow reduced to 14%. When comparing Re, larger values lead to less resistance through the water orifice (see Table 5-1).

Increasing the water orifice area above the size used in the standard bolt, did not significantly increase the down-flowrate, indicating the standard bolt design showed the optimum hole size for down-flow with smaller holes showing a reduced flow.

When three water orifices of the same combined orifice area as the two-orifice system were symmetrically positioned around the bolt, up-flow rate increased by about 14%.

Resistance through the nozzle-bolt system might be controlled by the *vena contracta* effect. A slight increase in up-flowrate was observed when the base of a bolt was modified to create a cone.

Given the crucial role even backwash water distribution in granular media filtration plays, this study provided additional understanding of the flow dynamics in a nozzle-bolt combination. Reduced resistance to flow during backwashing can be achieved by re-designing the bolt.

Chapter 6

Pilot Plant Evaluation Trials

6.1 *Introduction*

Two basic effluent types were studied, primary treated effluent (PTE) and secondary treated effluent (STE). The motivation of the majority of the work was to test the feasibility of using porous ceramic dual media (PCDM) filtration using SS, silica sand and TM media on PTE as an extension to the existing primary treatment. These experiments were based at the Hamilton Wastewater Treatment Station (HWWTS). For completeness, the PCDM system was also tested in a tertiary role on STE with and without flocculation. Experiments with STE were located at the Morrinsville Wastewater Treatment Station (MWWTS).

L. Choo carried out initial work on pilot plants with SS and TM media under contract to WFS. The same applies to pilot plant trials on secondary treated effluent carried out by A. White, as mentioned in Appendix A9. Data gathered during both pilot plant trials were later evaluated as part of this study.

During the work with PTE, the investigations of coarse SS at bed depths >1000 mm followed from conclusions based on work described in section 4.7. SS offers the advantages of long filter runs, due to its large solid holding capacity and high hydraulic conductivity. Its surface characteristics further contribute to effective filtration. In addition to traditional backwashing methods between filter runs, the short-pulsed bed backwashing system was also of interest on a SS filter bed.

Investigations of TM followed because it, like SS is an indigenous material, but has a high density, small grain size and magnetic properties, which may lead to novel applications in filtration. TM was included either as a polishing filter in series with two conventional PCDM filters and also separately as a single medium filter where the effect of pH adjustment of the effluent was tested. As part of the under drain in the PCDM evaluation plants, TM ceramic (TMC) tube nozzles were installed, because they retained the fine TM medium yet permitted its effective fluidisation.

The experimental sections that follow will summarise the results and give an outline of trends observed. Some of the data from field trials are listed in the Appendix in Tables A-6 and A-7.

6.2 *Materials and Methods*

6.2.1 Hamilton Wastewater Treatment Station

The HWWTS serves a population of approximately 110,000 people, including a small number of industries (see Section 2.8.1). Average daily flows of 4.5 x 10^4 m^3 were recorded with a medium TSS content of 350 mg/L. During the existing primary treatment process, contaminants and grit from the wastewater are removed firstly through bar screens, followed by a grit aeration tank and finally by three sedimentation tanks, reducing the solid content to about 85 mg/L TSS. Solids removed by these processes are currently compressed and concentrated by anaerobic digestion tanks from where the resulting sludge goes through a filter press. It is then composted on site. Liquid from the press is discharged back to the plant together with the raw sewage. After the sedimentation tanks, the supernatant is being chlorinated and sulphur dioxide injected, to neutralise the excessive chlorine, before discharge in to the Waikato River. A schematics representation of this process is shown in Appendix 12. During the HWWTS studies, evaluation units were installed to filter primary treated effluent, following the settling tanks.

6.2.2 Morrinsville Wastewater Treatment Station

The (MWWTS) treats the sewage from 5,500 inhabitants and some major industries. Average daily flows range from 4.4 x 10^3 to 1.1 x 10^4 m^3 (see Section 2.8.3). Treatment systems comprise a sequenced batch reactor (SBR), which operates with a cycle of biological aeration and settling. The SBR-process is followed by decanting and by further settling in a balance tank. At the time of this study, the secondary treated wastewater was chlorinated and discharged into the Piako River directly after chlorination. During the MWWTS studies on STE, an evaluation unit was installed after secondary treatment, following the balance tanks.

6.2.3 Evaluation units for pilot trials

Evaluation units, incorporating the essential features of the PCDM technology (Langdon et.al. 1996 and Section 1.14) were constructed for pilot plant trials. These

6.2.3 Evaluation units for pilot trials

Evaluation units, incorporating the essential features of the PCDM technology (Langdon et.al. 1996 and section 1.14) were constructed for pilot plant trials. These units needed to be small enough to manage frequent changes of the media but large enough to mimic real conditions. Glass walls were fitted to allow visual monitoring of the behaviour of the media during filtration and backwashing. Four evaluation unit designs (conventional 1.0 m, modified 1.0 m, 2.0 m and 2.6 m) were used. They are shown below in Figures 6-1, 6-2 and 6-3.

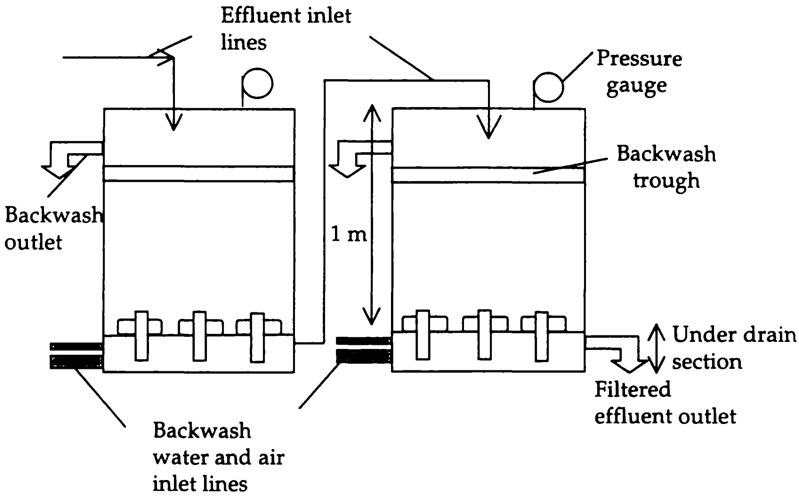


Figure 6-1. Conventional 1.0 m PCDM evaluation unit with TMC disc nozzles; two units connected in series.

The base dimensions for each unit shown in Figure 6-1 were 500 mm by 380 mm.

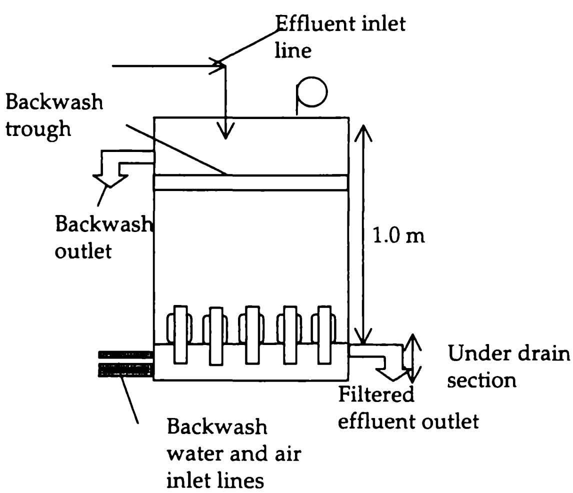


Figure 6-2. Modified 1.0 m PCDM evaluation unit with TMC tube nozzles for TM trials.

The base dimensions for each unit shown in Figure 6-2 and Figure 6-3 were 340 mm by 420 mm.

The short, conventional PCDM evaluation units (Figure 6-1 and 6-2) were closed filters, where the effluent was pumped to the system.

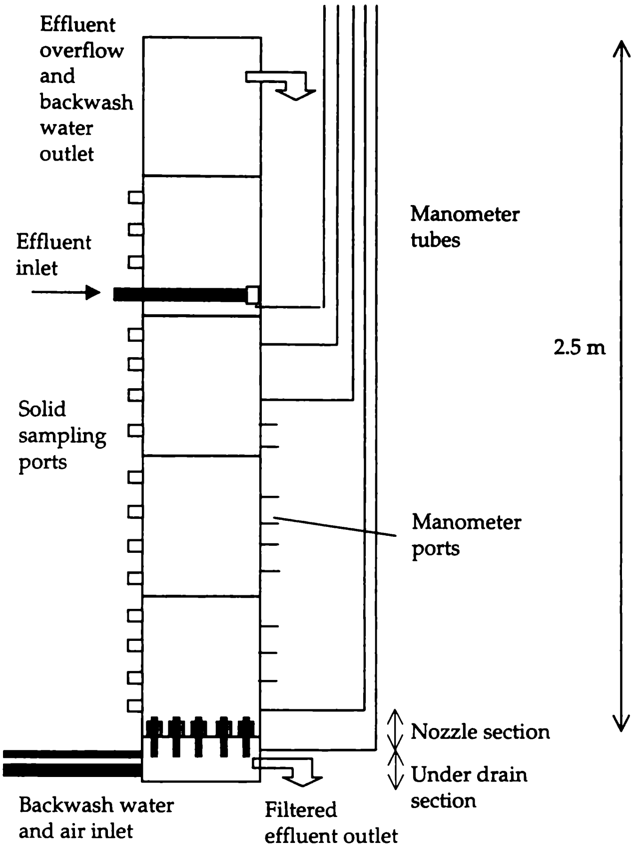


Figure 6-3. Tall PCDM evaluation unit for PTE bed depth effects, currently showing inlet pipe and manometer tubes setting for the 1500 mm deep bed.

The tall evaluation unit was uncovered and a constant head was maintained with an effluent overflow near the top (Figure 6-3). The effluent inlet line for the tall unit was a 40 mm ID galvanised pipe, capped at the end and perforated with 21 distribution holes (ID = 4 mm), facing up.

Nozzle shape and spacing was altered between the units. The conventional 1.0 m PCDM unit contained disc nozzles spaced at 40 nozzles/m^2. Nozzles were replaced with TMC tube nozzles in the modified 1.0 m PCDM unit, and remained in the same way for the tall PCDM unit, spaced at 70 nozzles/m^2. Tube nozzles were used to accommodate the fine TM media. Later, when the tall unit was used for STE trials, tube nozzles were replaced with TMC double nozzles, spaced at 42 nozzles/m^2. More details of the nozzle-bolt combinations used were shown earlier in Section 5.1 and Section 5.2.

Under drain depth was 160 mm for all the units. Piping for the effluent lines were 40 mm diameter galvanised pipes. The same size was used for the backwash water inlet and filtered effluent outlet lines. Inlet hose for pressurised air was 15 mm ID.

The tall unit, described in Figure 6-3 and used for PTE and STE trials was equipped with manometer tubes. These consisted of clear PVC tubing (ID = 10 mm), tacked onto a pole alongside the unit, over a graduated measuring tape. A valve was attached near the base of each tube, close to the unit wall. They were closed during filter shut down or backwashing periods, to decrease the chance of development for air locks inside the tubes. Metal mesh was used at the manometer wall opening, to contain the media. This prevented filter media from potentially clogging up the manometer tubes. On the opposite sidewall, ports for solid media sampling were installed. These comprised short, galvanised threaded pipes (ID = 40 mm) of 50 mm length with a screw cap. Spacing of both, the manometer tubes and the solid sampling ports was at a rate of four per 500 mm section. Additional manometer tubes were installed, immediately above and below the nozzle system, to provide data for the nozzle section.

6.2.4 Operation of the units:

During the first and the second trials with the shorter evaluation units (Figure 6-1 and Figure 6-2), PTE was pumped to the filter unit through a pipe in the top. The pump setting remained constant during a run. In normal runs, valves at the outlet(s) of the unit(s) were adjusted to give initial flowrates of approximately 8 m/h. Disturbance of the media surface was minimised by directing the inflow stream into the backwash trough suspended across the centre at the top of the filter. A complete filter run with SS media lasted between 45 - 50 h and much less for the TM trials. Sampling was carried out between the hours of 8 AM and 6 PM at regular time intervals. Backwashing, carried out between replicate runs, is outlined below. Access for replacement of media for different media configurations was through the top.

During the pilot plant tests, the flowrate was not controlled but was monitored regularly during the run by hand. The rig did not allow the increased heads necessary for maintaining constant high flow. This resulted in the declining filtration rate operation. In preliminary experiments where flowrates at the early stage of a run were varied, little effect on filtration efficiency was observed (see section 6.5.3).

For PTE work carried out with the taller units (see Figure 6-3), effluent distribution occurred through a perforated inlet pipe installed directly across the centre of the unit.

The height of this inlet pipe was adjusted when the bed depth was changed. Effluent overflow on the top edge of the unit provided a constant head of 2.7 m between the effluent overflow and the open outlet of the filtered effluent line. Filter flowrate was limited by a combination of headloss across the filter medium and the resistance to flow through the TMC nozzles. The high headloss across the tube nozzles was recognised. These nozzles would not be used for other than TM medium. For the experiments of this study, the restriction presented by the nozzles did not affect the experimental setup, they were serving as headloss control. A complete trial lasted between 25 h and 52 h. Sampling occurred between the peak times of 10.00 AM and 3.00 PM, when effluent characteristics were reasonably constant. At the end of a run, the filter media was sluiced out through the solid sampling ports. Between replicate runs, media was backwashed by using a combination of raw river water and pressured air. Details of the backwashing procedures are explained below.

For the final trials, carried out at the MWWTS, effluent was pumped to the tall unit and flowrate was controlled by hand with an outlet valve. The same effluent distribution system as was described for PTE in Figure 6-3 was used for the STE trials. Flocculants were administered with a separate dosing pump (Milton Roy, LM1 p146-151, from Foster & Associates) installed in-line, prior to the filter unit. Media depth changes for each trial as achieved by adding medium through the top. Backwash water supply at the MWWTS was clean drinking water. Backflow contamination was prevented by manual operation. Alternatively a backflow prevention valve could be installed.

6.2.5 Bed backwashing

Backwashing procedures and sources varied for the four different trials described.

During the shallow bed PTE trials, backwashing of the filters was with clean water from the drinking water supply while care was taken to prevent cross contamination to the town supply. A flow rate of 20 m/h for 1 min was followed by air for 30 sec (37 - 40 psi) and by a final water rinse until clean backwash water (< 10 NTU) flowed out through the backwash water outlet, which occurred after about 10 min.

During the second PTE trials with the taller unit, untreated river water was used for backwashing and pulsing, because there was no drinking water supply available with sufficient pressure to fluidise a deep bed. For the majority of the work, backwashing between successive runs consisted of a 4 min water wash (60 m/h) followed by a 1 min air scour, again 3 min water wash and a 30 sec air scour. The last wash cycle was repeated. Backwashing during pilot plant trials was not optimised. It was simply part of the trial to clean the media. Backwash water consumption was measured manually by using the flowrate at the backwash water outlet, (Figure 6-3) multiplied by the time required until turbidity levels in the backwash water overflow dropped below 10 NTU. Air-scouring was carried out on a bed with the backwash water drained to a level of about 100 mm above the media surface. The resulting backwash water consumption used during the trials was about 10 m^3/m^2.

During the final, STE trials, media was backwashed for 5 min at 20% - 30% bed expansion. This was followed by chemical cleaning, with a 0.08% Sodium hypochlorite solution, which was pumped together with the backwash water through the media at 30% expansion for another 5 min. The filter bed was then rinsed with backwash water at 20% bed expansion for 15 min.

Bed conditioning in addition to backwashing

During the TM trials, media conditioning consisted of filling the filter unit with PTE through the inlet, while the outflow valve remained in closed position. In three separate experiments, the pH of PTE was adjusted to 4.0, 5.2 and 8.9 with HCl or NaOH respectively. After a settling period of 15 min, the effluent was drained. Turbidity levels of the effluent were taken before and after the filter. The process of filling, pH adjustment and draining was repeated until the filter run was terminated, by the filter bed blocking.

Backwash pulsing

Backwash pulsing was carried out, following draining of the filter bed to a residual unfiltered effluent level of 100 mm above the medium surface. A 10 sec water pulse at 60 m/h, resulted in the trapped air from the under drain to dispersing throughout the bed together with the 25 L of backwash water supplied. The resulting backwash water sludge was released through the side-ports, just above the top level of the drained filter bed.

6.2.6 Performance measures

The main performance measure for all PTE trials was turbidity removal, expressed as percentage. This was determined as described in section 2.11. Generally turbidity correlates with TSS values with the relationship applicable to this study shown in Figure 2-1. During this project, the poor correlation between turbidity and TSS was realised, especially in the lower turbidity range, applicable to secondary treated effluent as is shown in Figure 2-1. Where it was critical, during pilot plant studies using secondary treated effluent, samples were also tested for TSS and BOD removal.

Additional performance measures included flowrate, headloss and solids distribution. Flowrate was measured by timing the filling of a 10 L container after the filter unit and headloss was measured from the water level in appropriately spaced manometer tubes placed along the bed depth.

Solids distribution in a drained filter bed was measured, using the method outlined in Section 2.7.1. After the effluent solids were extracted, the media was dried at 105^0 C to a constant weight. Solids were determined, using the calibration curve for TSS shown in Figure 2-1 and normalised to represent g of dry solids/g of dry SS.

For additional information, some of the samples were tested for trace metals, for Phosphorus and for BOD by Hill Laboratories in Hamilton, New Zealand, who also carried out some of the TSS analysis during this study.

6.2.7 Duration and timing of trials:

An initial 14 runs were carried out between the months of February and May 1995, using SS (14/24) and silica sand as filter media, and the units shown in Figure 6-1. A further 16 runs were carried out between the months of June and August 1995 with the single evaluation unit shown in Figure 6-2 filled with TM.

Deep bed trials, were carried out between May and August 1998 using the tall evaluation unit (Figure 6-3). The final trials with STE were carried out during January 1999. A list with of all trials is shown in Appendix 9 and a research plan in section 2.5.

6.3 Shallow bed PCDM trials of filtration of primary treated effluent

6.3.1 Introduction

Initial trials compared inlet turbidity with outlet turbidity after a single pass through SS (7/14). Further trials investigated the effect of a second filter connected in series with the first. The effect of altering the flowrate was also measured. Mono and dual media, including SS, silica sand or TM were compared. The effect of reduced backwash times was recorded. The conventional 1.0 m PCDM unit, shown in Figure 6-1 was used for these trials. TM trials compared acid with caustic treatment of the effluent as well as the effect of TM bed depth. The modified PCDM unit shown in Figure 6-2 was used for these trials.

6.3.2 Filtration by a mono medium bed of SS

In an initial experiment, a single unit with a 640 mm bed of SS was used to filter PTE. Results are summarised in Figure 6-4.

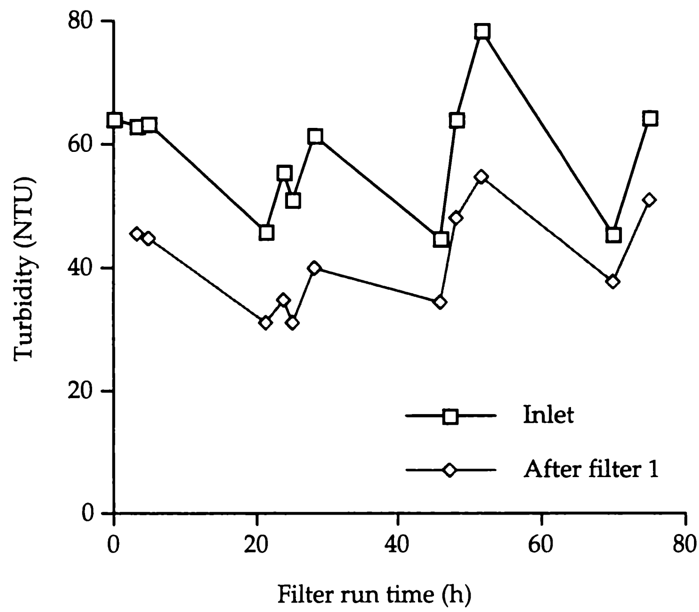


Figure 6-4. Turbidity levels of PTE before and after passage through 640 mm Silicon Sponge (7/14).

As indicated by the turbidity measurements immediately before filtration, the water quality of the incoming effluent varied according to a daily cycle with a maximum of up to 100 NTU occurring for approximately 4 h at about noon and a minimum of about 50 NTU in the early morning. After passage through the bed of SS, turbidity was reduced by about 30% but remained variable.

6.3.3 Filtration by single medium, dual bed systems

The effect of passing the filtered effluent through a second SS bed was tested by placing a second evaluation unit with 450 mm SS in series with the first unit containing 640 mm SS. Results of 5 runs are summarised below in Figure 6-5. The variability of turbidity removal is listed in the Appendix 10. Standard deviations varied quite widely particularly towards the end of a filter run.

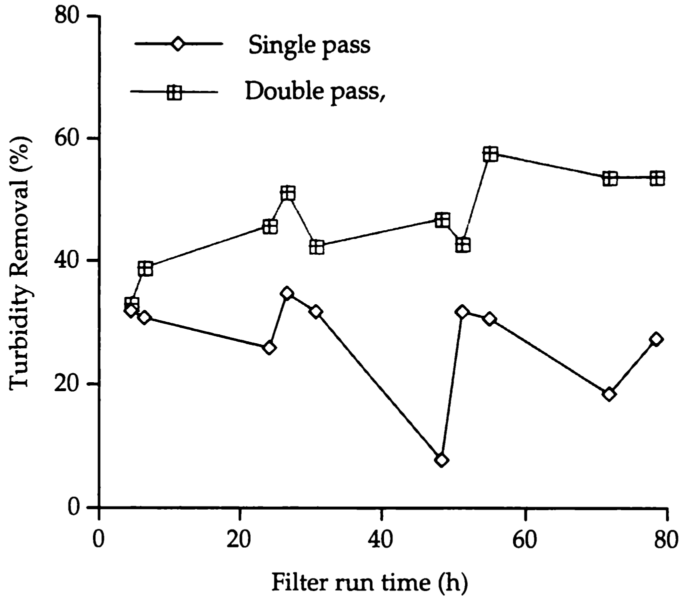


Figure 6-5. Filtration efficiency for filtration with SS (7/14). Data shown include single pass with 640 mm in one unit and a double pass with 450 mm in the second unit in series with the first unit.

In all runs, a time delay of 3 h to 5 h was evident, before the first SS bed achieved its maximum effectiveness. The flow rate decreased relatively linearly from 8 m/h to 4 m/h after 60 h run time. Two runs were continued for 80 h. The flowrates continued to decrease over the extended run time and two erratic and high turbidity data points were recorded, indicating the possibility of turbidity break through after that time.

It is clear that the passage through the second SS bed has resulted in a further reduction of turbidity. Again, a lag time of several hours was observed before the second filter became fully effective. Further small improvements in efficiency occurred for the remainder of the run. After about 60 h, the filtration efficiency of the two units in series had risen to approximately 60%.

6.3.4 The effect of dual media, dual bed systems

The effectiveness of a dual media filter as the second unit was tested. As previously, 640 mm SS was used in the first unit and combinations of fine and coarse SS were used in the second unit. Results are summarised in Figure 6-6 below.

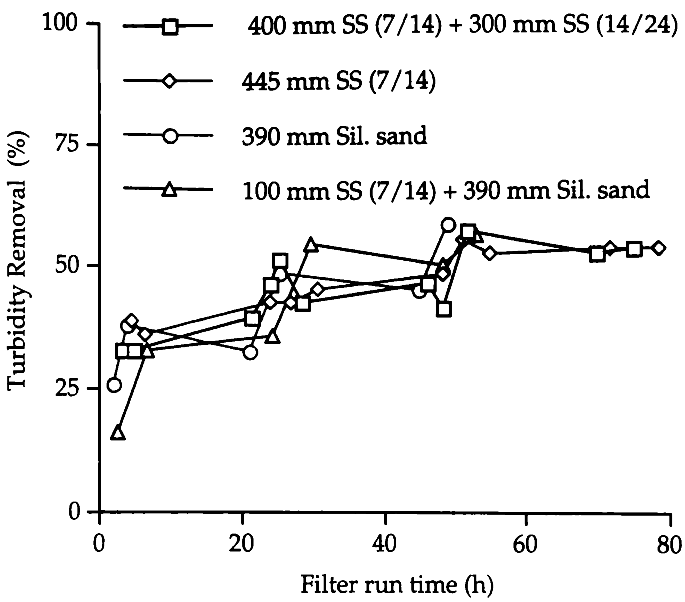


Figure 6- 6. Effect on filter performance when changing media in the second filter. Legend shows media configurations in the second filter unit with a constant 640 mm SS (7/14) medium in the first unit.

Considering the variability of the experiment, all media configurations in the second unit gave essentially the same performance in combination with the SS (7/14) in the first unit.

6.3.5 Effect of flowrate on turbidity removal

The configurations of these trials were such, that at a constant pump setting some increase in pressure and reduction of the flowrate occurred as runs progressed and media became partially blocked. Initial flowrates were similar, starting at 8 m/h and reducing to 4 m/h with run times of about 40 h. The effect of flowrate on turbidity removal was tested, by controlling flow with a globe valve at the outlet (samples were taken after approximately 15 min equilibration time). This resulted in regulated initial flows of 8.0 m/h, 6.0 m/h, 5.4 m/h and 4.7 m/h. Results from these experiments were compared with the data for normal initial flow and are shown in Figure 6-7.

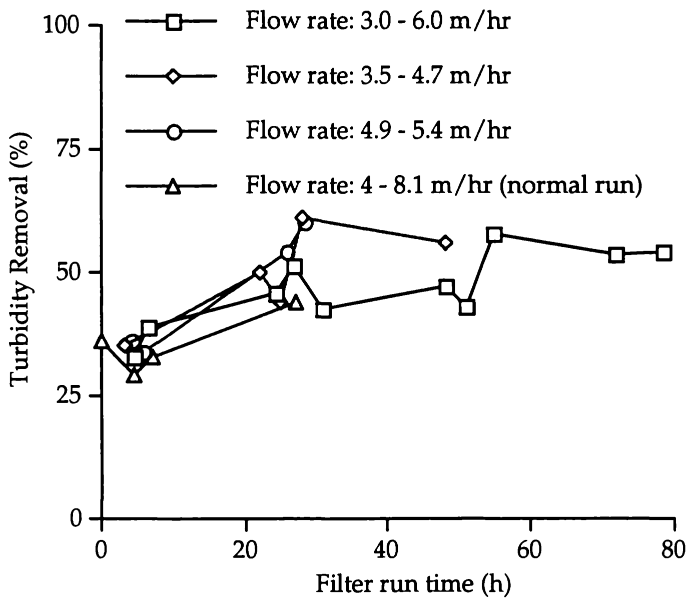


Figure 6 - 7. Effect of different flowrates, adjusting the flow in the beginning of a run and allow for natural reduction during the run. Filter media in the first unit: 640 mm SS 7/14 and in the second unit: 450 mm SS 7/14.

Filter media in the first unit: 640 mm SS 7/14 and in the second unit: 450 mm SS 7/14. When the flowrate in the beginning of a filter run was reduced from 8m/h to 4.7 m/h, turbidity removal fluctuated between 30 - 60% but improved on most data points by about 5 - 20% at lower flows.

6.3.6 Effect of reduced backwashing on filtration efficiency

Usually, filters were backwashed between runs, as described in Section 6.2.5. Reduction of backwash time by leaving the media in a partially blocked state may lead to subsequent improved filtration efficiency. In the experiment shown here, backwashing was limited to 4 min, after which time the backwash water was still noticeably loaded with solids. Results of a normal run, and two successive cycles with short backwashing are shown in Figure 6-8.

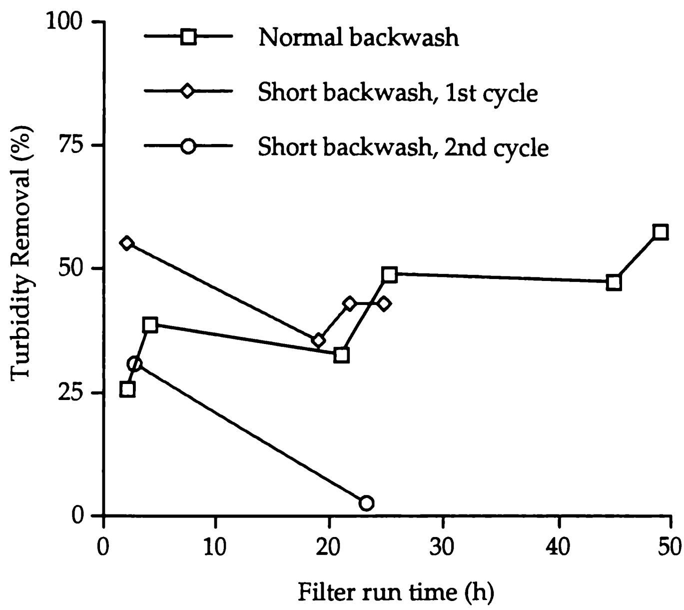


Figure 6 - 8. Effect of reduced backwashing time on filtration efficiency. Filter medium in the first filter unit: 640 mm SS 7/14 and in the second unit: 390 mm silica sand.

While a single short backwashing may have reduced the problem of initial delay time, two successive short backwash cycles appear to leave the media in a condition that turbidity breakthrough occurred within 20 h of the second filtration cycle. During the current trials filters have been left running over extended periods, without the formation of mudballs. In the case of SS drinking water filters, operation for several years has not provided evidence for mudball formation, even when flocculants were used (Liu, 1997).

6.3.7 Usefulness of titanomagnetite as a filter medium

In preliminary experiments the modified evaluation unit (see Figure 6-2) containing fine TM medium was placed as a third unit in series with the conventional 1.0 m PCDM evaluation double unit. The experiment was continued for 16 h. Results from the turbidity data showed that over that time, there was no evidence of improved filtration efficiency from using TM as a final polishing fine filter.

In order to test the effectiveness of TM alone in a single medium system, the modified unit was connected directly to the PTE supply.

Filter efficiency data for bed depths ranging from 90 mm to 190 mm and for acid (pH 4.2) and alkali (pH 8.9) adjustment of the effluent are summarised in Table 6-1.

Table 6-1. Filtration performance by a single medium, shallow bed TM filter, including modifications to effluent pH.

Medium depth (mm)	Initial flow (m/h)	NTU Inlet	Turbidity Removal (%)
90	2.9	31	38
90	2.2	64	41
140	2.5	50	44
140	1.4	43	44
140	1.4	22	50
140 R	1.4	47	53
140	1.3	43	35
190	1.2	28	35
190	1.1	56	41
190	1.1	24	62
190	1.3	28	65
190	1.1	40	70
100 I	1.1	38	50
100 AC	1.1	17	41
100 AC	1	44	48
100 AL	0.9	49	47
100 AC	1	49	75

R = reduced backwashing time; I = increased backwashing time; AC = acid treatment; AL = alkali treatment

Turbidity removals of between 35% and 75% were obtained. Surprisingly, there was no apparent effect of bed depth ranging from 90 mm to 190 mm. Adjustments of the effluent pH to acid conditions improved filtration efficiency in one of the experiments, but had no effect in a second experiment. The acid pH chosen was about or slightly above the isoelectric point of TM so the absence of a reliable effect is understandable. At a constant pump setting, similar to the previous SS trials, initial flowrates ranged from 1.1 m/h to 2.9 m/h and were significantly lower than those for the SS media and the SS/sand combinations. Rapid blocking of the filter resulted in relatively short run times, lasting from 20 min to 2 h.

6.3.8 Summary of outcome from shallow bed trials

The combined depth of SS beds in series reported here was 1100 mm and this combination resulted in a turbidity reduction of 60%. Of the options tested, the performance slightly improved by reducing flowrates from 8 m/h to 4 m/h independent of media type.

It is unlikely that TM will find application in conventional filtration of PTE. The flowrates and run times produced when using TM, are too low. However, pulsing technologies that make use of its high filtration efficiency and overcome early blockage show some promise. A pH value < IEP may have occurred in the last trial where turbidity removal improved from 40% to 75% upon acidification.

TMC nozzles performed well in the filtration of PTE. No reduction in initial flowrate after backwashing was observed during the 14 filtration trials, a combined filtration time of 540 h and a filtered volume of 470 m^3.

6.4 *Deep bed PCDM trials of filtration of primary treated effluent*

6.4.1 Introduction

The results from the shallow bed trials were sufficiently promising to warrant further studies. Bench-scale studies showed improved to filtration efficiency with deeper beds. Accordingly, a second trial was carried out between April and October 1998. The objectives of this work were principally to determine the effects of bed depth of SS (7/14) on filtration efficiency, the distribution of the retained solids, headloss variation through the bed and the effectiveness of pulsed backwashing.

6.4.2 Effect of bed depth on turbidity removal at a constant headloss.

The turbidity removal by various SS bed depths (190, 300, 500, 600, 1070, 1200, 1500, 1600 and 2100 mm) was compared, following the filtration of five bed void volumes of effluent. New medium was used each time the bed depth was changed. For repeat runs using the same bed depths, backwashing between runs as described in Section 6.2.5 was used. In an attempt to ensure the media had been similarly conditioned, turbidity measurements were taken after five bed void volumes of effluent had passed through the filter. Results on the effect of bed depth are summarised in Figure 6-9.

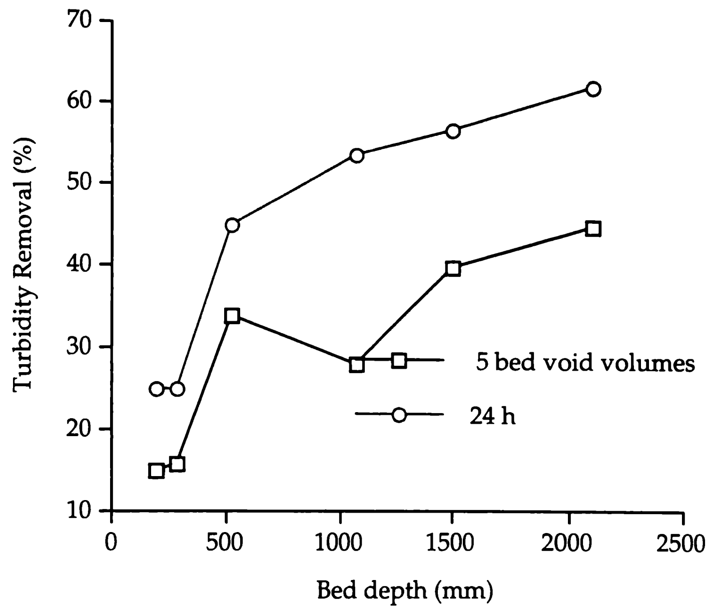


Figure 6 - 9. The effect of bed depth on turbidity removal.

Performance after 5 bed void volumes is compared with a 24 h run. Five bed void volumes represent filter run times of between 20 min and 1 h, depending on the filter bed depth. From data shown in Figure 6-9, it is apparent that filtration efficiency increases steeply with bed depth to a depth of 500 mm. Beyond this depth, improvement of filtration efficiency with depth was much less pronounced.

The detailed effect of run time on turbidity removal at various depths is shown in Figure 6 –10 and Figure 6-11 below.

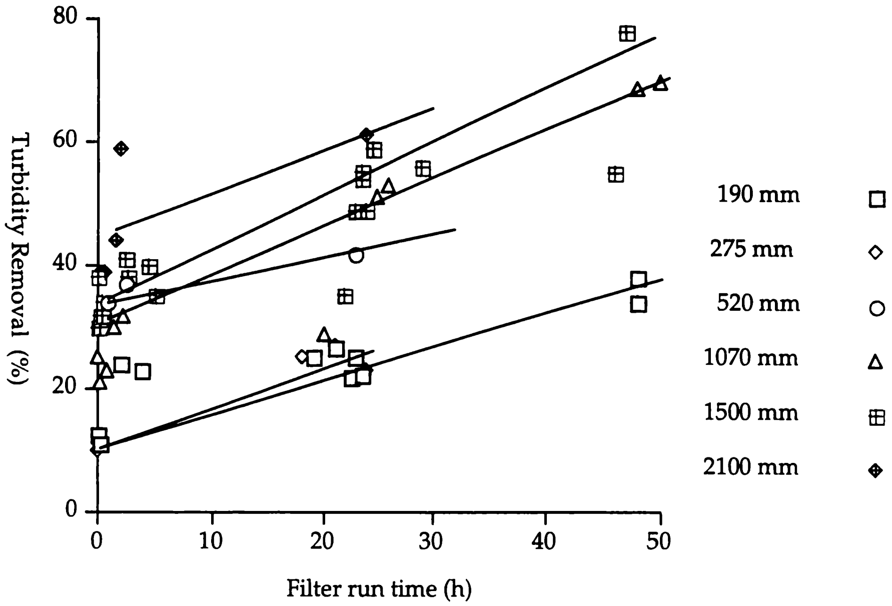


Figure 6-10. Effect of bed at short times.

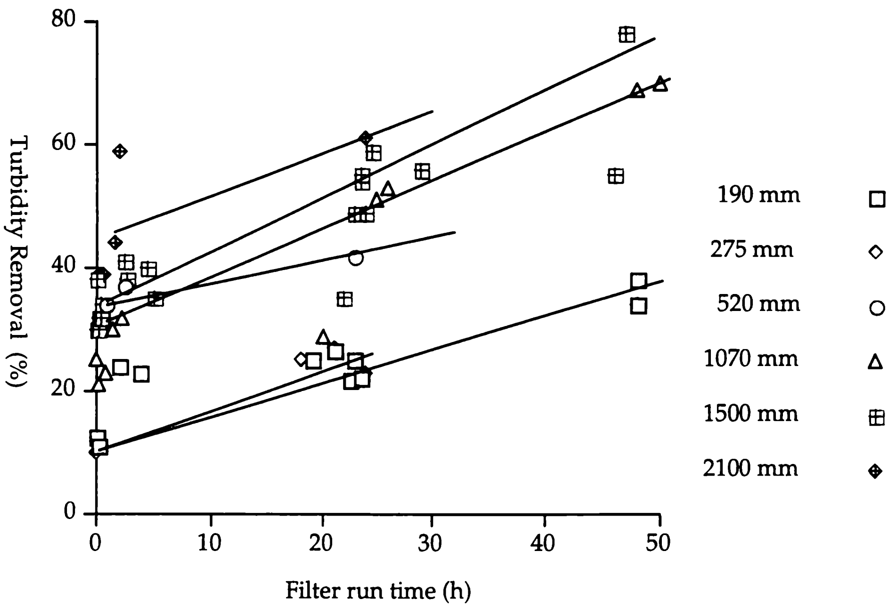


Figure 6-11. Effect of bed depth over entire run. The initial effect of bed depth is maintained over the entire filter run.

Note: significant increase of turbidity removed during the first hour and less pronounced increase over the reminder of the run. Over the first hour of the run, the filtration efficiency increased steeply, reaching approximately 40% for the 2100 mm deep bed. Further but less pronounced increases in filtration efficiency occurred over the remainder of the run, reaching up to 60% efficiency after 50 h for the deeper beds (in case of the 1500 mm deep bed, points at 20 h and 50 h appear anomalously low and were not used in averaging the results).

Ripening occurs across the entire bed depth over time and for each bed depth an increase in turbidity removal was found.

6.4.3 Effect of bed depth on flowrate at constant headloss

For most of the runs, a flowrate of 8 - 18 m/h was established, following a backwashing cycle. The effect of run time and bed depth on the flowrate is shown in Figure 6-12.

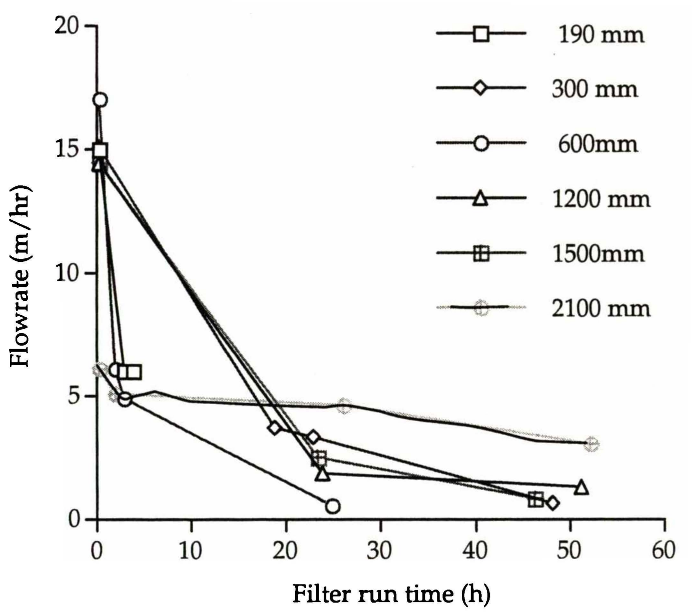


Figure 6-12. Effect of bed depth and run time on flowrates filtering at a constant head of 2700 mm.

At a given headloss of about 2.7 m, the flowrates generally dropped to about 3 m/h after 20 h filter run time, regardless of the initial flowrate. The steep initial drop in flow was followed by a less pronounced decrease after about 20 h. Towards the end of the trial period, flowrates dropped to 2 - 5 m/h.

6.4.4 Effect of bed depth on turbidity removal at controlled flowrate.

As in the earlier experiment described in Section 6.3.5, for shallow beds, the effect of flowrate for a range of bed depths was tested. Samples were taken after approximately 15 min run time after flow adjustment. Flow rate was controlled manually.

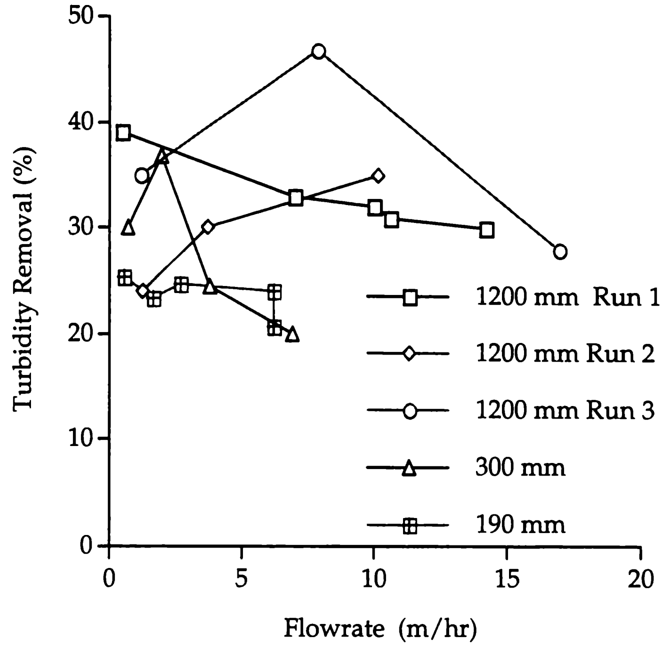


Figure 6-13. Effect of manual adjustment to the flowrate during the first 3 h of a filter run, legend showing bed depths.

By changing the flowrate during a run from 15 m/h to 1 m/h, the turbidity removal fluctuated between 25 - 50% for the deeper beds. For the shallow beds, turbidity removal did not exceed 40%. The results were variable and inconclusive.

However if the same reduction of flow is caused by the build up of a dirt-layer (Schmutzdecke), the effect on filter performance was greater and turbidity removal improved from 20% to as much as 60% (Figure 6-11). On this basis and also supported by earlier results obtained for shallow bed trials, no attempt was made to adjust flowrates during a run.

6.4.5 Chemical analysis

In order to get some information about the effect of filtration on non suspended particles, a chemical analysis was performed. Heavy metals such as arsenic and cadmium are likely to be bound to particles and give an indication of the filterable material in the effluent. The sample was collected before and after filtration, when the filter performed at about 54% turbidity removal, during a deep bed experiment after 50 h run time with a 2100 mm deep filter. Data are summarised in Table 6-2.

Table 6-2. List of components and the change after filtration by a 2100 mm deep SS filter bed.

Component (g/m^3)	Detection limit g/m^3	Effluent In	Effluent out
Total Phosphorus	0.004	8.83	6.98
$CBOD_5$	1	126	86
Total Arsenic	0.002	0.004	0.005
Total Cadmium	0.0001	0.00017	0.00007
Total Lead	0.0002	0.0036	0.0012

Heavy metals such as Arsenic and Cadmium are likely to be bound to particles and are filterable. The large content of unfilterable CBOD probably comprises soluble organics, such as lipids, polysaccharides, proteins and nucleic acids. In most cases the levels of contaminants have been reduced. The data indicate that a large fraction of $CBOD_5$ and inorganic constituents are present as potentially filterable suspended solids.

6.4.6 Solids distribution with bed depth and run time

After completion of 24 h and 50 h runs, using four different bed depths, samples of the filter media were taken from appropriate depths and analysed for solid contents. Data are summarised in Figure 6-14.

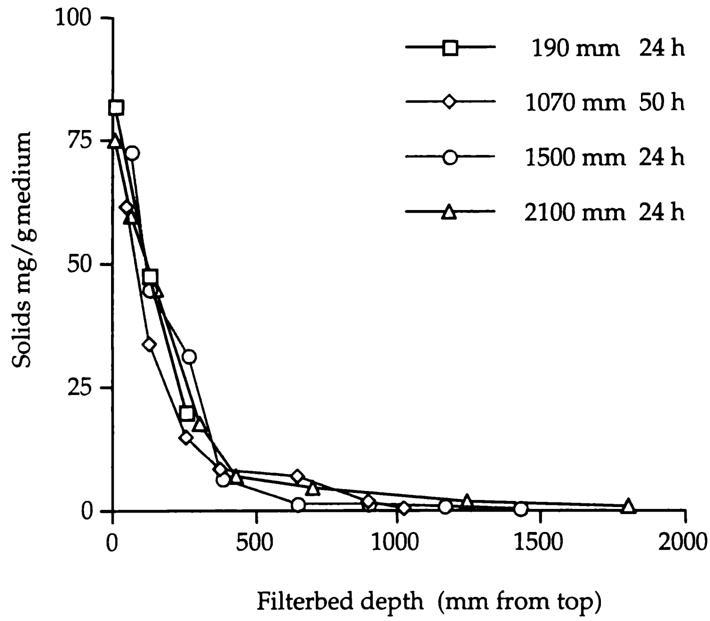


Figure 6-14. Distribution of solids, sampled at different heights from used filter beds. Legend shows the total bed depth of the original filter.

Most of the solids accumulate in the top 300 mm of a SS filter. Depth of the filter had little effect on solids distribution. Only small amounts of solids penetrated deeper than 300 mm and there were virtually no retained solids detected at levels below 1 m in the deeper beds. The poor filtration performance of the 190 mm bed described above can thus be attributed to solids break through. Conversely there does not appear to be any significant advantage in using beds deeper than 1 m under the conditions used in the current work.

6.4.7 The effect of depth and run time on headloss

During a filter run, using four different bed depths, manometer readings were taken from appropriate levels to determine headloss. Typical headloss data for various run times and depths are summarised in Figure 6-15.

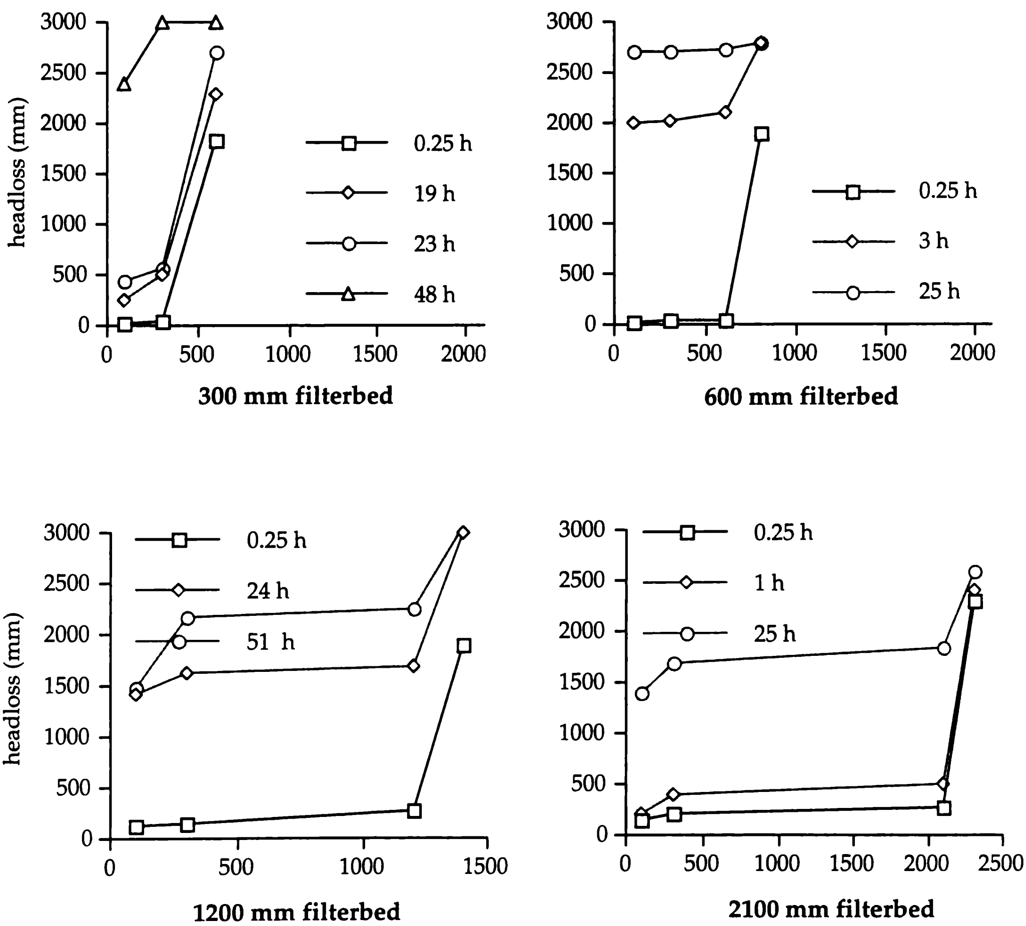


Figure 6-15. Headloss through filter beds at different run times with bed depth measured from the top and nozzle section shown as the deepest data point.

Directly after backwashing, total headloss across all sections for all bed depths was only a few mm. The major component of initial headloss was through the section containing the nozzles. Headloss in the nozzle section remained approximately constant throughout the various runs, indicating that little blockage of the nozzles occurred. Early in each run, headloss built up over the upper sections of the bed. Initially this headloss affected 100 mm, but as the run progressed, headloss built up over progressively deeper sections. By the end of a 50 h run, headloss had increased to about two thirds of the applied head and typically had affected the top 300 mm of the bed. This is similar to the depth where most of the solids were stored (see Figure 6-14).

A feature of all the headloss plots is the relatively high headloss contributed by the nozzle section of the bed. It is clear that the nozzle section was less than optimal in relation to flow resistance. However, the operational head was sufficient to overcome this resistance and allow desired flows for most of the runs.

6.4.8 Variation of hydraulic conductivity with bed depth

In order to determine how the media was being affected during the runs, hydraulic conductivity was calculated at various depths (see Section 2.5.1). The increment of headloss (Δh) across each segment of the bed was measured as the difference in water levels (m) in the successive manometer tubes which were evenly spaced down the filter bed. Data are summarised in Tables 6-3 and 6-4.

Table 6-3. Hydraulic conductivity data (m/s) for a filter run

	300 mm filter bed				600 mm filter bed		
Bed depth (mm)	0.25h	19 h	23 h	48 h	0.25 h	3 h	25 h
100	0.02	0.00022	0.00019	0.0000065	0.015	0.00006	0.00002
300	0.02	0.00039	0.00015	0.000026	0.09	0.00013	0.01
600					0.135	0.0048	0.0155
nozzle section	0.0004	0.0001	0.000096	0.00001	0.00033	0.000095	0.00004

Table 6-4. Hydraulic conductivity data (m/s), continued from Table 6-3.

	1200 mm filter bed			2100 mm filter bed		
Bed depth (mm)	0.25 h	24 h	51 h	0.25 h	1 h	25 h
100	0.0031	0.00028	0.00002	0.002	0.0014	0.00011
300	0.04	0.00033	0.0001	0.01	0.003	0.001
1200	0.028	0.0069	0.0037			
2100				0.08	0.05	0.018
nozzle section	0.0005	0.00007	0.000047	0.00038	0.00034	0.00017

The data show that conductivity loss is consistent with the media being blocked by removed solids. Deeper in the beds, the value of hydraulic conductivity changed very little from the value determined for fresh media (see Table 3-1). When values for the SS media from Table 3-1 are compared to values of residual solids obtained at various depths (Figure 6-14) in the bed, the effect of solids content is clear. The high headloss across the tube nozzles was recognised. These nozzles would not be used for other than TM medium. For the experiments of this study, the restriction presented by the nozzles did not affect the experimental set up, they were serving as headloss control. During the progress of the trial, headloss mainly occurred across the top 500 mm of the bed. The nozzles no longer inhibited the flow.

6.4.9 Removal of unfilterable PTE solids by backwashing

An experiment was performed to determine the effectiveness of backwashing in removing solids from the media. The total bed depth was 1200 mm and the run time was 48 h. The results are summarised in Figure 6-16.

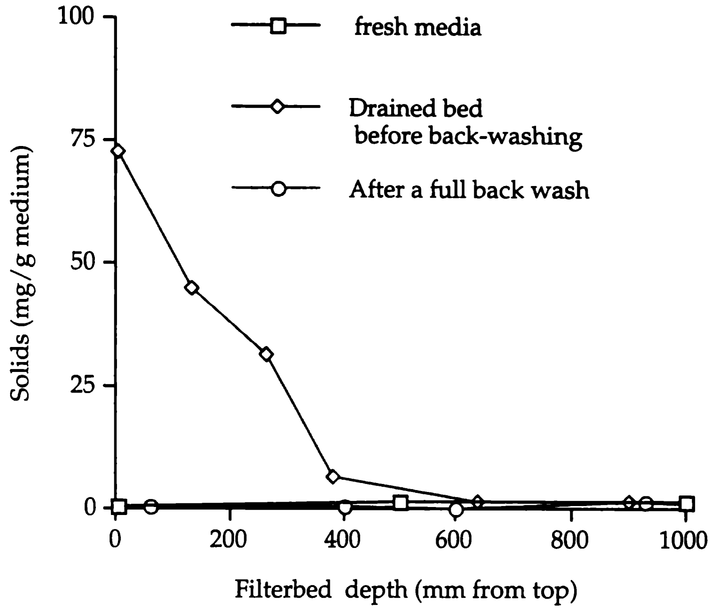


Figure 6-16. Efficiency of backwashing for the removal of solids from a filter bed. Residual solids extracted from backwashed media were similar to the levels extracted from fresh media.

It is clear that a normal backwashing procedure (see Section 6.2.5) removed all retained solids indicating that PTE solids were loosely held by the SS medium. Further evidence of the ease of cleaning of the SS medium of PTE solids was provided by a rainfall event, which dislodged solids from the upper section of a drained, 1070 mm bed. Data for solids extracted from various depths of the bed before and after rain are shown in Figure 6-17.

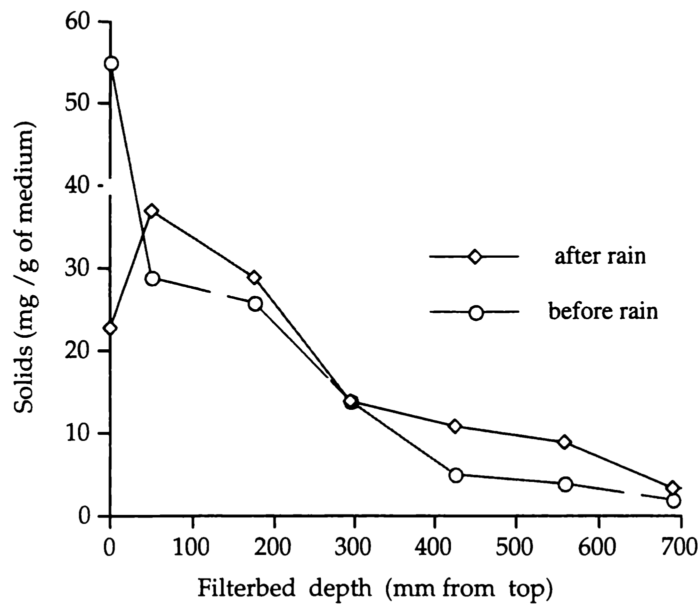


Figure 6-17. The effect of rainfall on solids distribution in a used and drained filter bed after a 48 h filter run.

The effect of rain appears to have dislodged solids from the surface layer of the bed. These solids do not appear to have penetrated more then a few mm into the bed. The solids distribution after the heavy over-night rain showed a displacement down the column. The solids content of the surface layer was reduced by more than 50%.

6.4.10 Pulsed backwashing of a deep SS filtration bed

It has been shown that solids retained from un-flocculated effluent were readily removed during backwashing. In order to take advantage of the easy removal of retained solids, short pulse back washing was investigated. The effluent was drained to within 100 mm of the medium surface and a 10 sec pulse of backwash water was applied. In the process, air trapped in the under drain was forced through the bed, providing a three-phase: medium-air-water backwash motion. The total volume of water used for a pulsed backwash was 25 L. The dislodged solids were sluiced away with unfiltered water remaining above the bed as a 1% TSS sludge after each pulsing cycle. After the pulse, all surface water above the bed, including the water from the pulse was released through the overflow and the run was re-started with fresh effluent.

The filtration efficiency for a 1500 mm deep bed after 4 successive pulses is summarised in Figure 6-18.

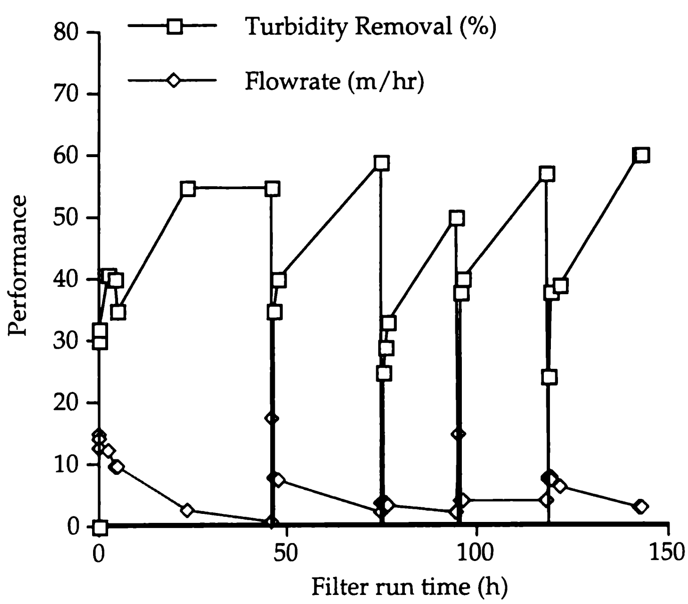


Figure 6-18. Pulsed back washing of a 1500 mm deep bed with SS. Turbidity removal of 40 - 60% and flowrates of about 12 m/h are re-established after each successive pulse.

For each of the 4 pulses, the 1 - 3 m/h flow observed after 24 h filtration was restored to the original flow of 12 m/h. After each backwash pulse, it took about 3 min of run time before filtration efficiency of 40% was restored.

The experiment was terminated during the fifth pulse when the glass wall of the evaluation unit cracked.

Distribution of solids before and after pulsed backwashing

In order to determine how solids distribution changed after pulsed backwashing, medium was sampled at appropriate depths from a drained bed immediately before and after backwashing. Data are shown in Figure 6-19.

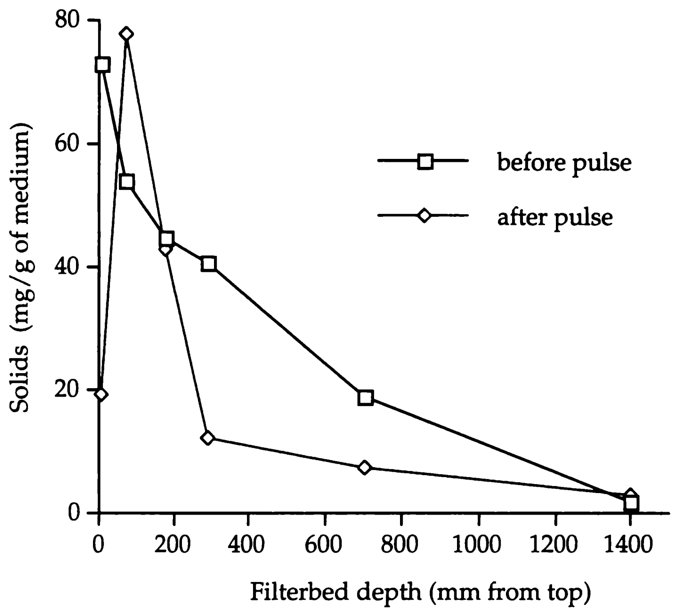


Figure 6-19. Solid retention in a 1500 mm deep bed, following one backwash pulse of 10 sec.

Pulsing removes about 75% of the trapped solids, with some of the solids remaining at a depth of 150 mm. For all but one of the depths sampled, the solids content of the media has been reduced. The point for 100 mm appears anomalously high. Some solids may settle and become buried by overturning the medium. Medium samples from the 100 mm level were visibly more coloured than samples from the surface layer after the pulse. High solids at depth of 10 mm after pulsing haves also been reported elsewhere (Rengel - Aviles, 1986).

Further samples were taken after each filtration cycle following pulsing, to determine the change in concentration and distribution of solids. Data after each pulsed backwash run is shown in Figure 6-20.

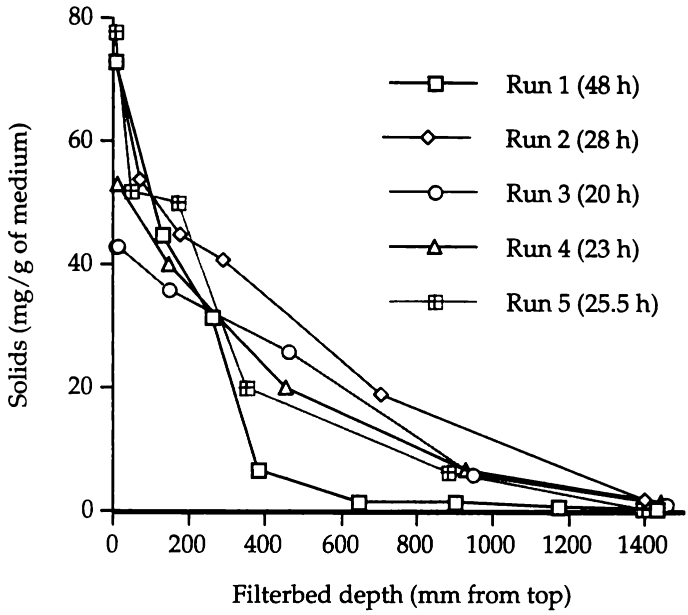


Figure 6-20. Solid retention in a 1500 mm deep bed of SS (7/14) following five successive pulsed filter runs.

Beyond the second run, distribution of solids remained essentially the same for successive runs, indicating that after the first cycle, solids retained during filtrationbuild up during a run were removed by the 10 sec back wash pulse.

6.4.11 Solid recovery and sludge concentration in the backwash water after a pulsed or a full backwashing cycle.

Total amount of backwash water used in one pulsing operation (10 sec) is about 39 L. Total volume required for a full backwashing cycle (eg 14 min water backwashing) is about 2 m^3. Given a 24 h filter run time between pulsing, at an average flowrate of 3.5 m/h, total amount of effluent filtered in the evaluation plant is 12.6 m^3. At a filtration performance of 50% turbidity removal, the amount of solids trapped in a filter bed before pulsing is about 0.5 kg. If 75% of these solids are dislodged during pulsing, a solid concentration in the backwash sludge of approximately 5% can be expected. If a conventional backwash of 14 minutes at flow rate of 60 m/h was used, the volume of backwash water required would be approximately 2000 L and a solids content of approximately 0.5% could be expected. The ratio of volume of backwash water to filtered water for traditional backwashing was calculated to be is 1:34. Pulsed bed operation has the potential to massively increase this ratio.

6.4.12 Deterioration of nozzle performance

During the pulsed backwash trials some deterioration of system performance was observed. Untreated river water provided the source of the backwash water. The river rose above its normal level on the 10th July and stayed high for about 7 days. It is likely that flood-water from the river partially blocked the nozzles during backwashing and led to failure of the system. Large particulate matter in PTE was observed during that period indicating changed solids composition possibly due to storm water-inflow and shorter residence time in the primary settling tanks.

6.4.13 Conclusions from deep bed trials

The principal finding from the work with deep beds was that while a large percentage of turbidity removal occurred over the first 500 mm of the bed, further improvement of effluent quality was achieved with a deeper bed. In a typical run, 45% of turbidity removal occurred over the first 500 mm of bed depth and a further 20% over the remaining 1500 mm of a 2 m bed. For runs of 24 h or less, the majority of headloss and solids retention presented itself within the upper 300 mm of the bed. For longer runs deeper penetration appeared. SS media has the ability to store solids deep into the filter bed, without significant reduction in flowrates. These solids are removed easily by backwashing.

Pulsed backwashing offers the possibility of large reductions in backwash times and backwash water volumes. Resistance to flow contributed by the nozzle section was high, but did not affect the flowrate in the beginning of a filter run. Hydraulic conductivity remained relatively constant throughout the run. During pulsed backwashing, build up of solids occurs, probably due to mixing of soil with the media. No breakthrough occurs, although an initial rinse is probably needed. After a series of pulsed runs, full backwashing would be expected to clean the media. The accumulation of solids throughout the bed is due to pulse backwashing. However the media is capable of retaining these solids during filtration without breakthrough effects at the flows used. More work would be required in order to optimise filter configurations.

6.5 *Multimedia PCDM trials for filtration of secondary treated effluent*

6.5.1 Introduction

Towards the end of the study period, an opportunity arose to test PCDM filtration on secondary treated effluent (STE). The need for an upgrade to the MWWTS resulted in an invitation to WFS for a tender to commission a system for reduction of BOD, N, P and TSS. The work described was planned and carried out in consultation with WFS who provided most of the logistics required. This report summarises the findings from a series of short filter runs with 5 different media configurations.

The trials compared turbidity, TSS and BOD before and after filtration. It was of interest to test the effect of filter bed depth, the media grain size and the consequence of altering the flowrate. In addition, the outcome of adding a flocculating agent was also tested in one run.

6.5.2 Modifications to the tall PCDM evaluation unit

The main modification to the unit described in Figure 6-3 was the installation of 6 TMC double nozzles (see section 5.2.2), resulting in a nozzle density of 42 nozzles/m^2.

Additional modifications included the reduction of the unit height to 2.0 m. A pumping system was installed to provide in-line dosing of the flocculant prior to filtration. Drinking water supply was used for backwashing.

6.5.3 Operation and performance measures of STE trials

Operation of the evaluation unit was described in Section 6.2.4. Filter run times lasted between 5 h - 7 h and the evaluation unit was backwashed between runs. Flowrates were kept constant at either 3 m/h or 6 m/h, by manual adjustments to the outlet valve.

Samples for TSS and carbonaceous biochemical oxygen demand ($CBOD_5$) were collected in the beginning and towards the end of a run. Turbidity readings were carried out on an hourly basis. The following sections summarise the performance, following variations to the filter bed depth, grain size and additions of flocculants. A detailed list of turbidity readings is given for two of the runs in the Appendix 12.

6.5.4 Effect of media depth

For the majority of the trials, a shallow bed with total depth of 680 mm (390 mm SS 7/14 over 290 mm silica sand) was used. Bed depth was increased from additions of silica sand to 400 mm in one trial and from further additions of SS (7/14) to 700 mm in another trial. Flowrates were maintained at 6 m/h. Figure 6-21 shows the effects on $CBDO_5$, TSS and turbidity removal, using deeper beds.

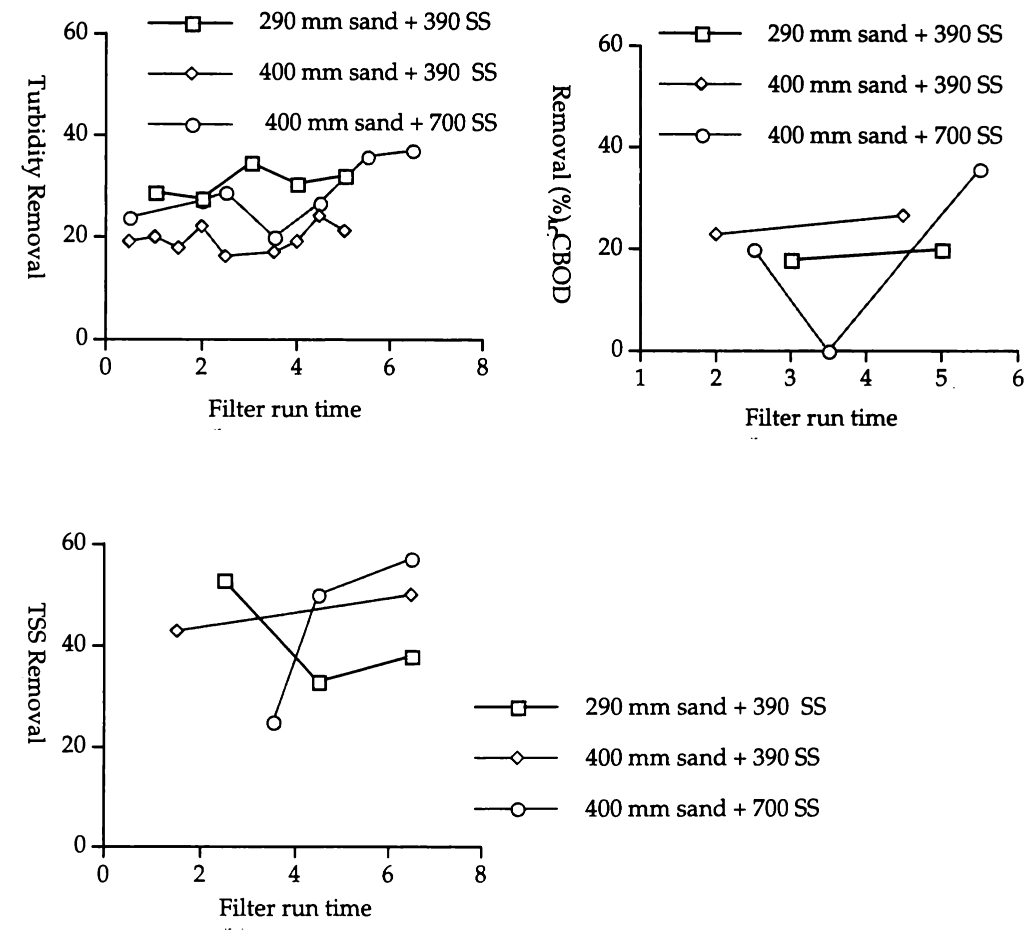


Figure 6-21. Effect of bed depth on filter performance: comparison between a dual media bed with 290 mm Silica sand below 390 mm SS (7/14) and deeper layers of silica sand or SS.

Turbidity removal for the three systems fluctuated between 20 - 30%. Increasing bed depth did not affect the level of turbidity removal during the first 6 h of a run.

Increasing the bed depth with additions of silica sand, improved the TSS removal from about 40% for the shallow bed to about 55% in the deeper beds. Additional SS (7/14) did not improve the TSS removal any further.

The removal of $CBOD_5$ was not affected by bed depth. About 20% removal was recorded in the three systems trailed.

6.5.5 Effect of media grain size

In order to test the effect of finer media size, the deep bed, comprising 700 mm SS 7/14 over 400 mm silica sand, was compared with additions of finer media, such as SS 14/24 and TM. For the first trial, a finer grade of SS 14/24 was used to replace the coarse SS (7/14) on top. This resulted in a multi media system of 400 mm silica sand at the bottom, 400 mm SS 7/14 in the middle and 300 mm SS 14/24 on the top. For a second trial, 150 mm of TM replaced some of the finer Silicon Sponge from the top. In that case, after fluidisation of this multi media bed, the majority of TM sank to the bottom of the filter bed, mixing with the pea-metal. Flow rate was maintained at 6 m/h. Results on the filtration performance are shown below in Figure 6-22.

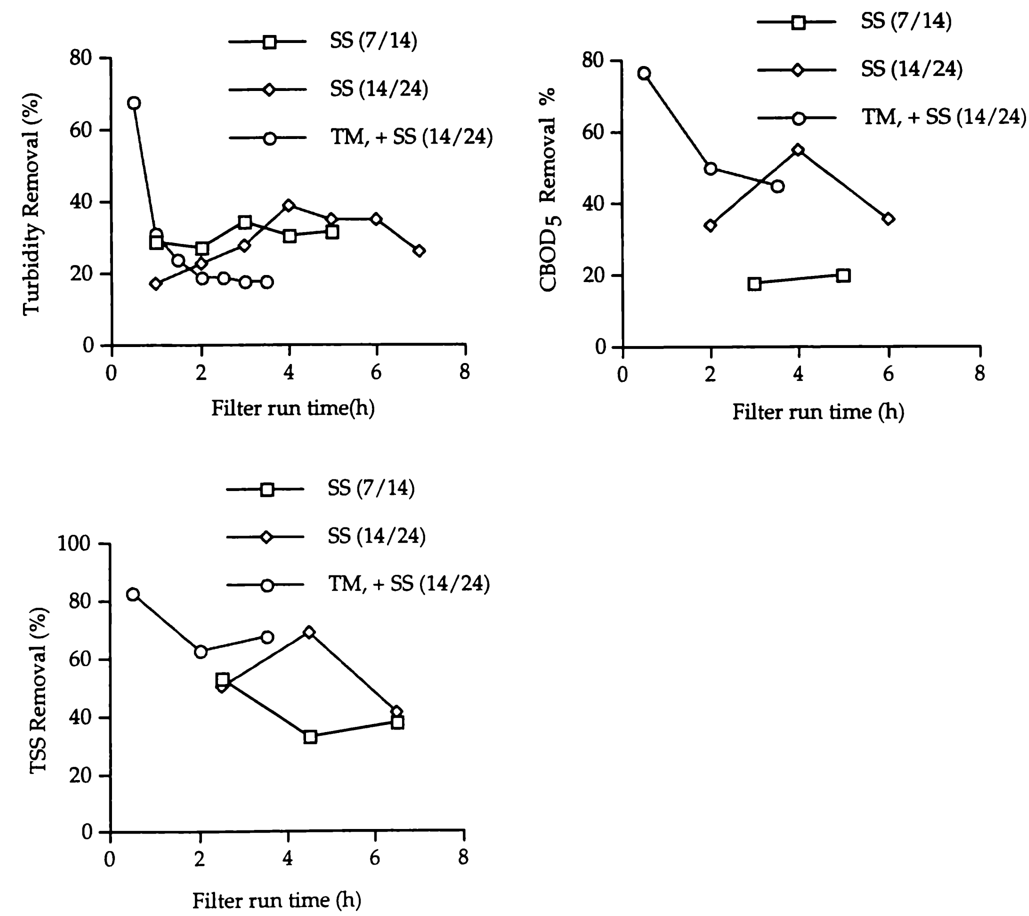


Figure 6-22. Effect of grain size, comparing 400 mm SS14/24 with SS 7/14 over 400 mm silica sand and with TM and SS 14/24 as a second and a third layer. Legend is showing the variable only, being media placed on top of 400 mm silica.

Replacement of coarse SS with finer SS produced similar filtration efficiency in terms of turbidity and TSS removal. The reduction of $CBOD_5$ doubled from about 20 - 40%, following the additions of fine SS. When TM was added to the system, an increase in TSS removal was recorded with the values increasing from 40% to about 55%. As with the fine SS, additions of TM led to a doubling of the removal of $CBOD_5$. However the heavier TM media would settle at the bottom of the bed, following backwashing.

6.5.6 Effect of flowrate

Two different flowrates were tested on the shallow bed system. Both runs lasted about 6 h. Figure 6-23 shows the performance, comparing 2 runs at a flowrate of 3m/h with a run using a faster flowrate of 6 m/h.

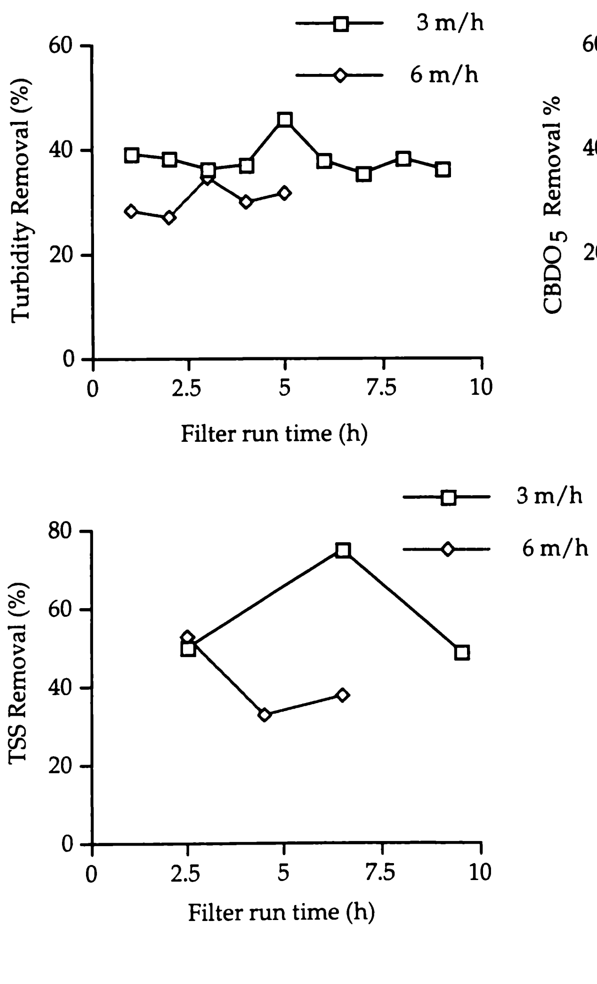


Figure 6-23. Effect of flowrate, comparison between 3 m/h and 6 m/h, using a dual media system with 390 mm SS 7/14 over 290 mm silica sand.

Slower flows led to a significant improvement in TSS removal; up to 60% of TSS was removed which was almost doubling the efficiency seen on a faster flow. Similar improvement rates were observed for $CBDO_5$ where 20% removal at faster flows improved to 40% at the slower flows.

6.5.7 Effect of flocculant

The addition of PFS at a concentration of 100 ppm to the effluent inflow stream, was tested on a shallow bed, using 390 mm SS (7/14) over 290 mm silica sand. A comparison between flocculated and un-flocculated effluent is shown in Figure 6-24.

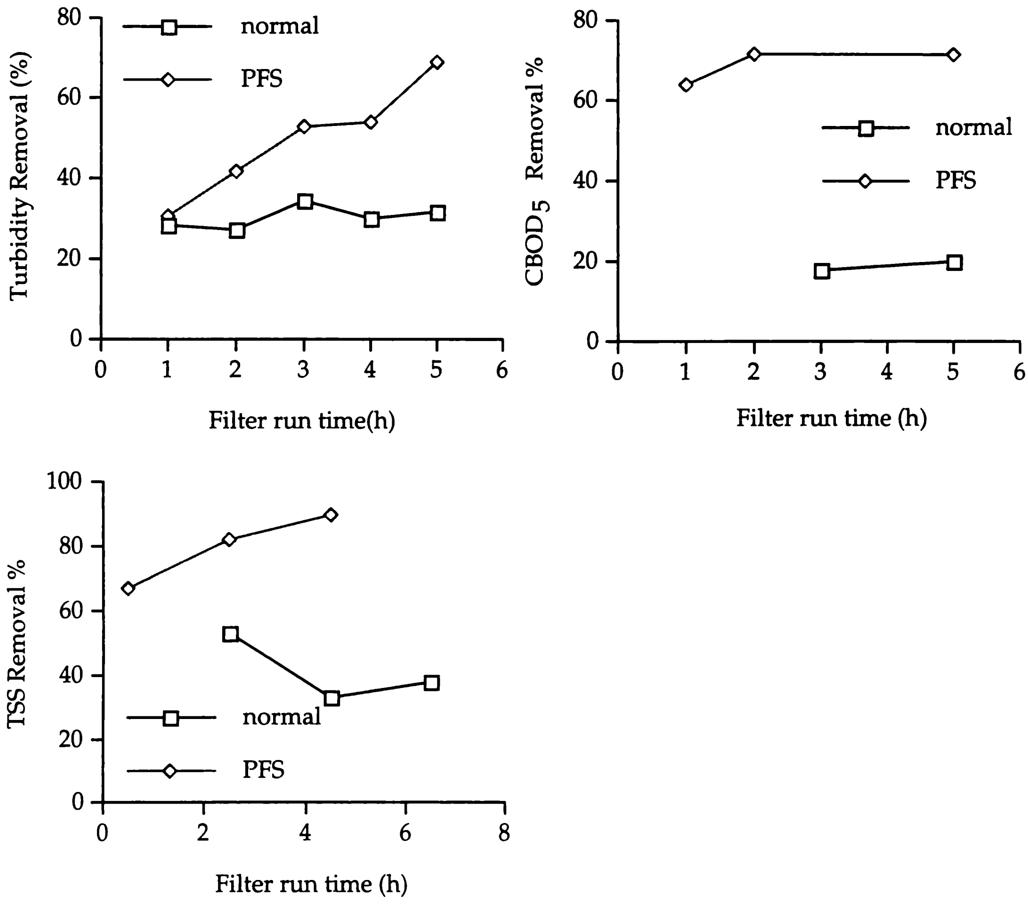


Figure 6-24. Effect after addition of PFS at 100 ppm to a shallow dual media bed.

The addition of 100 ppm PFS led to a significant improvement for all three components tested. The largest improvement resulting from un-flocculated effluent filtration was measured for $CBOD_5$ reduction, indicating that in un-flocculated effluent, most of this component is un-filterable. Turbidity removal increased from about 30% to 60%, following flocculation and TSS removal improved from about 40% to 80%. As the filter run progressed lower turbidity and TSS levels were recorded in the filtered effluent and the filtration efficiency improved.

6.5.8 Summary of STE filtration trials

Un-flocculated effluent

Of the parameters tested during this study, the biggest filtration effect for un-flocculated STE was measured on the removal of TSS. Level of turbidity and CBOD were much less affected by changes to the media configurations.

Increasing the bed depth had less effect than reducing the flow rate. For TSS removal, deeper beds improved TSS removal by 15% whereas reduced flows achieved a 25% improvement. The levels of $CBOD_5$ were not affected by increasing the bed depth. Only when either the grain size or the flow rate was reduced, improved performance led to a 40% reduction, thus doubling the effect from 20% to 40% $CBOD_5$ removal.

When TM with its very small grain size was added to the filter bed, the removal of TSS improved even further, from 40% to about 60%. However, by using the smaller SS media, no significant effect on either TSS or turbidity removal occurred. Only the $CBOD_5$ reduction improved from 20% to 40% by using smaller SS.

Flocculated effluent

Flocculation led to a significant filtration improvement for all parameters tested in this study. A shallow bed with coarse SS (7/14) media performed superior to deeper beds containing smaller grains.

6.5.9 Discussions on STE trials

The STE filter runs were an indication of potential reductions in TSS, turbidity and $CBOD_5$, obtainable using PCDM filtration. It is obvious that flocculation would be necessary to achieve the reduction in turbidity required by MWWTS.

The trials reported here were only of relatively short duration. Performance indicators during un-flocculated effluent filtration consistently drop towards the end of a filter run. Considering the relatively labour intensive backwashing procedures, trials involving longer filter runs should be carried out to test if performance could be maintained over a longer period.

The relative high concentration of flocculant addition is potentially costly. Current market value for 200 L PFS is $280.00 (Jamieson, 2000). To treat a daily flow of 8000 m^3 at the Morrinsville plant, the cost for PFS alone would be over NZ$ 1000/day. Lower

PFS concentrations between 10 - 25 ppm, such as those shown in Figure 4-29, should be investigated. Critical coagulation point was indeed reported by others (Boller et.al. 1981; Strohmeier, 1992) at lower concentrations (4 - 10 ppm Fe III).

Reports on deep bed filtration for STE (Koch et.al. 1992) refer to total bed depths of 1500 mm for coarse multi media. Alternatively, shallower beds were used but the grain size was much reduced and frequent backwashing by suction of the surface layers maintained continuous flows. In both, deep and shallow bed flocculation - filtration operations, TSS reductions of > 90 % were achieved.

Removal of TSS from un-flocculated STE of up to 85% were reported (Hegemann et.al. 1997), using 1400 mm pumice (d = 2.5 - 3.5 mm) over 400 mm Silica sand (d = 0.7 - 1.25 mm).

Flocculation filtration of STE has been reported by Boller et.al. (1981). $FeCl_3$ was introduced twice, first in a pre-settling tank (residual TSS = 75 mg/L) at a concentration of up to 20 mg/L Fe III, to remove 85% of TSS. After that stage, residual TSS values of 12 mg/L, were reduced further by in-line dosing of Fe III at a concentration of 4 - 5 mg/L in combination with poly electrolytes (0.2 mg/L). This was followed by deep bed filtration to remove TSS by up to 85%.

At MWWTS, concentrations of TSS following the existing biological treatment and settling tanks, are probably low enough for in-line dosing of flocculants, which was the approach taken during this study. A combination of flocculant - polyelectrolyte should be investigated in order to reduce chemical consumption. Removal of soluble and suspended matter during the flocculation process has the advantage of achieving P reductions as well. While there is currently no statutory obligation to reduce P levels it is envisaged that New Zealand will need to match current overseas practises in the near future.

Chapter 7
Summary and General Discussion

7.1 *Introduction*

The overall aim of the research presented in this thesis was to consider the possibility that PCDM filtration of primary effluent without the use of chemicals, may have advantages as an effluent treatment strategy. The specific investigations that were undertaken to further this aim included studies of the two novel NZ filtration media (SS and TM) used in the PCDM system, bench trials with FDE and PTE, studies of the flow characteristics of the novel PCDM nozzles and pilot plant trials. Principal findings from these studies are summarised. The improved understanding of the performance of the media and other aspects of the PCDM system, allows an assessment of the use of this technology in the treatment of effluents and the potential advantages of primary filtration. Further work that would be useful to optimise this technology is indicated.

7.2 *Media studies*

7.2.1 Summary of physical characteristics of SS and TM

Chemical composition

Proprietary treatment of river pumice by Works Filter Systems (WFS), involving thermal and chemical processes produce the SS filter medium.

SS shows a similar chemical composition to levels reported for New Zealand pumice (Briggs et.al. 1993). The major oxide is silicon, followed by aluminium and iron oxides. Of the trace elements, the arsenic content in SS is slightly higher when compared with pumice. This might be an effect of seasonal variations in river sediments (Mc Laren and Kim 1995), reflected in SS produced from river pumice.

TM is obtained as an iron oxide concentrate through magnetic separation processes. The TM media used in this study showed an iron oxides content of 77%, and a titanium oxide content of 8%. These levels are similar to those reported previously for ironsands, following magnetic separation (Graham and Watson 1980; Dasler, 1980;

Stokes, 1989). Aluminum oxides of 4%, followed by magnesium oxides represent the next common components in TM.

Grain size and filtration mechanisms

A range of different grain sizes was chosen for the filtration experiments. For the smaller SS 14/24 medium the majority of grains measured between 0.6 and 0.8 mm with a range between 0.6 – 1 mm. For the larger grade SS 7/14 medium, the majority of filter grains measured between 0.8 and 1.5 mm, ranging between 0.4 – 2 mm. The uniformity for the finer grade was relatively high, producing an UC of 1.3, whereas the larger grade showed a wider spread of grain sizes, which lead to an UC of 2.2. High UC may lead to segregation after multiple backwashing, into smaller grains on top and large grains on the bottom.

TM with a grain size between 0.07 – 0.18 mm was used for most of the experiments, where the majority of grains were found in the size range between 0.09 and 0.11 mm, producing a very uniform medium with an UC of 1.3. For specific experiments, sieved fractions were separated from the bulk TM supply, producing small lots of specific sizes between 0.063 mm and 0.18 mm.

Transport and adhesion steps were explained in Section 4.5.3. The fine TM medium leads to an early formation of a surface cake where most of the particles are being strained by the small pores of this medium in the top 10 mm. The coarse SS medium implies different capture mechanisms. Particle accumulation occurs over the top 500 mm bed depth. The particles remained trapped in these top layers and did not rearrange during longer runtimes. Ripening of the coarse media (SS 7/14) is an effect of gradual pore clogging over time. Perhaps coarser media would allow a more even distribution in a deeper bed, thus taking advantage of the greater capacity that a deeper bed has to offer.

The relatively poor filtration efficiency (up to 50% removal) of SS 7/14 medium may be a reflection of the amount of TSS occurring as aggregates large enough to be retained by the relatively coarse SS medium. The addition of flocculant to increase the percentage of aggregated particles could be expected to improve filtration efficiency.

Hardness and resistance to abrasion

Abraded pumice modified to SS has the potential to replace anthracite as a top layer in water and wastewater filters. One of the negative characteristics of anthracite is its relatively low resistance to abrasion. SS was therefore compared with anthracite, using two different methods. Following optical analysis, typical values on the moh hardness scale of values ranging between 6-7 were recorded for SS and between 3-4 for anthracite. When abrasion of SS was tested by loss of weight following continuous backwashing for 22 h (Langdon et.al. 1995), 4.2% loss due to backwashing occurred. The value for anthracite under the same conditions was 13%. Rigorous backwashing is therefore not a problem with SS medium.

Morphology, atypical grains and specific advantages

The surface of SS appears grey-white to brown-grey, it is rough and sponge or froth like and vesiculated. In contrast, TM has smooth edges. It is black with some exposed layers on the surface. An advantage of the SS grain is the increased attachment of biological matter on SS grains during exposure to PTE under aerobic conditions. The relatively long filter run times and greater solid holding capacity can be attributed to this. Some SS grains contain a larger fraction of sealed vesiculated spaces leading to a density < 1 g/cm^3 even after prolonged soaking. These grains are currently discarded. However they could find application in upflow filtration of PTE such as reported by (Flakentoft et.al 1999).

Hydraulic conductivity

High flows can be expected when using the coarse SS media. A 100 - fold increase in the $-K_p$ value was measured when compared with the finer TM. Higher values in hydraulic conductivity are also reflected in larger porosity values. A large porosity can lead to higher solid holding capacity, a potentially useful characteristic for the filtration of high solid PTE. This was reflected in filter run times lasting up to 50 h when using SS and lasting only between 0.5 h – 1 h when using TM.

Fluidisation

In order to achieve sufficient fluidisation during backwashing of the filter beds, a backwashing flowrate for SS of about 60 m/h was established. This agrees with backwashing flowrates recommended for pumice of 55 m/h by Hegemann et.al. (1996).

At this flowrate, bed porosity was 0.7, which agrees with recommendations for maximum hydrodynamic shear (Cleasby et.al 1975). Much lower backwashing flowrates were sufficient to fluidise the finer TM media. Bed fluidisation occurred at 20% bed expansion, requiring as little as 10 m/h in backwash flowrate. Considering the short filter run times for TM filters, frequent backwashing is more feasible where the demand on backwash water is low. A negative effect of easy fluidisation is the development of channeling through the filter bed. Ceramic nozzles have been designed to enhance the distribution pattern during backwashing operation, thus preventing this effect.

Biofilms

When exposed to PTE solids, SS showed large amounts of protein adhesion (3 mg/mL of medium). Slightly lower amounts were found on TM medium (2 mg/mL). The amount of protein attached to SS samples was largest under aerobic conditions, whereas the amount of protein attached to TM was largest under anaerobic conditions. There was also red stain on the sample bags containing the anaerobic TM samples. Given that there was no stain on any of the sample bags containing other media types, Fe released from TM under anaerobic conditions could be the reason. Similar red staining was noticed around the TMC nozzles, at base of the evaluation plants, following a plant shut down of 4 weeks.

7.3 *Bench trials of effluent filtration*

7.3.1 Choice of effluent system

A synthetic effluent containing $CaCO_3$ bentonite and meat extract was produced in an attempt to produce a liquid, which would represent the suspended particles found in PTE. Turbidity of the synthetic effluent was similar to primary treated effluent (PTE). However, when the particle size distribution was measured, it was found that the volume distribution of particles in the artificial effluent did not represent the particle size in PTE. When Farm Dairy Effluent (FDE) was prepared as a suspension from fresh material collected off a local dairy farm, the particle size distribution was similar to PTE. FDE was therefore used for the majority of bench trial work.

7.3.2 PTE filtration by SS, TM and silica sand

Bench studies were carried out where the performance of SS and TM media was compared with the more commonly used silica sand. A filter column with SS 14/24 (d = 0.6 and 1 mm) and silica sand (d = 0.2 – 0.6 mm) at a bed depth of 1000 mm removed about 35% of PTE turbidity. Increasing the bed depth of the SS 14/24 from 1000 mm to 1800 mm increased turbidity removal to 50%. Coarser SS 7/14 filter media (d = 0.4 – 2 mm) achieved only 30% reduction in turbidity at a bed depth of 1000 mm with no further improvements using deeper beds in a small glass column. The finest media, TM (d = 0.07 – 0.2 mm) removed about 40% of the turbidity and the optimum bed depth was 100 mm. Wall effects, when using the small glass columns may cause this low performance. Deeper bed trials on larger scale pilot plant filters are envisaged, using the coarse grain medium where wall effects contribute less to the filtration performance.

7.3.3 FDE filtration by SS, TM and silica sand

About 60% turbidity removal from FDE was achieved by using 100 mm deep TM filters. Flowrates of 1.5 m/h lasted for about 30 min, after which time the beds blocked. By placing 50 mm silica sand on top of an equal amount of TM, the problem of early bed blockage was overcome without any significant loss of filtration efficiency. However this is not a practical system as backwashing would lead to a mixing of the media. Coarser media such as SS and silica sand only removed about 10 – 25% of the turbidity from FDE. These media were therefore not considered as being suitable for the filtration of FDE.

7.3.4 The effect of fines additions

Increased performance was measured by using TM medium with 5% additions of fine media, either in the form of milled TM (d = 25 - 30 μm) mixed with the bed or by additions of 1% $CaCO_3$ (d = 0.15 – 0.6 μm) as a top layer. In both cases, the turbidity removal from FDE improved from about 40 – 60% to 70 – 80%. Additions of fines can further enhance particle size separation from particles < 5 μm to particles < 2 μm. This makes it a potential technology for the removal of cysts of Cryptosporidium and Giardia. The magnetic properties of TM medium may facilitate medium retention during backwashing and magnetic alignment of individual grains, may lead to increased flowrates during filtration.

7.3.5 Effect of effluent pH

Filtration efficiency improved significantly when the pH of the effluent was < IEP of TM but above the IEP of the effluent particles. This was not an effect of increased particle size of effluent particles. It was therefore concluded that increased turbidity removal was an effect of electrostatic attractions between the TM grain surface and the suspended particles in the effluent.

The isoelectric points of TM and SS are 3.75 and 4.1 respectively (see Section 3.5). Thus low pH filtration of effluent using SS may also lead to improved filtration efficiency.

7.4 Pulsed Backwash studies

Earlier studies have indicated a four fold saving of backwash water by using pulsed backwash operation compared with conventional methods (Matsumoto et.al. 1982; Rengel, 1986). Pulsed bed backwashing was tested on TM during bench-scale studies using FDE as model effluent and on SS during pilot plant studies, using PTE.

Titanomagnetite

The problem of early filter bed blockage, short filter run times and large volumes of backwash water required for TM filters can largely be overcome by pulsed bed backwashing operation. With conventional backwashing, the 2:1 ratio in filtered effluent volume and backwash water volume was unsatisfactory. Pulsed backwashing allowed this to be increased to 8:1. It was first tested on TM with FDE, where most of the solids were shown to accumulate in the top 10 mm of the bed. Solids concentration removed from the sludge during pulsed bed operation was about 0.3%.

Silicon Sponge.

Over a filter run time of 50 h about 250 m^3 of PTE/m^2 can be filtered before an increase in headloss results in unacceptably low flowrates. When the solids distribution was determined, it was found that the majority of solids were concentrated in the top 300 mm. Assuming 50% turbidity removal from an initial turbidity of 80 – 100 NTU (90 g/m^3 TSS), and an average flowrate of 5 m/h, a solids load in the top 500 mm of 22 kg/m^3 can be expected. In practise as much as 100 kg/m^3 was found in the upper 100 mm of the bed and about 17 kg/m^3 was found at a depth of 400 mm after a run of 50 h.

A high filtrate to backwash water ratio of ≥ 50 due to relatively long filter run times compensates for a relative large backwash water consumption of 10 m^3/m^2.

Solids loading of silica sand filters, not exceeding 2 kg/m^3 has been recommended (Cooper - Smith and Rundle, 1998). It is one of the advantages of SS that effective filtration can be achieved even at high bed solid loadings. While high solids loading can lead to formation of mudballs, these have not been observed in SS drinking water filters even after prolonged periods and the use of flocculants. (Liu, 1997).

Given a solids holding capacity of 100 kg/m^3 media in the top 100 mm, there is potential for large quantities of sludge to be harvested by pulsing. Based on a 10 sec pulse at 60 m/h flowrate, the expected TSS concentration in the sludge after pulsing could be as high as 5%. While the pulsing experiments showed that solids penetration occurred deeper into the filter bed after pulsing, the solids from the top section were easily removed which restored the original flowrate. Accumulation of solids in deeper regions only occurred between the first and the second cycle, with no further increase between the third and the fifth cycle, indicating that most of the accumulated solids were removed by a 10 sec pulsed backwash.

When pulsed backwashing was applied to SS media, adequate filtration was still observed after 144 h using 4 pulses of 10 sec duration. Efficiency in terms of turbidity removal and flowrate was similar to a conventional run.

Backwash water volume was reduced to about 0.3% of the total filtered effluent and represents a 28-fold decrease in backwash water consumption relative to traditional backwashing. Build up of solids, due to mixing of soil with the media, but no breakthrough occurs. An initial rinse of the backwashed bed may be needed. After a series of pulsed runs, full backwashing would be expected to clean the media.

A significant advantage of pulsed backwashing is that the backwash sludge is of sufficient concentration to be fed, with primary sludge, to anaerobic digestion.

7.5 TMC nozzles

The TMC nozzles of the PCDM system presented no additional resistance during the filtration of PTE. A low density of nozzle distribution of about $40/m^2$ is made possible by nozzle design incorporating the porous frit, which aids even distribution of the backwash water.

Following a trial period resulting in 14 trial runs, 540 h total filtration time and the passing of approximately 450 m^3 of effluent, the flowrate, immediately after backwashing remained constant. Hydraulic conductivity of the TMC nozzle remained unchanged during a filtration run. By replacing the traditional plastic umbrella nozzles with ceramic disc nozzles, overall hydraulic conditions remained the same.

It was of interest to determine the resistance through each component of the ceramic nozzle-bolt system. The water orifice diameter was the factor that determined the flowrate. When up-flowrate was compared with down-flowrate, about 30% higher resistance was found during up-flow. Throughout the experiments the evidence suggested that different flow patterns in the two directions caused this effect. The possibility of a coned entry to reduce the *vena contracta* effect during up-flow could reduce resistance and allow greater backwash flowrates for a given head. While resistance by the nozzle-bolt combination is advantageous in maintaining an even water distribution during backwashing (Geering, 1996), inadequate backwash flow is sometimes a problem and there is potential for energy savings by minimising backwash flow resistance.

7.6 Pilot plant trials

7.6.1 Effectiveness of PCDM system in effluent filtration

The relationship between turbidity and TSS for PTE and FDE (see Figure 2-1) allows the % removal of turbidity to give an indication for the removal of TSS. This makes it possible to compare the PCDM system with other filters used for effluent filtration. In the case of PTE, an effluent containing 100 mg/L TSS could be reduced by 50 - 60% by filtration through 1000 mm coarse SS in a PCDM system. Shallower beds of TM media obtained similar results.

Existing literature values for the removal of TSS from PTE range between 30% and 70% (England et.al. 1977; Brown and Wistrom, 1999). Other reports give turbidity removal between 20% – 60% (Matsumoto and Weber, 1988; Dahab and Young, 1977). In these studies, silica sand (0.35 mm and 0.9 mm) was commonly used with the filter bed depths ranging between 500 mm and 1000 mm.

7.6.2 Effect of bed depth on PTE filtration

Coarser SS media (7/14) was used to monitor accumulation of solids and removal behaviour. Filtration efficiency increased sharply with bed depth, up to a bed depth of about 500 mm. Beyond this depth, further bed depth had less effect. Efficiency increased with bed ripening. In a 500 mm deep bed, 35% of the turbidity was removed by a fresh filter bed, this increased to 45% after about 3 – 5 h of bed ripening. Accumulation of solids occurred mainly through the top 300 -500 mm of the bed and in order to utilise deeper beds, further trials should be conducted using coarser grains.

When a total bed depth of 1000 m was obtained by arranging two 500 mm filters in series, turbidity removal was greater than that obtained for a single 1000 mm bed, particularly after bed ripening. It is likely that surface effects contribute to this result.

7.6.3 Effect of flowrate

When operated with a constant head of 2700 mm, the flowrate of a 1000 mm deep bed SS 7/14 filter decreased from initially 8 m/h to about 3m/h after 48 h run time.

Initial flowrate decrease from 8 m/hr, to 4 m/hr had only a small effect on filtration performance. An increase of 5-20% in turbidity removal was achieved. The effect of flowrates was more pronounced for STE, where increasing flowrates from 3 m/h to 6 m/h decreased the removal of TSS from 60% to 35%. This is consistent with reports by Stevenson (1997), who explained break-through of particles caused by shear effects resulted at flowrates greater than 7 m/h. Brown and Wistrom (1999) also reported negative effects caused by increasing the flow rate from 10 m/h to 20 m/h.

7.6.4 Effect of grain size

The two grades of SS gave similar filtration performance. This contrasts with previous reports for silica sand, where reducing grain size from 1 mm to 0.5 mm resulted in a 100% increase of filtration efficiency, see Section 1.11. It is possible that much of the filtration effectiveness of SS is due to the vesicular structure and is less affected by particle size. SS with even larger grain sizes should be included in further trials to clarify when particle size becomes important.

7.6.5 Effect of effluent pH

As with laboratory trials, acidification of PTE in conjunction with filtration through TM enhanced filtration performance from 50% to 75% in one run. Electrostatic forces are likely to be responsible for the observed behaviour.

7.6.6 The effect of media grain size and type on effluent particle size distribution.

Expected trends were observed. A 100 mm bed of fine TM media was able to filter out 95% of particles greater than 5 μm from FDE. Most of the particles in the size range 8 – 100 μm remained in the top 25 mm of the bed. At deeper bed levels, the proportion of larger particles decreased. Following the addition of 5% fines (milled TM d = 30 μm) most of the particles > 2 μm were removed. An opportunity to remove pathogenic cysts therefore exists.

When PTE was filtered through the coarser SS medium, a 1200 mm deep bed was able to remove only the larger particles (> 30 μm) for the first hour of a filter run. This is consistent with earlier results using silica sand (d = 0.35 – 0.6 mm) (Boller 1993; Lawler et.al. 1993; Ellis and Aydin 1995).

7.6.7 Filtration of STE in pilot plant studies

Filtration by fine SS medium in combination with silica sand to a total depth of 1000 mm achieved a 55% TSS removal from STE. When combined with in-line flocculation, TSS removal increased to 70%. Flowrate had an effect on the filtration performance with slow flows of 3 m/h retaining more solids then faster flows of 6 m/h.

Filtration of STE is more widely practised than PTE filtration and has been studied in some detail by Boller et.al (1977, 1981, 1984, 1987). They found that by careful adjustment of the flocculant addition rate, up to 85% of TSS could be removed.

7.7 *Usefulness of the PCDM technology in wastewater filtration*

7.7.1 Design criteria for a primary effluent PCDM filter

Primary treated wastewater was filtered, using the PCDM system with SS and other media in combination with TMC nozzles. When operated as a deep bed rapid gravity filter, an average turbidity removal of 50% was achieved. Under a continuous head of 3000 mm and a declining flowrate, filter runs lasted up to 50 h with initial flowrates at 15 m/h, reducing to 4 m/h after that time. This is longer than the 8-12 h run times for wastewater filtration reported previously (Cooper-Smith and Rundle, 1998). Backwashing was completed after 14 min, using water and air in succession and about 8% of the filtered effluent volume. Backwash water consumption by conventional potable water filters typically varies between 1 – 5% of the total volume of filtered water (Degrémont, 1991). Larger volumes are required for backwashing of wastewater filters (Matsumoto et.al. 1982, Addicks 1992). Backwash water consumption was reduced considerably, by using pulsed bed operation.

When the fine TM medium was used as a filter medium, turbidity removal was as high as 70%, but filters blocked after less then 2 h run time.

Results obtained from these studies allow the design of a primary treatment system. Calculations of a PCDM filter area should be based on average flowrates of 8 m/h typically declining from initial 15 m/h to 4 m/h. As a filter medium for PTE filtration, SS 7/14 at a depth of 1000 mm should be used. A further 60% depth should be allowed to accommodate for bed expansion/fluidisation during backwashing. It is anticipated that the filter could be operated for up to 2 days, before the build up of headloss and reduced flowrates makes backwashing necessary. In order to fluidise the filter medium, treated water should be used to prevent blockage of the nozzles from the bottom side. A pump should be used which can deliver flowrates up 65 m/h or sufficient to fluidise the SS filter bed. There is a possibility to optimise the nozzle-bolt as part of the filter under floor. This can reduce the resistance presented by the current system by up to 30%. Accordingly a much less powerful pump may be sufficient to achieve fluidisation of the filter medium. Given these configurations, 50% of TSS can be reduced by PTE filtration. This process would reduce the TSS and BOD loading to secondary and tertiary treatment and in addition, due to the smaller particle size, the unfilterable BOD should be easier to digest.

The volatile suspended solids removed could than be treated by anaerobic digestion. For every 4 kg of TSS removed, as much as 1 m^3 methane can be generated, producing about 10 kW of energy. The associated reduction of CO_2 generation during secondary treatment lies between 0.2 – 0.3 m^3 /kg of TSS removed by PTE filtration (Tchobanouglous and Burton, 1998).

Recently PTE filtration efficiency of up to 85% has been reported (Odegaard and Liao, 2003). PCDM filtration technology could be improved by additions of low dose flocculants as shown during STE filtration trials. Recovery and reuse of flocculant could become part of the treatment system. This would firstly reduce chemical costs and secondly limit its possible interference during secondary biological digestion processes.

7.7.2 Application and potential advantages of primary filtration in treatment of Hamilton effluent

Analysis of daily gas yields

Following the primary settling ponds at the HWWTP, the pilot PCDM system removed about 50% of the remaining suspended solids, which in the case of HWWTP, would amount to 2900 kg/day. After primary settling, the filterable solids typically contain between 50 and 75% VSS (see Section 3.10, also Nazaroff & Alvarez – Cohen, 2001, Tchobanoglous and Burton, 1991). Assuming the conservative value of 50%, 1450 kg of additional VSS/day could be expected. Following anaerobic digestion of this material, additional daily methane production of 450 – 700 m^3/day would produce between 4430 - 6660 kW/day. This would be enough to supply between 110 - 167 houses with energy assuming a daily average power consumption of 40 kW. At a current commercial rate of NZ$0.09/kWh (Meridian Energy, 2003) a value of between NZ$142,000 and NZ$213,000 per year is indicated.

Reduction of secondary carbon dioxide production based on reduced VVS would range between 340 - 520 m^3 /day.

It is possible that these values could be improved by increased filtration efficiency that might be achieved by the use of flocculants.

Construction costs of a PCDM effluent filter

Using primary effluent filtration at an average flowrate of 7 m/h, in the case of the HWWTP, a total filter area of 750 m^2 would be required to cope with peak loading rates of 125 000 m^3/day, unless equalisation was build into the system. At a maximum single filter area of 70 m^2, about 10 filters would be required. At a cost of NZ$5000/$m^2$, construction of the filters, including shell, media and internal pipe work would amount to approximately NZ$3.7 M (Works Filter Systems, 2003). Using the higher estimate for returns from gas production, the capital pay back period would amount to about 17 years. With improved filtration efficiency, this period might be reduced to 10 years.

The HWWTP currently operates 2 thermophilic and 2 mesophilic digesters to stabilise sludge both from the clarifiers and the aeration tanks. A co-generation facility fuelled by a blend of biogas and natural gas provides sufficient heat for the thermo-mesophilic digestion system and the total power requirement of the site (Mc Donald & Leach, 2001), saving about NZ$100 000 /year in energy costs. Additional sludge generated from the primary filtration process could lead to energy being available to be fed into the national grid. Furthermore, shifting particle size distribution to a smaller size fractions would allow, secondary treatment to become more efficient (Odegaard, 1998).

7.8 *Other potential application of SS and TM*

Alternative applications of SS could include its use as a coarse media in upflow filtration processes. Its rough surface and high affinity for biological matter makes low density SS a suitable candidate for this technology.

Application of coarse SS as a roughing filter in horizontal direction could be a possibility. This flow direction would take advantage of the larger solid holding capacity and from particle collection in horizontal direction (Boller, 1993). The loose attachment of particles in SS beds (see Figure 6-17) would allow easy vertical drainage during wash-out regeneration cycles.

TM media can be used where short filter runs and intermittent cycles are an acceptable option. The majority of particles are removed by straining mechanisms and settle on the surface of a TM filter. This is specially the case for FDE where a fibrous mat was formed and the filters blocked after < 90 min. The ease of fluidisation of TM makes this medium a suitable system to use in conjunction with short backwash pulsing. In New Zealand, an average 250 unit herd would produce about 12,500 L dairy effluent with an average dry matter content of 0.9% or 112.5 kg/milking and related Nitrogen levels of up to 400 mg/L (Barton, 1994). Primary settling of FDE removes about 90% TSS, of the remaining 11.25 kg, 60% or 13.5 kg/day can be removed by filtration through a shallow TM bed. Bio-digestion of settled solids would produce about 100 m^3 methane/day or 940 kW. Solids from filtration would add a further 6 m^3. However, construction of a filter based on a flowrate of 1.5 m/h, would require a filter area of approximately 8 m^2 costing about NZ$40,000. It seems unlikely that the cost of this application can be justified. It is possible that TM filtration could be used after oxidative treatment to remove suspended solids before discharge. Alternatively it may also be possible to remove colour by flocculation prior to filtration. In this case SS would be a more suitable medium.

One property of TM that might find application is its ability to remove small particles. A problem of increasing importance to water supply authorities is the need to deal with cysts of Giardia and Cryptosporidium. Bench trials demonstrated that TM medium is capable of screening particles >2 µm, and thus should be able to protect water supplies from these pathogens. TM's magnetic property may be useful in such applications. Magnetic stabilisation of the bed may allow increased flow rates in both filtration and backwashing.

7.9 *Recommendations for further research*

7.9.1 Improvement of primary filtration with flocculation

In the present work, the primary effluent was allowed to clarify but no flocculants were added. This was consistent with the aim of the work to achieve treatment without the use of chemicals. Unfortunately TSS removal, even using the efficient PCDM system, was less than optimal, achieving approximately 50% solids removal. It is well known that flocculation, even at low doses, can dramatically improve filtration efficiency. Filtration of flocculated primary effluent using the PCDM system is therefore suggested. This work should determine the optimum flocculant, the optimum dose, options for recovery of the flocculant and the behaviour of the altered sludge during anaerobic digestion. It would be appropriate to aim at 80% solids removal.

7.9.2 Novel applications of SS medium

The affinity of SS for biofilms makes this medium a possible substrate for bio-reactors. The low density of some of the SS medium would make it an ideal for a medium for an up-flow bioreactor.

The use of deep bed coarse SS as horizontal roughing filter (see Section 1.8.8.) should be investigated to take full advantage of the large solid holding capacity, high degree of vesicular space and its affinity for loose attachment of organic material.

Surface modification of the SS may raise the isoelectric point of the SS surface. This should make the medium more effective in retaining negatively charged particles. Some progress in this direction has already been achieved (Yang, 2002).

7.9.3 Novel applications of TM

TM can be effective at very shallow bed depths. One of the limiting factors to reducing bed depth is the depth taken up by the nozzle. With improved nozzle design, beds of 50 mm or even less may be effective filters.

The ability to screen very small particles should be further studied. In particular the ability of TM to remove Giardia and Cryptosporidium cysts should be investigated as an alternative to the use of expensive membrane filters.

The effect of pH on filtration needs to be investigated in more detail. This effect is significant and might be achieved at high pHs by modification of TM to raise its IEP.

The magnetic property of TM may lend itself to filtration application in magnetically stabilised beds.

7.9.4 Applications of PCDM in tertiary treatment

Discharge of effluents to surface water is controlled by the RMA (1991). Performance standards are effect based and receiving waters often have high concentrations of TSS from natural sources (see Section 1.2). In most cases existing primary and secondary municipal treatment systems are able to meet TSS content requirements of current regulations. However criteria based upon nutrient loadings, particularly nitrate and phosphate, are becoming increasingly important. A standard method for reducing phosphate levels is flocculation followed by filtration.

The effectiveness of the PCDM system as a tertiary treatment step has been demonstrated in the third pilot plant trial and the technology is now ready for commercialisation. Optimisation of flocculant dosing, flow rates and bed depths will make the technology more competitive.

7.9.5 Application to small scale systems

The failure of small-scale effluent treatment systems such as those used for small communities, farms, schools etc is often due to the creeping failure of the drainage field beyond the septic tank. Filtration of the effluent from the septic tank may reduce this problem or allow the effluent to be disposed of by other ground-based systems.

7.9.6 Bolt, nozzle and under-drain design

Flow resistance through the TMC nozzle-bolt combination during backwashing operation can be decreased by alterations to the shape of the nozzle-bolt. Careful design of the under drain and fine-tuning of the coning-factor of the bolt has the potential of energy savings of up to 30%.

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Appendix

A 1. Cleaning up procedure for TM

Four different cleaning procedures were tested for TM: They included: cleaning of TM by NaOH, or HCl and mixing it inside a closed container on either a horizontal shaker or in a sonic bath. As an alternative, chemical backwashing in place was tested, using NaOH.

For the container method, dry TM media was placed into a 500 mL screw top bottle and mixed with cleaning solution at a ratio of 350 mL liquid to 250 mL TM. The cleaning solution was either 0.1 mol/L NaOH or 0.1 mol/L HCl. Mixing of the media-chemical mixture was achieved by a motorised shaker at 300 rpm in horizontal direction (Miltern). Further mixing was in vertical direction by hand at about 80 movements/min.

As an alternative to the horizontal mixer, a sonic bath was used and the sample container immersed for 20 min.

Chemical backwashing inside a single filter column was carried out by additions of NaOH to the media then air scouring it for 1 min, creating a media-NaOH mixture.

Efficiency of the washing procedure was evaluated by measuring the residual turbidity in each successive decanted wash water, after rinsing the media by hand shaking it with distilled for 1 min each cycle. Table 2.12 is a summary of the rinsing steps required to achieve constant low residual turbidity.

Table A - 1: Summary of washing procedures for TM to remove turbidity.

Technique - wash solution	Water rinsing cycles required to stabilise residual turbidity values	Lowest turbidity value (NTU)
HCl, 1 min, hand	8	60
	12	50
NaOH, 1 min, hand	7	40
	12	50
HCl, 2 min, shaker	5	10
HCl, 2 min, shaker plus 1 min, hand.	15	50
NaOH, 2 min, shaker	7	150
	7	10
NaOH, 2 min, shaker plus 1 min, hand	18	60
Sonic bath, 5 min	10	20
Backwash air pulsing in NaOH	1	<10

Media washing inside a filter glass column, with NaOH and air scouring produced lower residual turbidity levels when compared with mechanical shaking. This is indicated by low number of distilled water rinses required to establish a close to zero baseline from the backwashed media. It became the adopted method for TM media preparation following grain size elimination by dry sieving.

A 2. Calibration for Malvern particle sizer

Certified poly-styrene beads, type 73 – B1-598 were supplied by ATA scientific PTY Ltd.

Result: Analysis Report

Sample Details

Sample ID: 5/1/00
Sample File: BEADS
Sample Path: G:\SIZERS\DATA\HELENT\
Sample Notes:

Run Number: 1
Record Number: 2

Measured: Wed, 5 Jan 2000 10:31p.m.
Analysed: Wed, 5 Jan 2000 10:31p.m.
Result Source: Analysed

System Details

Range Lens: 300RF mm
Presentation: 3OAD
Analysis Model: Polydisperse
Modifications: Active --

Beam Length: 2.40 mm
[Particle R.I. = (1.5295, 0.0000); Dispersant R.I. = 1.3300]

Sampler: MS17

Obscuration: 13.4 %
Residual: 0.485 %

Result Statistics

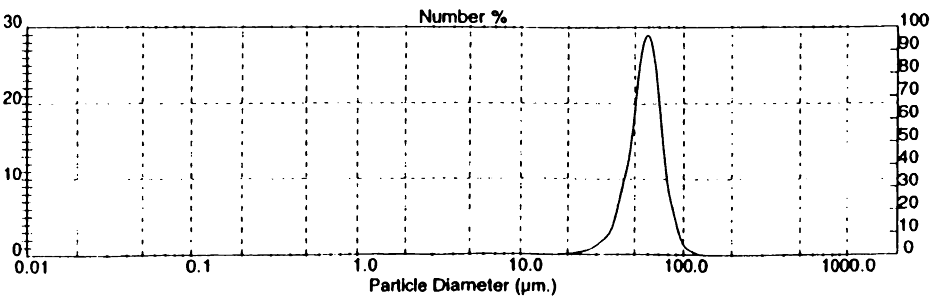
Distribution Type: Number
Mean Diameters:
D [4, 3] = 69.72 um

Concentration = 0.1312 %Vol
D (n, 0.1) = 42.86 um
D [3, 2] = 66.14 um

Density = 2.650 g / cub. cm
D (n, 0.5) = 58.61 um
Span = 5.685E-01

Specific S.A. = 0.0342 sq. m / g
D (n, 0.9) = 76.18 um
Uniformity = 1.832E-01

Size Low (um)	In %	Size High (um)	Under%	Size Low (um)	In %	Size High (um)	Under%
0.05	0.00	0.06	0.00	6.63	0.00	7.72	0.00
0.06	0.00	0.07	0.00	7.72	0.00	9.00	0.00
0.07	0.00	0.08	0.00	9.00	0.00	10.48	0.00
0.08	0.00	0.09	0.00	10.48	0.00	12.21	0.00
0.09	0.00	0.11	0.00	12.21	0.00	14.22	0.00
0.11	0.00	0.13	0.00	14.22	0.00	16.57	0.00
0.13	0.00	0.15	0.00	16.57	0.08	19.31	0.08
0.15	0.00	0.17	0.00	19.31	0.19	22.49	0.27
0.17	0.00	0.20	0.00	22.49	0.41	26.20	0.68
0.20	0.00	0.23	0.00	26.20	0.88	30.53	1.56
0.23	0.00	0.27	0.00	30.53	1.94	35.56	3.50
0.27	0.00	0.31	0.00	35.56	4.52	41.43	8.02
0.31	0.00	0.36	0.00	41.43	11.24	48.27	19.26
0.36	0.00	0.42	0.00	48.27	22.82	56.23	42.08
0.42	0.00	0.49	0.00	56.23	28.80	65.51	70.88
0.49	0.00	0.58	0.00	65.51	19.26	76.32	90.14
0.58	0.00	0.67	0.00	76.32	7.12	88.91	97.26
0.67	0.00	0.78	0.00	88.91	2.09	103.58	99.35
0.78	0.00	0.91	0.00	103.58	0.54	120.67	99.89
0.91	0.00	1.06	0.00	120.67	0.11	140.58	100.00
1.06	0.00	1.24	0.00	140.58	0.00	163.77	100.00
1.24	0.00	1.44	0.00	163.77	0.00	190.80	100.00
1.44	0.00	1.68	0.00	190.80	0.00	222.28	100.00
1.68	0.00	1.95	0.00	222.28	0.00	258.95	100.00
1.95	0.00	2.28	0.00	258.95	0.00	301.68	100.00
2.28	0.00	2.65	0.00	301.68	0.00	351.46	100.00
2.65	0.00	3.09	0.00	351.46	0.00	409.45	100.00
3.09	0.00	3.60	0.00	409.45	0.00	477.01	100.00
3.60	0.00	4.19	0.00	477.01	0.00	555.71	100.00
4.19	0.00	4.88	0.00	555.71	0.00	647.41	100.00
4.88	0.00	5.69	0.00	647.41	0.00	754.23	100.00
5.69	0.00	6.63	0.00	754.23	0.00	878.67	100.00



Malvern Instruments Ltd.
Malvern, UK
Tel:01684 892456 Fax:01684 892789

Mastersizer S long bed Ver. 2.14
Serial Number:

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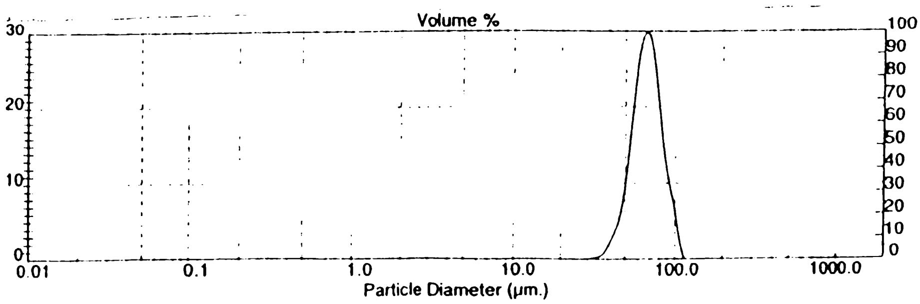

Malvern Instruments Ltd.
Malvern, UK
Tel:01684 892456 Fax:01684 892789

Mastersizer S long bed Ver. 2.14
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A 3. Interlaboratory comparison for TSS testing

Inter-laboratory comparison on TSS between Hill Laboratories, Hamilton and this study

Identical samples were analysed by Hill laboratories and by the author at Waikato University. Results of 5 samples are shown in Table A-2.

Table A - 2. Total suspended solids (mg/L) in 5 effluent types, comparison between two laboratories.

Sample type	University laboratory	Hill laboratories
PTE (16.4.97)	105	118
PTE (20.5.98)	55	58
FDE (20.4.97)	64	72
FDE (8.8.96)	85	87

The difference between the two TSS results varied by < 8%. This was less than the experimental error.

A 4. Calibration tables and results for Protein assay method and calibration curve

Reagents and method for the Biuret assay:

Reagents:

(A) 2% $NaCO_3$ in 0.1 mol/L NaOH (50 mL) + 0.05% $CuSO_4 \cdot 5\ H_2O$) in 1% Na tartrate (1 mL)

(B) Folin-Ciocalteau reagent: (solution of sodium tungstate and soldium molybdate in phosphoric acid).

Method for the Biuret assay:

Pipette 1 ml of each cell suspension into a clean labelled test tube. A phosphate buffer blank which will act as control. Pipette 3 mL of alkaline copper reagnt (reagent A) into each test tube. Mix using the vortex mixer carefully and place in boiling water bath for 15 min. The alkaline copper reagent degrades the cell membrane and allows the reagents to react with the intracellular protein. After cooling to 30° C, add 0.5 ml Folin Ciocalteau's reagent (reagent B) and mix at once. Leave to stand for 30 min. to allow development of the blue colour and read the optical densities (O.D.) in the spectrometer at 750 nm. The standard curve will go up to an O.D. of about 0.7 to 0.8. Above this the linear relationship between OD_{750} and protein concentration breaks down. Solutions should be diluted with water to fall within this range.

O.D values are then converted into µg protein /mL from the standard protein vs OD. The standard calibration curve was established with bovin lactalbumin and is shown below in Figure A-1.

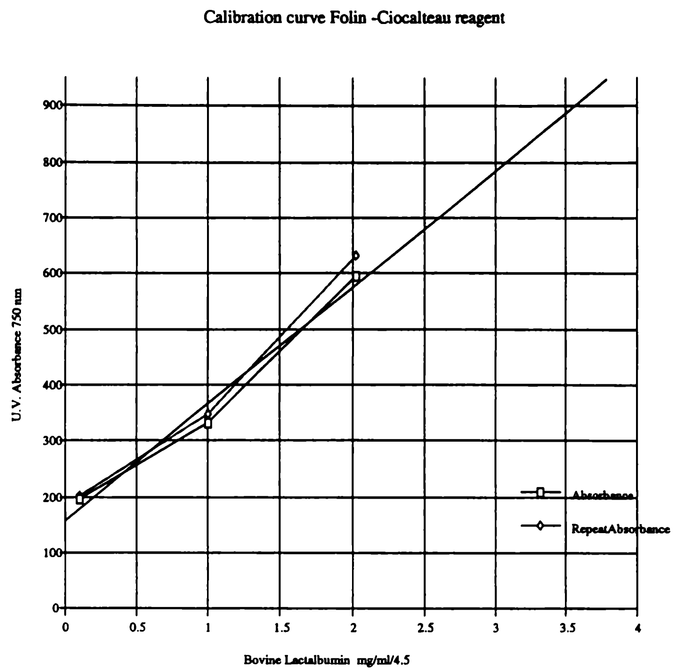

Figure A - 1 Calibration standard curve, using Folin-Ciocalteau reagent and Bovine lactalbumin.

Table A - 3 Results of protein on filter media samples.

Material	Protein (μg/mL media)	Protein μg/m^2 media surface
SS controll	0.012	0.007
SS aerobic	3.0	1.79
SS anaerobic	0.623	0.373
Sand control	0.006	0.0051
Sand anaerobic	0.78	0.67
Sand aerobic	0.47	0.405
Glass control	0	n/m
Glass aerobic	0.71	n/m
Glass anaerobic	0.0125	n/m
TMC control	0.0055	n/m
TMC aerobic	0.39	n/m
TMC anaerobic	0.33	n/m
TM control	0.015	0.023
TM aerobic	0.338	0.56
TM anaerobic *	2	3.3 *

* suspected contamination from colloidal Iron seen as reddish stain on the polyethylene bag.

Working sheet to calculate μg protein/mL surface area.

Density and surface area have been listed in Sections 2.4.3 and 2.4.4.

Additional ρ_A values for TMC and glass have been established during this study, using methods described in Section 2.3.4 (ρ_A glass = 2.6, g/mL; ρ_A TMC = 3.1 g/mL).

Table A - 4. Working sheet for translating protein assay results into mg protein/m^2 medium surface area.

Material	dry weight	Volume mL	Protein µg/mL	Protein mg/g medium	Protein mg/m^2 medium
SS control	1.9115	1.07	0.0133	0.007	0.0014
SS aerobic	0.882	0.49	1.55	0.87	0.174
SS anaerobic	1.164	0.65	0.426	0.24	0.048
Silica sand control	2.737	1.14	0.006	0.0025	0.000926
Silica sand aerobic	1.89	0.78	0.331	0.14	0.052
Silica sand anaerobic	1.938	0.81	0.56	0.23	0.085
TM control	5.619	1.44	0.017	0.00436	0.0054
TM aerobic	4.381	1.12	0.322	0.082	0.1025
TM anaerobic	5.426	1.39	2.37	0.608	0.76
TMC control	2.805	0.8	0.004	0.0013	n/a
TMC aerobic	2.6091	0.745	0.289	0.093	n/a
TMC anaerobic	3.065	0.876	0.29	0.0935	n/a
Glass control	3.446	1.32	0	0	n/a
Glass aerobic	2.707	1.04	0.71	0.27	n/a
Glass anaerobic	2.242	0.862	0.01	0.0038	n/a

A 5. Correction for linear velocity during fluidisation

To normalise the relationship between the three media types, a linear velocity was calculated, using the actual porosity of the expanded bed as the area of flow.

$Q_{lin} = Q_{exp.\,bed}$ /Void volume (mL)$_{exp.\,bed}$ x total bed volume (mL)$_{exp.\,bed}$

where

Q_{lin} = linear velocity
$Q_{exp.\,bed}$ = Velocity in expanded bed (
This resulted in the following flowrates:

- TM: 61.3 m/h
- Sil. sand: 68.7 m/h
- Silicon Sponge : 90 m/h.

A 6. Particle size of retained solids in effluent

Particle size of FDE was analysed during a filter run where flowrates changed during the run.

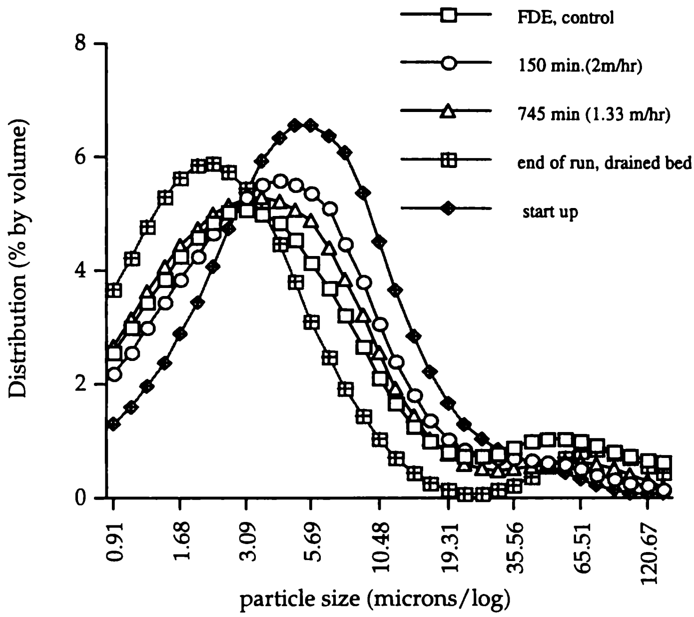

Figure A - 2. Particle size distribution in FDE after and during a filter run through 1700 mm SS 14/24.

Figure 3-37 shows that for unfiltered effluent, most particles are found in the size range between 3 - 20 µm. These sizes are not eliminated during the filtration process, using SS. When the particle size in the effluent, following draining of a bed is analysed, there is an indication that particles, > 3 µm are retained in the bed.

SS was not an effective filter medium for FDE. The removal efficiency never exceeded 20%, which only occurred after about 20 h ripening of the filter bed. When flowrates were reduced during a run, turbidity break through occurred.

A 7. Model relating headloss versus TSS removal by grain size

For completion, a model was developed based on experimental data, relating headloss with media grainsize.

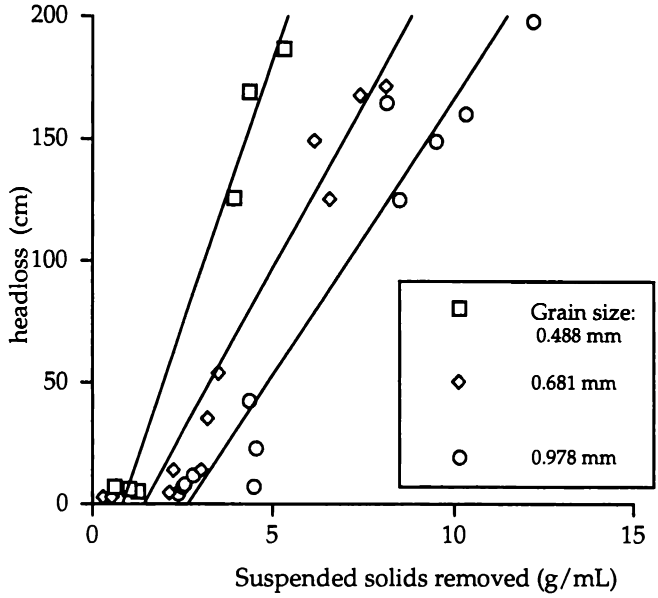

Figure A - 3. Effect of grain size on headloss during the filtration of STE. Adapted from (Tchobanoglous, 1970).

By adapting the model described above, headloss over media size can be predicted. This model does not account for build up of effluent solids during a run or for variation in UC of the media or multi media beds.

A 8. *Repeatability and standard deviation for nozzle flows*

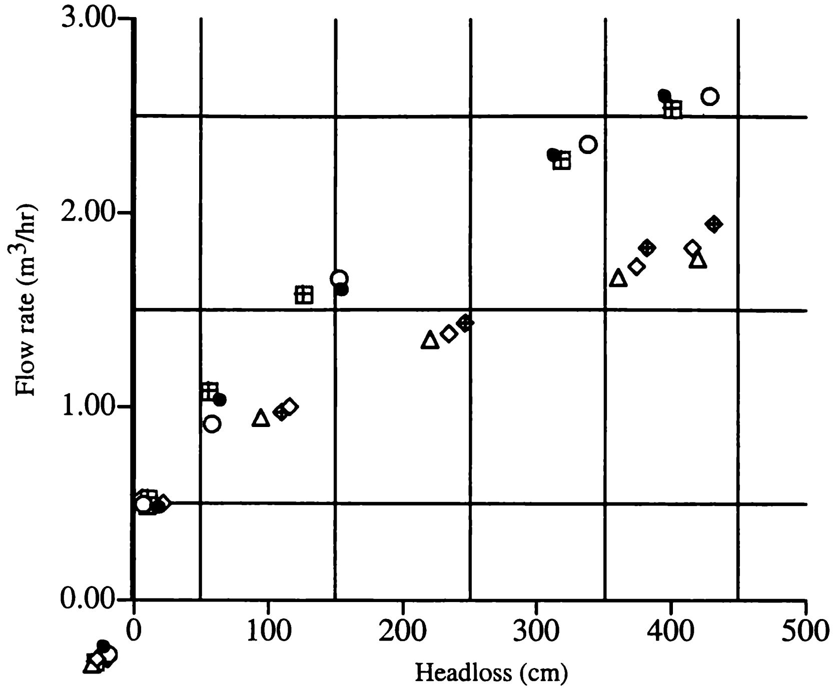

Figure A - 4. Repeatability between different runs through PCDM disc nozzles

Three runs, using standard stem in up flow and down flow direction.

Standard deviation (xσn) for down-flow at 2 m head was 0.0262 m/hr.

A 9. Dates and media configuration for pilot plant trials

Table A - 5 Dates for pilot plant trials

Date	Filter 1	Filter 2	Objective	
Feb. 1995	640 mm Silicon Sponge (7/14)		Turbidity inlet/turbidity outlet.	
Feb. 1995	640 mm Silicon Sponge (7/14)	445 mm Silicon Sponge (7/14)	Turbidity removal in two units with identical medium.	
March 1995	640 mm SS (7/14)	400 mm Silicon Sponge (7/14) + 300 mm Sil.sand	Effect of dual media in the sec. filter	
March 1995	640 mm Silicon Sponge (7/14)	390 mm Sil. sand	Effect of finer media after the first filter	
March 1995	640 mm Silicon Sponge (7/14)	100 mm SS (14/24) + 390 mm Sil. sand	Effect of dual media in sec. filter	
	Filter 1	Filter 2	Filter 3	
April 1995	640 mm SS	450 mm SS	90 mm TM	Effect of TM as third unit (x2)
April 1995	90 mm TM	Effect of TM		
	140 mm TM	Effect of TM (x3)		
	140 mm TM	Effect or reduced backwashing time		
	190 mm TM	Effect of TM bed depth (x4)		
	100 mm TM	Effect of increased backwashing time		
	100 mm TM	Effect of acid treatment (x3)		
	100 mm TM	Effect of caustic treatment		
PCDM trials with tall unit on PTE				
15.May 1998	empty	Testing flowrate at 3.1 m headloss, Backwashing, 60 m/hr		
16. May, 1998	1500 mm SS	First trial run.		
16.June 1998	1600 mm SS	Testing maximum filterrun time and pulsing		
10. June	1500 mm SS	Run for 48 hrs		

1998		
3. June 1998	1200 mm SS	Run for 51 hrs
22. June 1998	1500 mm SS	Run for 46 hrs
18. May 1998	300 mm SS	Bed depth effect, run for 5 hrs
18. May 1998	190 mm Silicon Sponge	Bed depth effect, run for 4 hrs
20. May 1998	300 mm SS	Bed depth effect, run for 48 hrs
25. May 1998	1200 mm SS	Bed depth effect, run for 50 hrs
28. May 1998	1200 mm SS	Bed depth effect, run for 4 hrs
23. July 1998	2100 mm SS	Bed depth effect, run for 25 hrs
11. August 1998	2100 mm SS	Bed depth effect, runl for 32 hrs
PCDM trials on tall unit with STE		
13.Jan 1999	SS (7/14) 390 mm + Sil. sand 290 mm, + Pea metal, 260 mm	
19.Jan. 1999	as above	
21.Jan.1999	as above	
22.Jan.1999	as above	Effect of PFS between 100 - 200 ppm
26.Jan.1999	SS (7/14) 390 mm + Sil. sand 400 mm + peametal 110 mm	Effect of multi media bed
27. Jan. 1999	SS (7/14) 700 mm, sil. sand 400 mm, peametal 110 mm	
28. Jan 1999	SS (14/25) 300 mm + SS (7/14) 400 mm + sil. sand 410 mm + peametal 110 mm	Effect of deep multi media bed
29. Jan. 1999	SS (14/25) 150 mm, SS (7/14) 400	Test he effect of TM as polishing medium mixed in with he peametal at the

	mm, sand, 250 mm, peametal - TM mix 250 mm	sil. base.	

The trials during 1995 were carried out by Choo L. on behalf of WFS and details of the exact dates are unavailable.

The trials during 1998 were carried out by Blattler A. for this study.

The trials during 1999 were carried out by White A. on behalf of WFS.

A10. Repeatability studies for NTU on PTE filtration.

Table A - 6. Repeatbility of turbidity removal performance based on filtration of PTE through 2 filters with 450 mm Silicon Sponge each.

Filtration time	Turbidity removal %	mean	STD (xσn)
4 - 5 h	32.5, 28, 26.5, 32.3,	29.66	1.91
20 - 25 h	30.5, 37.5, 35.4, 29.5, 27	31.98	3.7
25 - 30 h	39.4, 33, 34.4, 33, 51, 37, 27, 22, 37	34.86	7.6
45 - 50 h	9.4, 20.7, 25.1, 25.7, 30.3, 30.4	23.6	7.16
70 - 80 h	3, 20.5, 18.4, 27.4	17.3	8.9

A 11. Data of a full filter run during deep bed SS 97/14) trials

The list below illustrates daily fluctuations in turbidity of PTE and shows actual sampling times.

Table A - 7. PTE filtration by SS (7/14) at 1700 mm depth during pilot plant studies.

Date	time	Run time (h)	Turbidity in (NTU)	Turbidity out (NTU)
16.6.98	11.25	0	76	51
	11.45	0.5	77	49
17.6.98	9.30	22	44.7	28.9
	10.30	23	66	33
	10.35	23.1	68.5	35.4
	16.45	28	119	51.4
18.6.98	10.35	47	70.6	15.5

A 12. Data on Morrinsville trials

In order to illustrate the daily range of turbidity, Table A- .. below shows date of two filter runs.

Table A - 8. Filtration of STE, using 390 mm SS (7/14) over 290 mm silica sand

Date	Time	Turbidity in (NTU)	Turbidity out (NTU)
12'01.02	8.30	8.5	1.3
	9.00	8.7	5.3
	9.0	7.1	4.9
	10.00	8.9	6
	11.00	9.7	6.1
	12.00	7.9	5.8
	13.00	7.8	5.5
	14.00	8.0	5.6
	15.00	7.4	5.2
	16.00	7.9	5.9
	17.00	7.8	5.7
	18.00	7.6	5.9
	18.30	8.1	6
13.01.99	9.00	7.8	1.8
	10.0	8.2	5.2
	11.0	8.2	5.7
	12.00	8	5.6
	13.00	7.3	5.7
	14.00	7.7	5.6
	15.00	7.3	5.9
	16.00	8.2	5.1
	17.00	7.0	5.2

A 13. Schematic of main treatment process at HWWTP

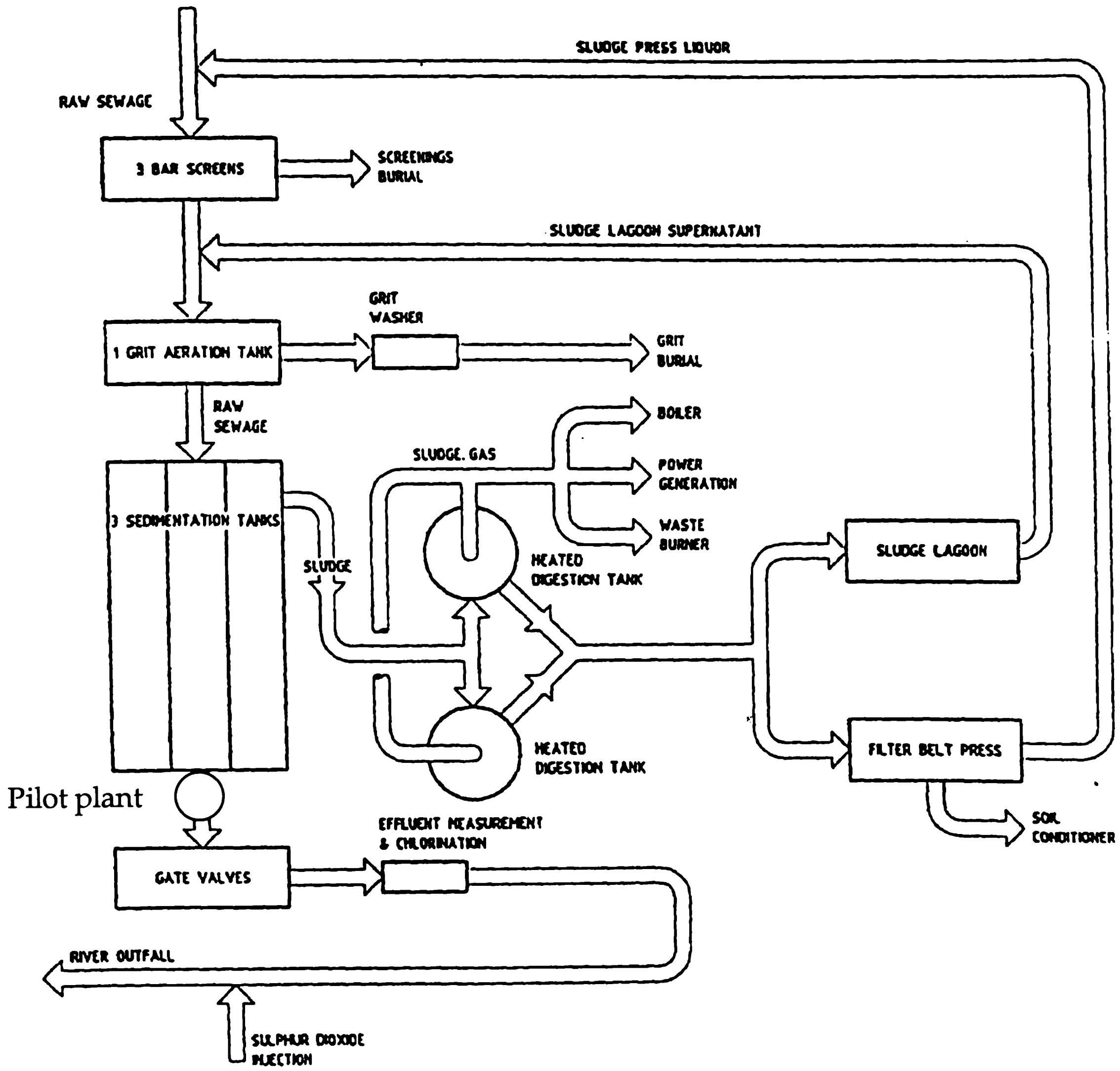

Figure A - 5 Schematics of HWWTP with indication of pilot plant position

A14. Calculations on TSS before and after PTE during wet and dry medium flow at HWWTP.

Dry weather based on 45000 m^3

15 hrs@50 $g/m^3$6 am - 12 pm 28 000 m^3 = 1400 kg

3 hrs @120 g/m^3peak conc. 12 - 3 pm5625 m^3 = 675 kg

6 hrs @40 $g/m^3$10 pm - 6 am 11250 m^3 = 450 kg

Daily total = 2525 kg, medium = 56.1 mg/L

wet weather based on 125000 m^3

15 hrs@70 $g/m^3$6 am - 12 pm78125 m^3 = 5468 kg

3 hrs@120 g/m^3peak conc. 12 - 3 pm15625 m^3 = 1875 kg

6 hrs@ 50 $g/m^3$10 pm - 6 am31250 m^3 = 1562 kg

Daily total = 8905 kg, medium = 71.2 mg/L

Inflow, based on 70% reduction through PTE, see section 1.6.1:

Dry weather: 4690 kg187 mg/L

Wet weather 25443 kg237 mg/L

A. 15 Specialised media types and modifications

Specialised media types

Less common media types include high-density garnet (4.5 g/cm^3) and perlite (volcanic glass with density of 0.8 - 1 g/cm^3). The first is used as fine polishing bottom layer and the second as a flotation media for the removal of emulsified oil products (Spevakova, 1995). In Turkey, a limited availability of silica sand and the abundance of perlite have resulted in this medium being widely used. Filtration efficiency is similar to silica sand with the advantage that perlite is more resistant to acid abrasion (Ulatam, 1992).

Deep beds of walnut shells have been used against oil spills and to reduce oil concentrations from 100 ppm to 1 - 2 ppm. After the filtering cycle, the shells were pumped through a scrubbing system for cleaning (Toohill et.al. 1999). Another medium that has successfully been used to remove oil contaminations is diatomaceous earth. It require a large media to TSS ratio and is commonly used as filtration media in swimming pools, supported by silica sand (Kemmer, 1987).

Zeolites can be used as polishing filters where nitrogen removal is also required. They operate as ammonium selective ion exchangers during tertiary treatment. During trials with a bed depth of 1 m and a flowrate of 3 - 5 m/h, a reduction of NH_4 from about 35 mg/L to 10 mg/L was achieved (Witte and Keding, 1992). In a patent, described by Kumamoto et.al. (1986), zeolite was coated onto pumice by immersion in NaOH, SiO_2, and Al_2O_3 mixture at 80 - 150^{0}C.

Granular activated carbon can remove dissolved contaminants by adsorption to its large surface area. The carbon is sourced from coal, peat, lignite wood, bone; petroleum based residues or nutshells. Carbon filters are used in the beverage and sugar industry to remove odour and colour producing contaminants. These filters can also be used as a tertiary treatment step for wastewater. In addition to regular backwashing, an extended regeneration cycle by heat treatment is required to restore the adsorption capacity (Salvato, 1992).

Coated media

Silica sand, as reported by Ahammed and Chaudhuri, (1996) can be coated with silica sand with iron, aluminium, magnesium and calcium hydroxides. In each case, removal of heterotrophic bacteria was improved when compared with uncoated sand. Iron and iron-aluminium hydroxide showed superior removal qualities compared with other metal coating, with an increase from 64% for uncoated sand to 84%. The higher charge of the metal-coated sand facilitated adsorption of the bacteria and viruses.

Carbonised pumice formed from the controlled oxidation of films of phenol-formaldehyde resins and sucrose has been trialed. Improved adsorption capacities for humic acid and for H_2S were achieved (Langdon et.al. 1994). More recently, pumice was modified with a coat of aluminium hydroxide, to increase its isoelectric point (IEP) from 4.1 to > 8. The effect was to increase the removal of negatively charged colloidal particles (Yang et.al. 2000). The same author, by coating pumice with magnetite, achieved vertical magnetic stabilisation of the filter bed allowing improved hydraulic conductivity through fine filter media. The effect of charge on the medium has been tested by coating sand with FeAl- hydroxide (Shaw et. al., 2000), where an electropositive zeta potential on the grain surface has lead to enhanced adsorption of the electronegative Cryptosporidium oocysts.