Chapter 6

Particle Removal by Filtration Processes

6.1 INTRODUCTION

While a great variety of filters is used in water and wastewater treatment practice, they can be differentiated by their mode of action into two broad groups: (a) deep-bed filters and (b) surface filters.

The traditional filter type used in water treatment is the deep-bed filter, consisting of a permeable granular medium (commonly silica sand) through which the water to be filtered flows, and within the pores of which particulate material is to be retained, hence the description 'deep-bed'.

Surface filters separate particles by a sieving or blocking mechanism; they typically contain a fabric or membrane which permits water flow in response to a pressure difference or gradient but constitutes a barrier to particle transport. The use of membrane filtration in water and wastewater treatment is a more recent development in filtration technology, made possible by the introduction of suitable synthetic membranes in the 1960s and sustained by the ongoing development in membrane technology since that time.

6.2 DEEP-BED FILTERS

The process of filtering water through beds of granular media in order to purify it is in general use throughout the world. Many different types of filter are used with the principal objective of removing microscopic suspended particles from water. Deep-bed filters may be broadly classified as 'rapid' or 'slow' according to the rate at which they operate, with the further distinction that in slow sand filtration there is very significant biological activity, whereas in rapid filtration physical removal is the important factor. The process is a dynamic one in which the change in concentration of the suspension flowing through the bed is a function of depth and time.

Table 6.1 General features of slow and rapid filters

Parameter	Slow sand filtration	Rapid sand filtration
Rate of filtration (m h^{-1})	0.1-0.2	5-10
Size of bed (m^2)	5-200	5-200
Depth of sand (m)	0.6-1.2	0.6-1.2
Sand effective size (mm)	0.15-0.30	0.6-1.2
Sand uniformity coefficient	<3	<1.5
Grain size distribution	unstratified	Stratified by backwashing
Filter floor	Designed for filtrate collection	Designed for filtrate collection and backwash flow
Loss of head limit	Water depth on bed	Water depth on bed
Duration of filter run between cleanings (d)	20-60	1-3
Penetration of suspended matter	Superficial	Deep
Method of cleaning	(1) scraping off surface layer and renewing at intervals (2) mechanical washing of surface layer insitu	Backwash with air and water or by water only at high rate
Amount of washwater used for cleaning sand (% filtrate)	0.2-0.6	2-6
Pre-treatment of water	Usually none	Chemical coagulation + clarification
Supplementary treatment	Disinfection	Disinfection
Relative capital cost	High	Low

The general features of slow and rapid filters are summarised in Table 6.1. Slow sand filters are conventionally operated at a filtration rate in the range 0.1-0.2 $m^3m^{-2}h^{-1}$ (approximately 0.029 mm s^{-1}) and rely to great extent on the formation of a biological layer or 'Schmutzdecke' for their cleansing action. In rapid filters the rate of treatment is typically in the range 5-10 $m^3m^{-2}h^{-1}$ and the full depth of the filter bed contributes to purification and consequently has to be cleaned. Natural silica sand is generally used for both filter types but other media are possible, for example crushed anthracite.

Rapid filters operate at a much higher rate and hence require frequent cleaning. This is done by backwashing the sand insitu using filtered water, with the aid of mechanical, air or water scouring.

6.3 SLOW SAND FILTRATION

Slow sand filters have been used (Graham, 1988) for the purification of public water supplies since the early part of the nineteenth century. In the early days, prior to the introduction of chlorination for water disinfection, slow sand filtration provided an effective barrier against the transmission of waterborne pathogens. For example, Dublin City constructed slow sand filters at Roundwood, Co. Wicklow in the 1860s to treat impounded River Vartry water. This scheme is still in operation in 2006, supplying about 77 Ml d-1.

A typical schematic layout of a slow sand filter cell is shown in Fig 6.1.

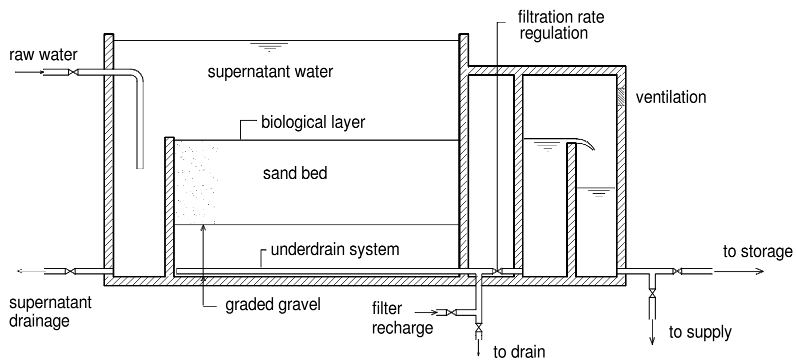

Fig 6.1 Schematic layout of a slow sand filter cell

Slow sand filtration effects a modest removal in colour and can only be used as a sole treatment process where the raw water colour is less than 15-20 ^oH. Where the raw water silt content is high or variable, pre-treatment by roughing filters is required. Because of the relatively high colour of many surface waters, slow sand filtration is not widely applicable as a sole treatment process. Worldwide, however, it is still a significant unit treatment process in modern water process technology, albeit with rather limited application.

A typical slow sand filter unit is composed of 0.15-0.3 mm effective size sand and a uniformity coefficient <3 (see section 6.8), with a bed thickness in the range 0.6-1 m. The sand bed is unstratified and is supported on an ungraded gravel layer, designed to prevent the ingress of sand into the underdrain system. The latter, which is designed to convey the filtrate to a central collector channel, may typically consist of a system of perforated lateral pipes discharging to a central manifold pipe or collector channel. Many other types of proprietary underdrain systems are also available, including perforated floors and narrow slit nozzles, the use of which may eliminate the need for gravel layers. Very small aperture systems should be avoided as they are prone to the development of increased head loss with time due to the accretion of material, which reduces the aperture size. Silica sands and gravel are used because of their durability and non-reactive qualities.

Biological activity plays an important role in slow sand filtration. After a few weeks operation the uppermost layer of sand grains become coated with a gelatinous biological film, which forms a continuous surface mat or ‘Schmutzdecke’ consisting of bacteria, algae, protozoa and colloidal matter derived from the water. This membrane-like structure is considered to be a key factor in the treatment efficacy of the slow sand filtration process. Bacterial action, resulting in the oxidation of biodegradable substances, extends some distance into the sand bed.

The length of filter run is typically of the order of 30 days. The head loss across the bed, which initially may be about 50-75 mm, increases as the pores of the upper layers of the filter are reduced in size by the deposited material. To avoid negative pressure development within the sand bed (see Fig 6.8) the permissible head loss is usually limited to a value equal to the depth of water above the sand surface, which is typically in the range 1.0-1.5 m. When the head loss has reached the limit value, the bed is drained down and cleaned. The usual method of cleaning is by manual skimming the top 12-20 mm of sand. The filter is then recharged slowly by the upflow of filtered water from storage. The skimmed sand is washed and stored for subsequent re-use. Re-sanding is carried out when the skimming operations have reduced the sand bed thickness to about 0.5 m. At large slow sand filter installations the process of filter cleaning has been mechanized, thus reducing filter downtime.

6.4 RAPID FILTERS

Rapid filters are most frequently used in water treatment following pre-treatment by chemical coagulation/clarification; they are also used in wastewater treatment for tertiary effluent ‘polishing’. Rapid filters are usually constructed as open-top, free surface units (rapid gravity filter, RGF) and less frequently as in-line pressure filters. Schematic arrangements are shown in Figs 6.2 and 6.3. Pressure units are used mainly in small installations. While the use of pressure filters may eliminate a pumping stage, they have the disadvantage that the condition of the sand bed cannot be visibly inspected.

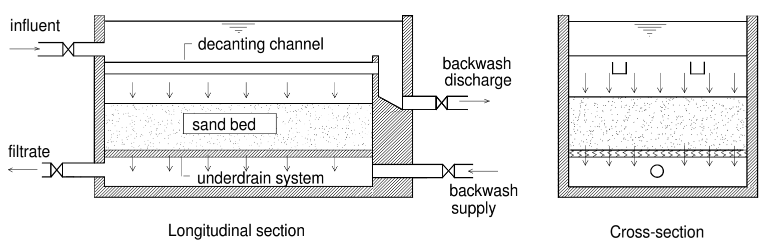

Fig 6.2 **Schematic layout of a rapid gravity filter**

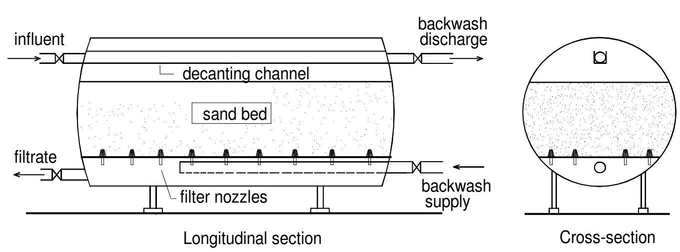

Fig 6.3 **Schematic layout of a pressure filter**

Rapid filters typically contain about 0.6-1 m of sand overlying an underdrain system. The depth of water overlying the sand surface in RGF units is generally not less than 1m.

6.4.1 Rapid filter underdrain systems

The underdrain system in rapid filters is designed to transmit filtrate and to ensure a uniform distribution of back-wash water and, where used, air. It must also prevent loss of sand with the filtrate. Many types of underdrain system have been devised, two of which are illustrated in Fig 6.4.

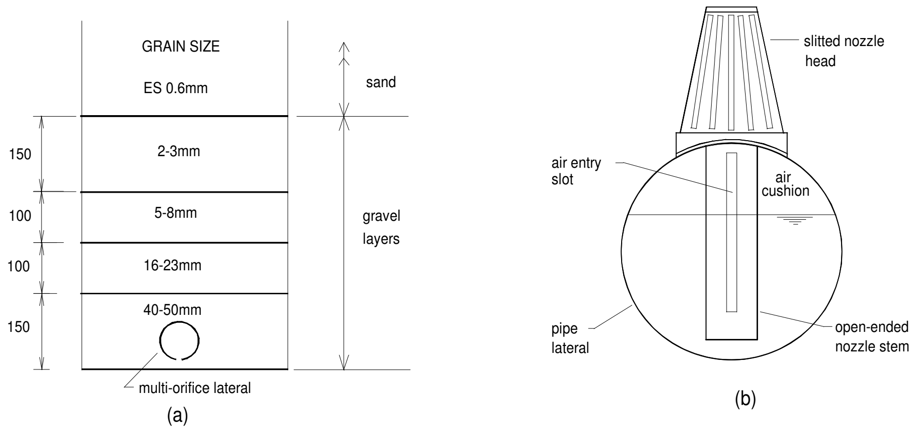

Fig 6.4 **Examples of filter underdrain systems**

The most widely used system consists of a set of perforated pipe laterals surrounded by graded silica gravel layers, as illustrated in Fig 6.4 (a), which shows a four-layer underdrain system with a grain size ranging from 2 mm in the uppermost layer to 50 mm in the bottom layer. The main function of the gravel layers is to distribute the back-wash upflow evenly over the bed area and to prevent the penetration of the sand into the pipe lateral system. Pipe laterals are usually 75-100 mm diameter and are spaced at 150-225 mm centres, with orifices of 6-12 mm diameter at similar centres, discharging downwards. They are connected to a central pipe or channel manifold. The manifold and lateral pipe system are hydraulically designed to ensure a uniform distribution of back-wash water over the filter area. Where air is used for filter cleaning, a separate air manifold and lateral system is provided.

Aternative underdrain systems incorporating nozzles or no-fines porous concrete may partly eliminate the need for gravel layers and may also provide for air distribution, as shown in Fig 6.4 (b). In terms of freedom from clogging and long-term performance reliability, the pipe lateral system is probably still the best available system. Casey (1992) presents an analytical procedure for the computation of flow in pipe lateral/manifold systems of the type used in filter underdrain systems and also in flow distribution systems in sludge-blanket clarifiers.

6.5 HYDRAULICS OF FILTRATION

In 1856 Darcy showed that for the steady laminar flow of a liquid through a homogeneous granular medium, the rate of flow was linearly related to the pressure gradient:

$$Q = KA\left(\frac{\Delta p}{\Delta l}\right) \qquad (6.1)$$

where Q is the volumetric flow rate, A is the cross-sectional area, $\Delta p/\Delta l$ is the pressure gradient and K is the permeability coefficient, the value of which is dependent on the structure of the porous medium and the liquid viscosity. To separate the properties of the liquid from those of the solid material, the permeability coefficient was subsequently modified to

$$K = \frac{k}{\mu}$$

where k is the specific permeability of the medium and μ is the viscosity of the liquid. The specific permeability, k, is an empirical constant which depends on the nature of the material, its mode of packing, porosity and other physical properties. Kozeny (1927) postulated an hydraulic radius model for the flow resistance through a porous medium such as sand,. Using the Hagen-Poiseuille equation for laminar flow through a capillary tube:

$$\frac{\Delta h}{\Delta l} = 32\frac{\mu}{\rho}\cdot\frac{v_c}{g}\cdot\frac{1}{d_c^{\,2}} \qquad (6.2)$$

where v_c is the mean velocity and d_c is the tube diameter. Expressing d_c in terms of hydraulic radius, $R_h=d_c/4$:

$$\frac{\Delta h}{\Delta l} = 2\frac{\mu}{\rho}\cdot\frac{v_c}{g}\cdot\frac{1}{R_h^{\,2}} \qquad (6.3)$$

The analogous parameter to hydraulic radius (flow/wetted perimeter) in flow through a sand bed is taken to be the pore volume/grain surface area:

$$R_h = \frac{\text{pore volume}}{\text{grain surface area}} = \frac{\varepsilon}{(1-\varepsilon)\frac{A}{V}}$$

where ε is the porosity, i.e., the ratio of pore volume/total volume, V is the volume of an individual grain and A is the surface area of an individual grain. Equation (6.3) may therefore be adapted to the following form for flow through a uniform sand bed:

$$\frac{\Delta h}{\Delta l} = C\frac{\mu}{\rho}\cdot\frac{v_c}{g}\cdot\frac{(1-\varepsilon)^2}{\varepsilon^2}\left(\frac{A}{V}\right)^2 \qquad (6.4)$$

Expressing the capillary velocity v_c in terms of the approach velocity v, using the correlation $v_c = v/\varepsilon$, yields the Kozeny equation:

$$\frac{\Delta h}{\Delta l} = C\frac{\mu}{\rho}\cdot\frac{v}{g}\cdot\frac{(1-\varepsilon)^2}{\varepsilon^3}\left(\frac{A}{V}\right)^2 \qquad (6.5)$$

where the empirical coefficient C has a value of about 5.

For a spherical sand grain of diameter d_s, A/V has the value $6/d_s$. For a non-spherical sand grain passing through a sieve aperture size d_s, A/V can be expressed as 6/ψds, where ψ is a sphericity factor, being the ratio of the surface area of the equivalent volume sphere to the actual surface area. Sphericity and bed porosity values for various sand grain shapes are given in Table 6.2.

Table 6.2 Typical porosity and sphericity values for filter sands (Fair et al., 1968)

Grain description	Sphericity (ψ)	Porosity (ε)
Spherical	1.00	0.38
Rounded	0.98	0.38
Worn	0.94	0.39
Sharp	0.81	0.40
Angular	0.78	0.43
Crushed	0.70	0.48

Assuming a value of 5 for the coefficient C and a sand bed of uniform size spherical sand grains of diameter d_s, the Kozeny equation can be written as follows:

$$\frac{h}{l} = 180\frac{\nu}{g}\frac{(1-\varepsilon)^2}{\varepsilon^3}\frac{v}{(\psi d_s)^2} \qquad (6.6)$$

where the kinematic viscosity $\nu = \mu/\rho$. Equation (6.6) is valid while conditions remain laminar, i.e. when the Reynolds number R_e is less than 10, where $R_e = vd_s/\nu$. For example, at a filtration rate of 12 m h-1, a sand grain size of 1 mm and a water temperature of 20 ^oC, the filtration Re vale is calculated to be 3.33. As this set of data relates to conditions at the upper limit region of filtration practice, it would be very unusual for flow to be outside the laminar range in sand filtration practice.

In filter back-washing at high rates the upflow through the sand bed may be outside the laminar range. The following empirical equation is suggested for flow in the first part of the transition range, $10<R_e<60$:

$$\frac{h}{l} = 130\frac{\nu^{0.8}}{g}\frac{(1-\varepsilon)^{1.8}}{\varepsilon^3}\frac{v^{1.2}}{(\psi d_s)^{1.8}} \qquad (6.7)$$

The maximum head loss in back-washing equals the submerged weight of the sand bed and is reached at incipient fluidization:

$$\left(\frac{h}{l}\right)_f = (1-\varepsilon)\left(\frac{\rho_s}{\rho} - 1\right) \qquad (6.8)$$

For typical values of $\varepsilon = 0.4$ and $\rho_s/\rho = 2.6$, $(h/l)_f$ has a value of 0.96, indicating that the maximum head loss in upflow through a typical sand bed is approximately equal to the bed depth. Equation (6.8) can be combined with equation (6.6) or (6.7), as appropriate, to compute the minimum upflow velocity, v_{mf}, at which fluidisation will occur. An increase in upflow velocity beyond this value will not lead to an increase in head loss but will cause the sand bed to expand, increasing its porosity. Wen and Yu (1966) developed the following empirical equation for the expanded bed porosity, in terms of Reynolds number and Galileo number:

$$\varepsilon^{4.7}G_a = 18R_e + 2.7R_e^{1.687} \qquad (6.9)$$

where G_a is the Galileo number

$$G_a = \frac{d_s^3\rho(\rho_s - \rho)g}{\mu^2}$$

Wen and Yu also made the following observations:

(1) In multi-sized particle systems, the average diameter may be defined by

$$\frac{1}{d_{ave}} = \sum_{i=1}^{n}\frac{X_i}{d_i} \qquad (6.10)$$

(2) In sand beds of mixed sizes, intermixing occurs if the ratio of sizes is less than 1.3:1, and stratification occurs if it is greater than 3:1.

6.5.1 Head loss in a stratified filter

In practical applications filter sands are not of single size but extend over a size range, the size distribution within which can be measured by sieve analysis. The resulting grading curve can be used to approximate the grain size profile within a sand bed stratified by back-washing as consisting of a layered bed, as shown in Fig 6.5.

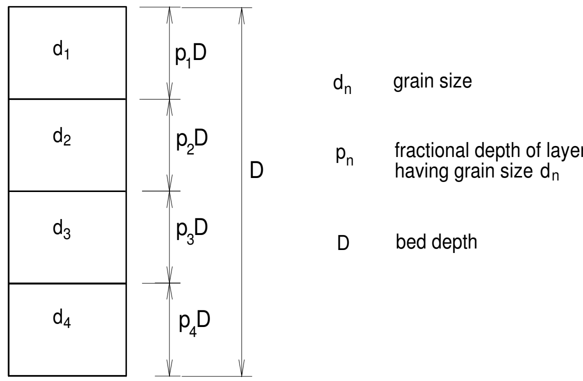

Fig 6.5 Layer model of filter bed stratified by back-washing

Equation (6.6) can be applied to each layer, assuming a uniform particle size within a layer:

$$h_n = 180 \frac{\nu}{g} \frac{(1-\varepsilon)^2}{\varepsilon^3} \frac{v}{(\psi d_n)^2} p_n D \tag{6.11}$$

where h_n is the head loss across layer n. The total head loss, H, is

$$H = 180 \frac{\nu}{g} \frac{(1-\varepsilon)^2}{\varepsilon^3} vD \sum_{i=1}^{n} \frac{p_i}{(\psi d_i)^2} \tag{6.12}$$

This estimation of head loss in a stratified bed assumes that ε and ψ are constant and independent of grain size.

6.6 REMOVAL MECHANISMS IN RAPID FILTRATION

While one might intuitively expect mechanical straining to be a major mechanism of solids separation in rapid filtration, this is not the case. Indeed, such a separation mechanism would rapidly lead to a blinding of the filter surface and a consequent very rapid increase in head loss. In fact, much of the particulate matter removed in rapid filtration is very much smaller than the average pore size. The separation of such material results from a combination of transport processes, which bring particles into contact with grain surfaces, and attractive forces, which secure their attachment. The transport mechanisms may include diffusion, hydrodynamic forces, interception and sedimentation, while the attachment mechanisms involve physico-chemical forces.

Diffusion or random Brownian motion is of particular significance for very small particles but has a negligible transport influence compared to gravity and hydrodynamic forces for larger particles. Sedimentation is considered to be significant because of the very large upward-facing horizontal surface area within a filter onto which particles may settle. Removal by sedimentation is a function of particle size and particle density and the velocity distribution within the filter pores.

Inertial forces do not constitute a very significant removal mechanism in rapid filtration because of the low fluid velocities involved and the retarding effect of viscosity.

Surface attachment influences include the universal force of mass attraction (van der Waals force) and electrostatic forces which depend on the sign and magnitude of the charges on the particles and on the sand grains (Coulomb force). These forces are only of significance when particles and sand grains are brought close together by one or more of the transport mechanisms noted above. The van der Waals force decreases in proportion to the sixth power of the distance between mass centres, while the Coulomb forces decrease in proportion to the second power of the distance between surface. Particle attachment is also promoted by the development on the sand grains of a sticky gelatinous layer of deposited material.

6.7 PROCESS KINETICS

The rate of decrease of suspension concentration *c* with distance *y* from the filter bed surface is considered to be proportional to the local concentration in accordance with the following relation:

$$-\frac{dc}{dy} = \lambda c \qquad (6.13)$$

where l is the filter coefficient (m-1) and is dependent on grain size, porosity and filtration rate. Integration of equation (6.13) subject to the boundary condition of $c=c_0$ at y=0 gives

$$c = c_0 e^{-\lambda y} \qquad (6.14)$$

where c is the concentration (mg l^{-1}) of suspended material in the filtrate water at depth y below the filter surface.

The rate of deposition in the filter pores corresponds to the rate of removal from the water. Hence, a mass balance equation may be written for a filter volume of unit plan area and thickness *dy* as follows:

Inflow – outflow = deposit

$$v \cdot c \cdot dt - v\left(c + \frac{\partial c}{\partial y} dy\right) dt = \frac{\partial \sigma}{\partial t} \cdot dt \cdot dy$$

which simplifies to

$$-\frac{\partial c}{\partial y} = \frac{1}{v} \cdot \frac{\partial \sigma}{dt} \qquad (6.15)$$

where σ is the mass of deposit per unit volume of filter (specific deposit). Combining equations (6.14) and (6.15):

$$\frac{\partial \sigma}{\partial t} = v \cdot \lambda \cdot c_0 \cdot e^{-\lambda y}$$

Integration subject to the boundary condition σ = 0 at t = 0 yields

$$\sigma = v \cdot \lambda \cdot c_0 \cdot e^{-\lambda y} t \qquad (6.16)$$

Under practical filtration conditions, however, the value of λ is not constant, varying over the filter depth and also varying with time. Its variation with filter depth arises from grain size stratification by back-washing which brings the finer material to the upper regions of the filter. Thus in a clean bed λ is likely to increase with distance from the bed surface. The increase in filter deposited material with time also causes a decrease in λ, leading to a deeper penetration of deposited material as filtration proceeds.

While much effort has been devoted by many researchers to the development of filtration theory (Camp, 1964; Mints, 1966; Ives, 1969; Burganos et al., 1991 and many others), the development of filtration technology has been largely based on empirical methods and experimentation. This is probably a reflection of the complex nature of the particle removal mechanisms under practical filtration conditions.

6.8 FILTER MEDIUM SELECTION

Silica sand is invariably used as the filter medium in rapid filters. It should be as near single size as possible, otherwise the bed becomes hydraulically stratified in back-washing with a grain size gradient from fine at the top to course at the bottom – the opposite to that required for efficient use of the full depth of the bed. A grain size gradient of course at the top to fine at the bottom can be maintained by using media layers of progressively increasing density, e.g. anthracite on silica sand on garnet sand, as exemplified in Table 6.3. The density difference between these materials is sufficient to prevent intermixing during back-washing, provided the size difference is not too great. However, most filters contain sand only because of its ready availability and lower cost than materials such as anthracite and magnetite.

Table 6.3 Examples of multi-media filters (Hall & Hyde, 1992)

Type	Position in bed	Medium	Depth of layer (m)	Specific gravity	Effective Size (mm)
Dual-media	Top	Anthracite	0.2	1.5	1.5
	Bottom	Silica sand	0.6	2.6	0.6
Triple-media	Top	Anthracite	0.2	1.5	1.5
	Middle	Silica sand	0.4	2.6	0.6
	Bottom	Garnet sand	0.2	4.2	0.4

The filter media size distribution may be fully specified, i.e. its grading curve may be required to be within specified upper and lower limits, as illustrated in Fig 6.6. Alternatively, the size distribution may be specified in terms of the Hazen effective size and uniformity coefficient. The effective size is the sieve size that passes 10% by weight of the material (d_{10}); the uniformity coefficient is the ratio of the sieve size that passes 60% by weight of the material to the sieve size that passes 10% of the material (d_{60}/d_{10}).

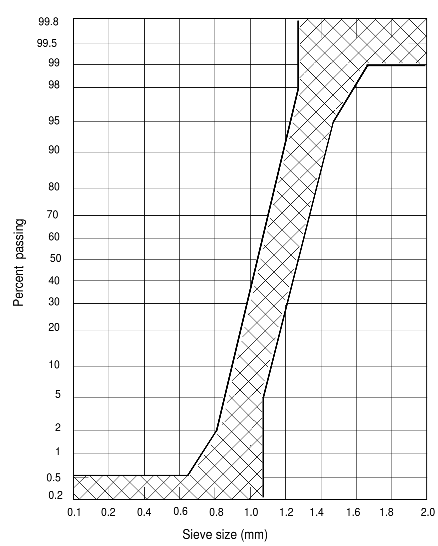

Fig 6.6 **Filter sand grading specification**
(d_{10} = 1 mm; d_{60}/d_{10} = 1.2)

Because the smaller grain sizes are of greater significance to filter performance, the ten percentile diameter was chosen as the effective size. The ten percentile by weight corresponds approximately to the fifty percentile by number. Ideally the uniformity coefficient should be as close to unity as possible and preferably less than 1.5. Obviously it is better to set full grading curve limits than to specify an effective size and uniformity coefficient only – the latter specification identifies only two points on the grading curve.

Selection of bed depth and sand size is frequently made on the basis of local experience, an effective size of 0.6 mm and a bed depth of 0.6 m being commonly used. Pilot plant tests provide the best basis for the determination of the design values for these parameters. Pilot units may be constructed in cylindrical tubes of 100-150 mm diameter. After a 'run-in' period, the filter performance is monitored by measuring the head loss at the end of a filter run of design duration T, while the residual solids concentration (or turbidity level) is measured after a greater duration, e.g. 1.1T. These measurements are carried out for various filter depths and the results are plotted in the form shown in Fig 6.7. The design depth is then based on a target residual suspended solids or turbidity value, as illustrated in the diagram. The use of a greater run-time for the filtrate quality measurement than for the head loss measurement allows the latter to be used for the control of the filter run duration with a built-in margin of safety in relation to filtrate quality.

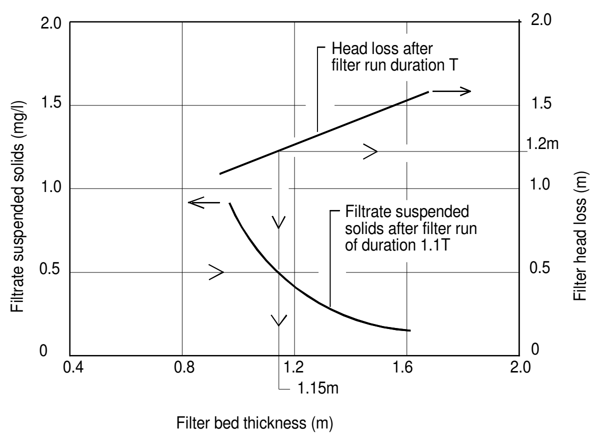

Fig 6.7 **Selection of filter bed thickness on the basis of pilot plant test results**

6.9 BACK-WASHING PRACTICE

Conventional rapid gravity and pressure filters are operated on a batch mode basis and are taken out of production for back-washing at the end of each filter run. Back-washing practice varies. The following are commonly used methods:

(1) Air-scour at an upflow rate of about 30 m h^{-1} for 3 min followed by water upflow at a rate of 20 m h^{-1} for 5 min.

(2) Up-wash with water only, at a rate of about 50 m h^{-1} for 5 min (used in the United States)

(3) Simultaneous air and water up-flow for 5 min, air at an up-flow rate of about 50 m h^{-1} and water at a rate of about 20 m h^{-1}, followed by water only at a rate of 20 m h^{-1} for 5 min.

Experimental comparison of these methods (Patterson, 1978) has shown that method (3), i.e. simultaneous use of air and water, provides the most efficient cleaning action.

The back-wash rate should be sufficient to achieve bed fluidisation, resulting in a bed expansion of 10 to 20%. The up-flow rate to achieve this is a function of grain size, grain specific gravity and water temperature. The following empirical expressions are a guideline to the minimum fluidisation rate for silica sand and anthracite media at a water temperature of 20 °C (rates at 5 °C would be about 75% of the 20 °C values, due to the increased water viscosity at the lower temperature):

Silica sand: $q_f = 40\ d_s^{1.4}$

Anthracite: $q_f = 8\ d_s^{1.6}$

where q_f is the upflow rate (m h^{-1}) and d_s is the effective grain size (mm).

The back-washing operation requires a relatively large water flow for a relatively short period (5-10 min) during which an individual filter unit is being washed. This can be provided by pumping from low-level storage or the appropriate volume may be stored at high level to provide a gravity back-wash supply.

The normal sequence of operations in back-washing is as follows:

(1) inflow to the filter is shut off.
(2) The standing water level on the filter is drawn down to near bed surface level.
(3) The air and back-wash valves are opened
(4) After the set duration of the back-washing process, the back-wash valves are closed.
(5) Filtration is re-started by opening the inflow control valve.

The back-wash water is collected in decanting channels set about 300 mm above the filter bed surface (sufficiently high so as not to interfere with bed expansion due to back-wash). The decanting channels discharge to waste, as shown in Fig 6.2.

While conventional filters are back-washed intermittently in the manner just described, innovative filter systems are being developed which permit continuous filtration and reduce back-wash water requirements (Boller, 1994).

6.10 CELLULAR SUB-DIVISION OF FILTRATION AREA

The required filter area A is determined by the required filter output and the design filtration rate, taking into account the variations in both of these parameters throughout the year. The required water output should include an allowance for the quantity of filtered water to be used for back-washing. The filter area should allow for at least one filter cell being out of use for repair or alteration.

The total filter area A may be divided into n individual units, each of area a in accordance with the relation:

$$a = \frac{A}{(n-1)} \cdot \frac{T}{(T-t_w)} \tag{6.17}$$

where T is the filter run duration and t_w is the filter wash duration. The factor (n-1) allows for one unit being out of use for repair or alteration; the ratio $T/(T-t_w)$ allows for downtime for back-washing. In general, the value of n should not be less than 3 and the value of a should not be less than 10 m^2. Individual units of very large plan area are undesirable because of the correspondingly large flows required for back-washing – units of plan area greater than 200 m^2 are rare.

6.11 CONTROL OF FILTRATION

The hydraulic control of filtration involves:

(1) equal sub-division of incoming flow among the filter units in operation at any time.
(2) Maintenance of a constant rate of flow through each filter unit for the full filter run duration

(3) Operation of the back-washing sequence

A filter run may be terminated when the filtrate quality has deteriorated below a set value, e.g. colour 10 ^{o}H, turbidity 0.5 NTU or when the head loss has exceeded a set value. Head loss is the preferred control parameter because of its convenience of measurement. A design value of terminal head loss, consistent with the achievement of the set quality target and which also avoids negative head development within the filter, is chosen. As illustrated in Fig 6.8, the development of a negative head can be avoided if the maximum head loss is limited to the depth of water above the filter bed surface.

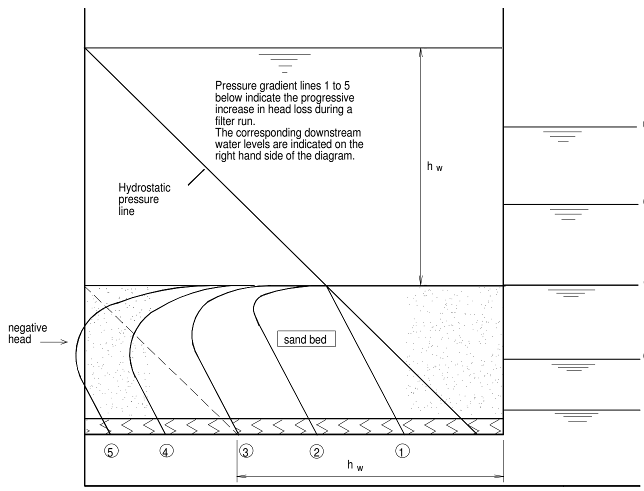

Fig 6.8 **Pressure gradient change during a filter run**

All filter control systems incorporate an automatic valve on the filtrate line, which gradually opens during a filter run thereby compensating for the increasing filter resistance, so as to keep the filtrate discharge rate constant. A common feed channel with a lateral discharge to each filter unit controlled by a weir or orifice is commonly used to sub-divide flow equally among filter units.

6.12 MEMBRANE FILTRATION

Membrane processes can be broadly classified in two groups on the basis of the nature of the driving force employed in their operation. The more widely used membranes are those that rely on an imposed pressure gradient to force water through the membrane, while retaining particulates and, in some cases solutes. The second type of membrane process uses an electrical potential gradient across the membrane to effect the selective migration of ions in a filtration process known as electrodialysis.

6.12.1 Pressure-driven membrane processes

Pressure-driven membrane processes can be categorised by the molecular weight or particle size cut-off of the membrane, as illustrated in Fig 1.2 and as summarised in Table 6.3.

Table 6.3 Classification of pressure-driven membrane processes

Membrane process	Size cut-off Range (µm)	Examples of materials separated
Microfiltration (MF)	0.05-1.5	Microbial cells, large colloids, small particles
Ultrafiltration (UF)	0.002-0.05	Macromolecules, viruses, colloids
Nanofiltration (NF)	0.0005-0.007	Viruses, humic acids, organic molecules, Ca^{2+}, Mg^{2+}
Reverse osmosis (RO)	0.0001-0.003	Aqueous salts, metal ions

MF and UF membranes are considered to effect species separation by a sieving mechanism and hence they remove particles larger than their cut-off size limit. The mode of action of both NF and RO membranes is considered to include both sieving and diffusion-controlled transport. NF and RO membranes are capable of separating both organic molecules and inorganic ions. NF and RO membranes are operated at high differential pressure (up to 100 bar) and have typical characteristic flux rates in the range 5-9 l m^{-2} h^{-1} bar^{-1}. UF membranes are operated at differential pressures ≥5 bar and achieve flux rates in the range 100-200 l m^{-2} h^{-1} bar^{-1}. MF membranes are operated at differential pressures ≤1 bar and achieve flux rates in the range 100-200 l m^{-2} h^{-1} bar^{-1}.

There has been ongoing development in synthetic polymeric membranes since the 1960s which has now reached the stage where membrane filtration offers a potential alternative to conventional potable water treatment processes such as chemical coagulation/sand filtration/disinfection. Membranes are marketed in module or cartridge housings, which are supplied in ready-to-use form. They are fitted with feed, permeate and concentrate connections, as illustrated in Fig 6.9.

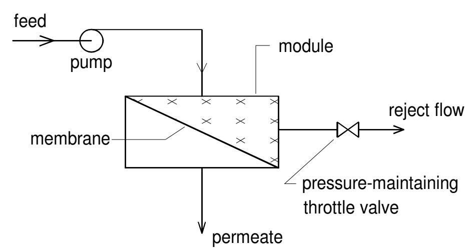

Fig 6.9 Schematic layout of membrane filtration module

In general, membranes may be operated in 'dead end' or in 'cross-flow' modes. In the dead-end mode of operation, the permeate is the only outflow from the module during a filter run and hence the removed particulate and solute species accumulate within the module. Cross-flow filtration, on the other hand, operates in a re-circulation mode with the feed introduced tangentially to the membrane surface. The tangential flow mode promotes the self-cleansing of the membrane surface with an associated reduction in membrane fouling. The suspension can be recycled until the required yield is achieved.

A variety of module configurations is available, including tubular, capillary, hollow fibre, spirally wound sheets and plate and frame. Since the permeate discharge per unit membrane area may be relatively low, modules are designed to maximise the membrane packing density, expressed as membrane filtration area per unit volume of module (m^2m^{-3}).

The tubular module consists of a tubular membrane sitting inside a porous sleeve, which in turn fits into a perforated pressure-tight tube between 12 mm and 24 mm diameter. Tubular membranes are normally operated in the cross-flow filtration mode.

Capillary modules consist of capillary tube bundles, each end of which is connected to a head plate. The internal capillary diameter is typically in the range 400-2500 μm and the packing density is of the order of 100 m^2m^{-3}. Capillary modules are available in both UF and MF ranges. Like tubular modules, they are normally operated in a cross-flow filtration mode.

Hollow fibre modules consist of hollow fibre bundles, usually bent in hairpin-like fashion, with the open ends cast into an epoxy resin head plate. The internal diameter of the fibres is typically in the range 10-40 μm and the packing density of the order of 10 000 m^2m^{-3}. Unlike the capillary modules, filtration is from the outside inwards, the permeate being discharged through the open fibre ends. Hollow fibre modules are available in the UF, NF and RO ranges.

The spirally wound modules consist of membrane sheets wound around a permeate-collecting tubular core. The rectangular membrane sheets are arranged in 'pockets', each of which consists of a highly porous sheet (permeate-side spacer) sandwiched between two membranes, which are glued together along three edges. The fourth edge of the pocket is attached to the collecting tube. Several such pockets are spirally wound around the collecting tube with a feed-side spacer placed between the pockets to form a so-called 'element'.

In plate and frame modules, the membranes are arranged in parallel flat sheets, with porous spacers, through which the permeate is discharged. The feed solution flows through flat rectangular channels between the membranes. Packing densities are in the range 100-400 m^2m^{-3}.

MF processes are used to remove turbidity and chlorine-resistant pathogens such as Cryptosporidium and Giardia. Based on work on a wide variety of membrane materials, Zahka and Grant (1991) concluded that absolute retention was obtained for particles that were 2 to 3 times larger than the manufacturer cut-off rating of the membrane. The average diameter the oocysts of Cryptosporidium parvum, the most frequently isolated cryptosporidial species, is 4.5-5μm. Giardia cysts are somewhat larger (Coop et al, 1998). Hence, membranes with a cut-off rating of 1.5μm or less should provide an effective barrier to the transmission of these pathogens. For example, the UK drinking Water Inspectorate (DWI, 2006) has approved membrane filtration capable of continuously removing particles greater than 1μm diameter as a means of complying with the UK Water Supply Regulations 2000 in respect of Cryptosporidium removal. Studies have shown that membrane filters with cut-off rating of about 0.2 μm or less provide an effective barrier to the transmission of bacteria and also up to 90% virus removal, even though the nominal pore size is much larger than the size of viruses (Cote et al, 1995).

MF processes are typically operated at a differential pressure of less than 1 bar and at flux rates up to 150 $l\ m^{-2}h^{-1}$.

UF is used in the food and process industries as well as finding increased application in water and wastewater treatment. UF membranes with a molecular weight cut-off of 100 000 were found to achieve a 6 logs rejection (i.e. below the detection limit) of seeded MS2 virus, while UF membranes with larger cut-off limits achieved partial virus removal (Jacangelo et al., 1995). UF hollow fibre membranes are being used in conjunction with powdered activated carbon for the direct production of potable water from surface water (Cornu at al., 1995). Membranes at the lower end of the UF cut-off range can be used to remove humic substances, including colour, from waters (Jensen and Thorsen, 1995).

NF and RO membranes have lower cut-off limits than UF membranes and are capable of rejecting inorganic as well as organic species. NF rejection of inorganics is determined by the composition of the water, the rejection of bivalent ions being significantly greater than the rejection of monovalent ions. RO membranes remove dissolved organics almost completely and also remove most of the inorganic salts. Hence, they are used for the production of potable water from brackish waters and seawater.

6.12.2 Membrane fouling

Membrane fouling, with a consequent reduction in specific flux, is recognised as a major limitation of the membrane filtration process. Fouling potential is a function of water quality and is also influenced by the membrane material. Membranes with hydrophilic characteristics are known to reduce particle adherence. The water quality parameters of relevance to fouling potential include turbidity, aluminium, iron, manganese, silica, hardness, pH, TOC and temperature.

Membrane fouling can comprise chemical precipitation (scaling), colloidal particle retention, organic matter deposition and biofilm growth.

Chemical precipitation occurs when the raw water is supersaturated with the precipitant chemical as, for example, waters that have a significant positive calcium carbonate precipitation potential. Chemical scale fouling can be eliminated by the chemical adjustment of the feed water composition to ensure its chemical stability (refer Chapter 11).

Colloidal fouling results from the deposition of relatively inert colloidal suspended solids, such as silica and silicates. Pre-treatment of the feed water by processes such as chemical coagulation, rapid gravity filtration or some form of cartridge filtration can greatly reduce particulate fouling.

Organic fouling results from the precipitation of the decay products derived from NOM, such as organic acids. Such deposits are easily removed at elevated pH.

Biofilm formation potential depends on the availability of assimilable organic matter in the raw water.

Pre-treatment of the feed water by processes such as chemical coagulation, rapid gravity filtration or some form of cartridge filtration can greatly reduce particulate fouling.

6.12.3 Membrane integrity

The maintenance of the integrity of membrane systems is of obvious importance where they are used as a barrier against the transmission of pathogens such as Cryptosporidium and Giardia. The system integrity can be breached by puncturing of the hollow fibres that are mechanically stressed during backwashing and also by leakage through mechanical seals as a result of wear or displacement due to transient pressures. The problem is aggravated by the physical configuration of membrane systems where installations typically have very large number of hollow fibres, each of which acts as an individual filter and also a large number of automatic valves and pipe connections.
Major leaks on large systems can be detected by on-line turbidity monitoring of total permeate. However, dilution effects inevitably mask smaller leaks at the individual module or fibre level. Integrity tests that can be used for UF and MF membranes include:

- Air tests
- Particle counting
- Hydraulic noise measurement.

Recent studies (Hoffman et al., 2004) in the Netherlands have shown fibre failure rates of approximately 0.1 per thousand in UF installations. Transient pressures caused by rapid valve switching and flow reversal in backwashing have been identified as the main causes of fibre breakage.

6.12.4.1 Electrodialysis

Electrodialysis (ED) is a membrane filtration process in which ionic species are removed under the driving force of an electric field. The water to be treated is pumped through a membrane stack which consists of alternately placed anion-permeable and cation-permeable membranes, as illustrated in Fig 6.10. Under the action of the applied electric field the anions migrate towards the cathode, resulting in the stripping of ions from alternate chambers and their concentration in the remaining chambers.

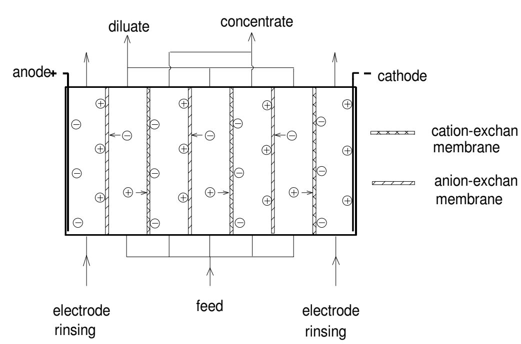

Fig 6.10 **General arrangement of ED module**

REFERENCES

Boller, M. (1994) *J. Water Supply Res. Tech.. – AQUA*, 43. No. 2, 65-75.
Burganos, V. N. Paraskeva, C. A. and Payatakes, A. C. (1991) *J. Colloid Interface Sci.*, 148, 167-181.
Camp, T. R. (1964) *J. San. Eng. Div., ASCE*, 90, No. Sa4, 1-30.
Casey, T. J. *Water and Wastewater Engineering Hydraulics*, Oxford Science Publications, Oxford University Press, Oxford.
Coop, R.L., Wright, S.E., and Casemore, D.P. (19xx) *Cryptosporidiosis*, Chapter 45 in Zoonoses, Eds. Palmer, Soulsby & Simpson, Oxford Medical Publications, 1998.
Cornu, S., Gelas, G. and Ansleme, C. (1995) *Proc. IWSA Workshop on Membranes in Drinking Water Production,* Paris.
Cote, P.Tazi-Pain, A. and Dard, S. *Membranes for Disinfection*: Potential and limitations, Proc. IWSA Conf., Paris, March 1995.
Darcy, H. (1856) *Les fontaines publiques de la Ville de Dijon*, Dalmont, Paris.
DWI (2006) *Cryptosporidium: Approval of membrane and other filtration systems for Cryptosporidium removal*, Drinking water Inspectorate, London, UK.
Graham, N. J. D. ed., (1988) *Slow Sand Filtration*, Ellis Horwood Ltd., Chichester, UK.
Hoffman, J.A., Ijpelaar, G.F., Heijman, S.G., Vrouwenvelder, J.S., Kruithof, J.C. and van der Meer, W.G. (2004) *Drinking water treatment in the Netherlands: outstanding and still ambitious*. Water Science and Technology: Water Supply, 4, No. 5-6, pp253-262.
Ives, K. J. (1969) *Proc. IWSA Conf.,* Vienna.
Jacangelo, J., Adham, S. and Laine, J. M. (1995) *Proc. IWSA Workshop on Membranes in Drinking Water Production,* Paris.
Jensen, K. and Thorsen, T. (1995) *Proc. IWSA Workshop on Membranes in Drinking Water Production,* Paris.
Kozeny, J. (1927) *Sityungsberichte der Wiener Akademic der Wissenschaften,* 136, Pt 2a, 271
Mints, D. M. (1966) *Proc. IWSA Conf.,* Barcelona.
Patterson, P. (1978) *MSc Thesis,* Queen's University, Belfast.
Rachwal, A. j., Bauer, M. J. and West, J. T. (1988) Advanced techniques for upgrading large scale slow sand filters, Slow sand Filtration, ed. Graham, N., Ellis Horwood Ltd., Chichester, UK.
Rautenbach, R. and Albrecht, A. (1989) *Membrane Processes*, John Wiley & Sons, New York.
Wen, C. Y. and Yu, Y. H. (1966) *Mechanics of Fluidisation*, Chem. Eng. Progress, Symposium Series, 62:62:100.
WRc (1992) *Water Treatment Processes and Practices*, eds Hall, T. and Hyde, R. A., WRc Swindon, UK.
Zahka, J.G. and Grant, D.C. *Predicting the performance efficiency of membrane filters in process liquid based on their pore size ratings,* Micrcontamination, pp23-29, December 1991.

Related reading

ASCE/AWWA (1990) *Water Treatment Plant Design*, 2nd. edn., McGraw-Hill Book Co., New York.
Lorch, W. (1987) *Handbook of Water Purification*, Ellis Horwodd Ltd., Chichester, UK.