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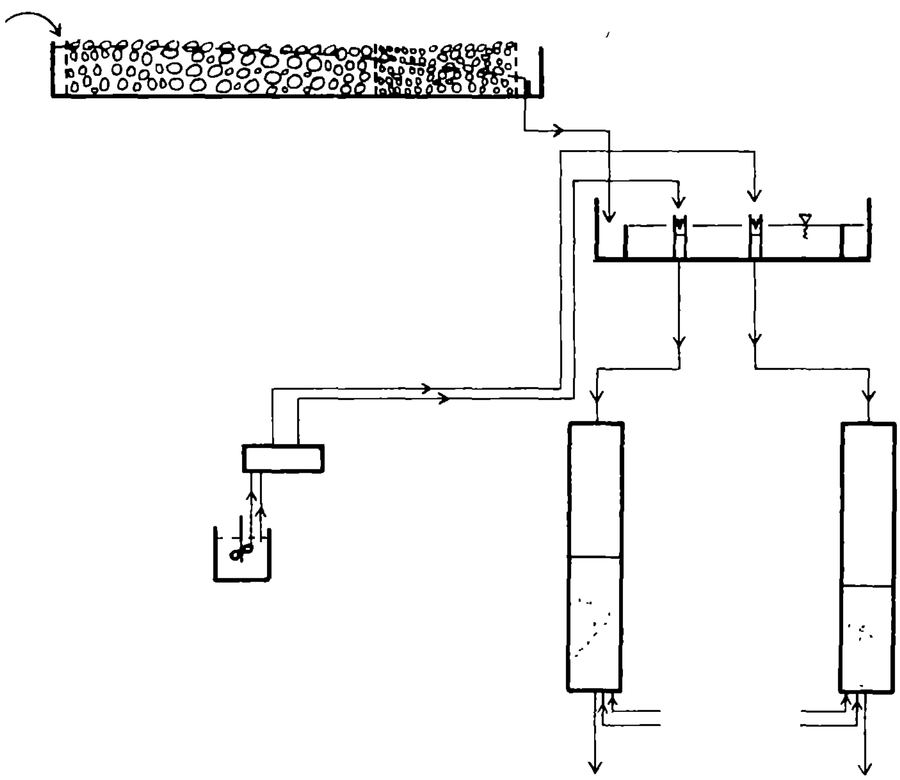
Tilahun Gebre-Tsadik

Direct Filtration with Horizontal Roughing Filter as Pretreatment



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DIRECT FILTRATION WITH HORIZONTAL ROUGHING FILTER AS PRETREATMENT

by

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Thesis submitted to the department of civil engineering, Tampere University of Technology in partial fulfillment of the requirements for the degree of Master of Science in Engineering

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Dedicated to:

Ato SEYOUM SERAWITU,
Woizero TAFFESECH SHIFERAW
and
my beloved wife GENET HAILEMICHAEL,

whose continued encouragement has been a strong source of inspiration in overcoming the formidable challenges that arose out of my stay abroad for the thesis work.

DIRECT FILTRATION WITH HORIZONTAL ROUGHING FILTER AS PRETREATMENT

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ABSTRACT

Investigations were carried out to study the feasibility of direct filtration with horizontal roughing filter as pretreatment. With the roughing filter, tests were run at filtration rates of 5, 10 and 15 m/h. Suspended solids concentrations of up to 6000 mg/l were investigated. The tests performed on the down-flow rapid filters were also carried out at constant filtration rates of 5 m/h, 10 m/h and 15 m/h; turbidities of up to 220 NTU were handled. Dosages of alum as low as 2,5 mg/l were tested and found satisfactory. Further, an attempt has been made to provide a rough guide for the design of the horizontal roughing filter including prediction of the filter length. The results of the investigations as well as the background literature review are of both practical and theoretical value. Direct filtration with horizontal roughing filter as pretreatment is proposed as a viable treatment method for the clarification of turbid surface waters.

1. INTRODUCTION

Surface waters in tropical countries are in most cases characterized by some pollution, high turbidity and silting. The turbidity is mainly due to suspended solids and dissolved colloids originating from erroded land surfaces. Uneven rainfall distribution, deforestation and land cultivation methods which promote soil errosion being the main causes.

Understandably, the concentration of suspended solids in waters from such sources varies according to the intensity of the reasonal rainfalls. During the rainy seasons the concentration from the heavy runoff is the highest. On the other hand, the waters are usually relatively clean during the dry seasons.

The conventional purification practice for the removal of turbidity and colloidaly dispersed solids from water predominantly consists of coagulation, flocculation, sedimentation and rapid filtration of the water to be treated through a bed of granular media.

With relatively low turbidity waters (up to 200 turbidity units) the flocculation and sedimentation steps could be eliminated from the conventional treatment process and direct filtration applied (Culp, 1977; Voss and Gross, 1981). The key to success in the direct filtration process is the filter itself because of its use both as a flocculation reactor and a floc storage reservoir (Shea et al, 1971).

The savings made from the elimination of units for flocculation and sedimentation together with the corresponding mechanical accessories as well as reduced chemical requirements, to mention but a few, make direct filtration an attractive alternative to the conventional treatment method for urban water supply needs. In brief, the problems of high construction, operation and maintainance costs could be appreciably reduced applying direct filtration for waters of low turbidity.

The objective of this thesis is to study direct filtration with horizontal coarse gravel roughing filter as pre-treatment. So far horizontal roughing filters have been studied in connection with the reduction of turbidity of waters for subsequent slow sand filtration (Wegelin, 1981; Riti, 1981; Voss and Gross, 1981; Mbwette, 1983). If the horizontal coarse gravel roughing filters could produce water of constantly low turbidity especially during the few rainy seasons, it is evident that the application of direct filtration would still be advantageous for urban water supply treatment needs in the tropical countries as compared to the conventional treatment system.

In view of the foregoing discussion, the relevance and practical significance of the study to the advancement of the water treatment methods for the developing countries in the tropics can not be overemphasized.

2. MODES OF FILTRATION

Within the filtration operation two distinct modes of filtration can be observed - surface filtration and deep bed or volume filtration.

2.1 Surface filtration

Surface filtration may be characterized by the formation of a cake of suspended particles on the surface of the filter medium, due to blocking of the pores in the uppermost layer of the filter. This is predominantly a physical type of removal mechanism - straining, although some small particles may be removed by adhesion to the surface cake. The removal efficiency is mainly constant during the time of filtration. Surface filtration occurs when certain conditions prevail with respect to particle characteristics and media size, flow rate and influent concentrations (Hedberg, 1976).

2.2 Deep bed filtration (volume filtration)

Water filters are deliberately designed as deep bed filters. This means that the retention of particles from suspension must be within the filter pores in the depth of filter media. Therefore, straining is precluded (Ives, 1982), particularly where the suspension particles are larger than the pore openings forming a mat or a cake at the inlet surface. Consequently, media design and operation must avoid this possibility, either by pretreatment of the suspension or by adjusting inlet face pore size and flow rate to allow particles to penetrate into the filter material. In contrast to cake filtration which is commonly encountered in chemical process industries and used to separate particles from relatively dense suspensions with solid volume fractions exceeding say, 2000 ppm, deep bed filtration is the most effective and economical in treating large quantities of liquids containing relatively low solid volume fractions (below 500 ppm) of particles with fine or colloidal size (less than 30 μm) (Tien et al, 1979).

The removal in the filter material of the particles in the suspension to be filtered, entails a change of quality of the filtrate as well as inducing increased head loss of the media due to clogging in the filter pores. According to Ives (1982) a linear head loss development indicates true deep bed filtration; if it curves upwards with time there is probably inlet surface deposit.

3. MECHANISMS OF FILTRATION

The flow of suspensions through porous media is a very complex phenomenon due to the diversity of the mechanisms involved. In the literature several factors which may play an important role in filtration have been discussed. The predominant mechanisms depend on the physical and chemical characteristics of the suspension and the medium, the rate of filtration and the chemical characteristics of the water.

Most researchers agree that filtration of suspended and colloidal particles from water involves two separate and distinct steps (O'Melia and Stumm, 1967; Yao et al, 1971; Hedberg, 1976; Bratby, 1980):

a) The transport of the suspended particles to the immediate vicinity of the filter grains.

b) The attachment of these particles to the filter grains or to another particle which has previously been deposited in the bed.

According to Ives (1982) and Bauman (1982) a third step - the detachment step is also possible.

3.1 Transport mechanisms

As the flow of water is laminar during filtration (Ives, 1982), that is the streamlines are ordered and do not cross and mix, forces must act on the particles in the streamlines to move them to the filter grain surfaces. The transport mechanisms which bring the small particles from the bulk of the fluid within the interstices alose to the surface of the media include interception, sedimentation and Brownian diffusion.

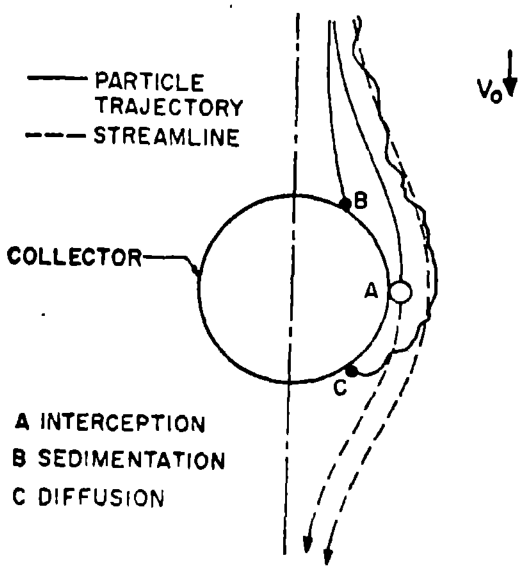


Figure 1. Basic transport mechanisms in water filtration (Yao et al, 1971)

It is evident from the literature that the relative importance of the transport mechanisms depends upon several physical parameters, the most significant of which has been found to be the size of the particles to be removed from suspension. Bauman (1982) refers to the work of Yao et al (1971) who reportedly found that there exists a size of suspended solids for which removal efficiency is minimum. This critical suspended solids size is about 1 μm. For suspended solids larger than 1 μm, removal is enhanced by transport mechanisms of sedimentation and/or interception, i.e. gravity forces. For suspended solids smaller than 1 μm, removal efficiency increases with decreasing particle size. Transport is made possible by the increasing effects of diffusion forces as particle size decreases. Thus, the effects of the applicable forces on particle transport are shown in figure 2.

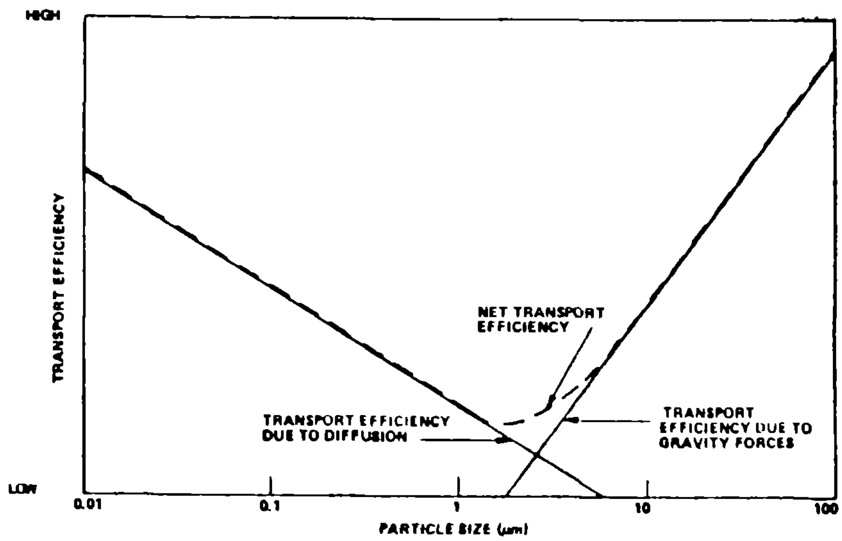


Figure 2. Relationship between transport and particle size (Bauman, 1982)

The most important conclusion that can be drawn today concerning the the transport mechanism is, perhaps, its relative insignificance in filter design. The reason for this is that the transport mechanisms involved in filtration are sufficient to do their job even for particles that have minimum transport ability (Adin et al, 1979). Kavanaugh et al (1978) also explicitly state "contrary to theoretical models of filter removal efficiency as a function of particle size, it has been shown that filters can remove particles in the fine size fraction (0,5 - 20μm) with efficiencies comparable to that for coarses particulates provided that the particles are destabilized with appropriate chemical pretreatment prior to filtration". Hence, control efforts in filtration might be made more fruitful and easier by viewing attachment as the major factor in thé filtration process.

3.2 Attachment mechanisms

As the particle approaches the surface of the medium, or previously deposited particles on the medium, an attachment mechanism is required to retain it. Attachment of particles to the media surface has been generally attributed to physicochemical and molecular forces. Ives (1982) writes that the attachment mechanisms exert their influence on the particle at distances of less than 1 μm.

The attachment mechanisms may be classified according to two models. The classic "double-layer model" is based on an interaction between the electrostatic repulsive forces and Van der Waal's attractive forces. The "bridging model" explains effects resulting from chemical bonding and bridging of suspension particles and medium through their reaction with coagulants and/or coagulant aids (Adin et al, 1979).

3.21 The double-layer model

When a colloidal particle is immersed in a solution, electrical charges will develope at the particle - water interface. The origin of these charges may be due to the dissociation of the ionizable groups of the colloid itself or to the adsorption of low - molecular - weight ions onto its surface (Committee Report, 1971). As a result of this charge development a charge balance must be established in the vicinity of the colloidal particle to fulfill the requirement of electro-neutrality. The arrangement of the charge balance is explained by the formation of the Stern-Gouy electrical double layer around the colloidal particle (Committee Report, 1971). The structure of the Stern-Gouy double layer and the corresponding potentials is presented in figure 3. Ions with the same charge as the particle are rare near the particle surface but gradually increase in number as the distance from the particle is increasing. Counter ions (ions of opposite charge to the particle) predominate near the particle surface and gradually decrease in number and concentration with increasing distance.

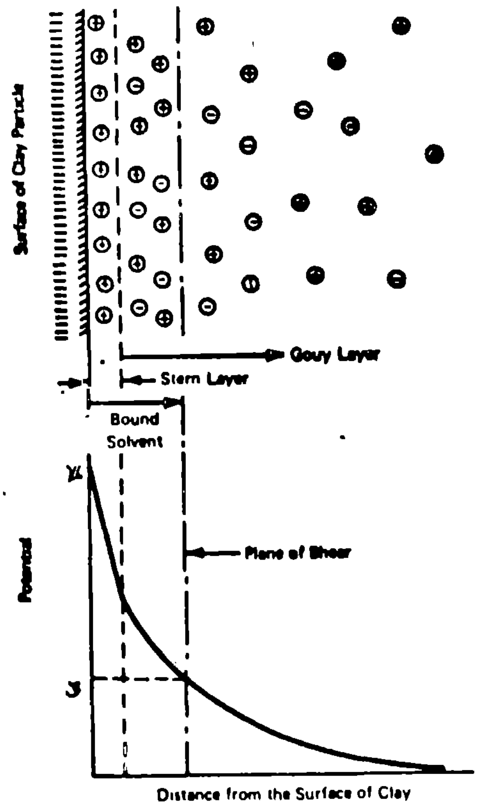


Figure 3. Structure of the Stern-Gouy double layer and corresponding potential (Committee Report, 1971)

Thus at the solid - liquid interface a tightly held layer of ions of opposite charge termed the "stationary layer" and the second, more loosely bound layer of ions termed the "diffuse layer" are produced. This double layer exerts a repulsive potential between similar particles in an aqueous suspension. The magnitude of this potential and the distance over which it acts are significantly affected by the chemical composition (ionic concentration) of the aqueous phase. Depending upon the different types of counter ions involved in the colloidal systems, the repulsive zeta-potential of the particle can be reduced by the compression of the double layer due to the incorporation of simple unhydrolysing counter ions (like Na^+, Ca^{++}) into the diffuse layer. This means that particles can come closer to each other as well as the filter grain surface more freely. Eventually, Van der Waal's attractive forces which vary inversely as the seventh power of the distance of separation (Craft, 1966) predominate. Thus removal of suspended particles is enhanced when the electrostatic repelling forces are at a minimum. However, the significance of electrokinetic effects as well as Van der Waal's forces have been questioned (Ives, 1964; O'Melia and Crapps, 1964). Figure 4 illustrates the forces acting on colloids and compression of the double layer.

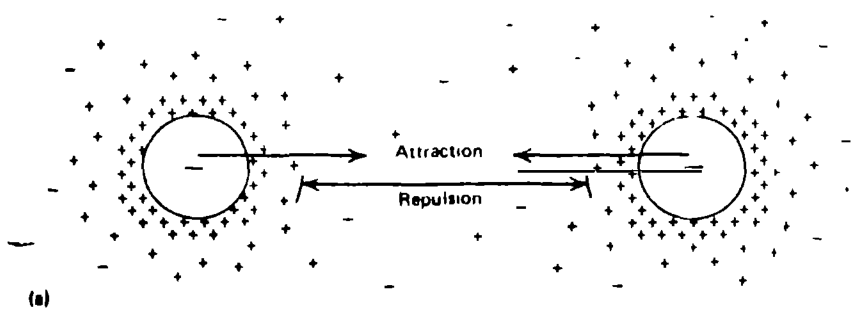


Figure 4 a. Electrokinetic repulsive and Van der Waal's attractive forces

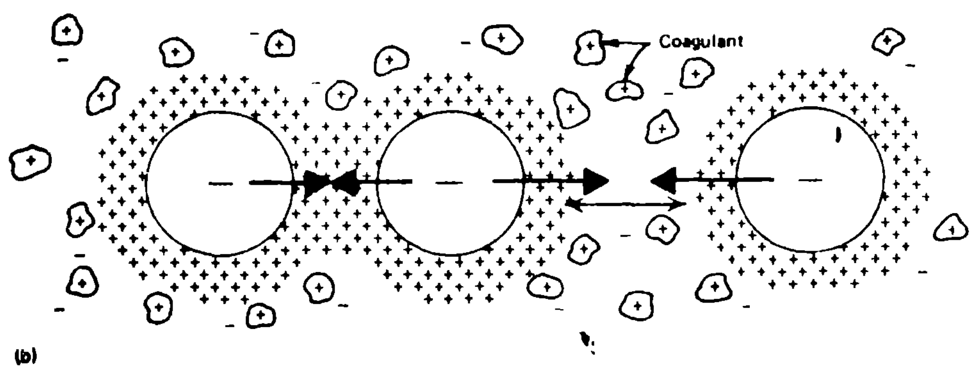


Figure 4 b. Compression of the double layer by coagulant addition

3.22 The bridging model

The repulsive zeta potential of particles in an aquous environment can also be reduced by the specific absorption of the counter ions onto particle surface. The action of conventional flocculants (aluminium and iron salts) is due primarily to their hydrolysates which are polymer chains with good adsorption properties that can form structures as a result of bonding. Flocculation with the aid of these materials takes place in two steps: neutralization of the particles negative charge by the positive hydroxide and formation of flocs by bridging between the particles as well as the grain surface as a result of the polymer chain adsorption (Adin et al, 1979). Flocculation with the aid of synthetic polymers occurs as adsorption of the polymer on the surface of the particles and bridging between them. The bridging in the case of synthetic polymers results in the formation of large and strong three dimensional structures.

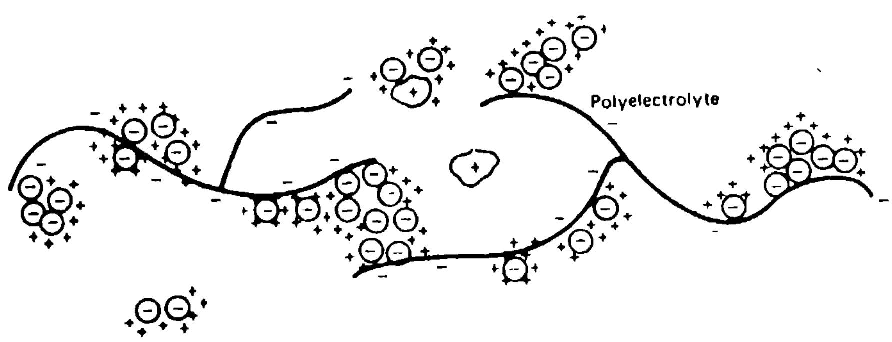


Figure 5. Agglomeration by specific adsorption and bridging

The attachment step is thus analogous to that of destabilization in the coagulation process (Bratby, 1980). With consideration of the foregoing the influence and destabilization of suspended material may readily be appreciated. Furthermore, the lag of working in period of the filter after backwashing is also understood.

3.3 Detachment mechanisms

There is controversy concerning the detachment of particles from the filter media during filtration. This is commonly referred to as the Mintz-Ives controversy. The disagreement centered around one of the fundamental theories of filtration i.e.

> a filter layer ultimately reaches a stage where it is no longer effectively clarifying the suspension and the concentration of suspension leaving the layer equals the concentration entering the layer - the so called equilibrium or saturation condition. (Anonymous, 1976)

According to Mintz

a) the rate at which particles are deposited on the filter grains remains constant, in terms of efficiency, throughout the filtration process, even at equilibrium;

b) particles, when they have been deposited on the filter grains, are subject to detachment by the flow, back into suspension at a rate proportional to the quantity of deposit on the grains.

Mintz's theory is consequently known as the "deposition and scour" hypothesis. The deposition rate, invariant with time, was attributed to the fundamental characteristic of filtration depending on the suspension, rate of flow and nature of filter grains, but not on the quantity of deposit present. The scour rate was attributed to the narrowing of the filter pores caused by the deposits, which locally increases the fluid velocity. This increases the fluid Shear stress on the deposits causing them to detach and be re-entrained in the flow. At equilibrium the rate of scour equals the rate of deposition, so the concentration of suspension entering and leaving any filter layer is unchanged. However, the particles which emerge at equilibrium are not those that enetered, but are scoured particles which had previously been deposited.

On the other hand Ives contended that the rate at which particles are deposited on the filter grains varies, in terms of efficiency, during the filtration process (first increasing, then decreasing, but this is a secondary argument) ultimately to become zero at equilibrium. Ives attributed this variable efficiency to geometric changes in the filter pores, due to the deposited particles, which changed the amount of surface available for deposition, which changed the flow pattern, and which locally increased the pore velocity. At equilibrium state a condition of no retention prevails because reduced surface area and high velocities sweep particles through pores before they can attach to grains or existing deposits. Consequently, the concentration of suspension entering and leaving any filter layer is unchanged, and the particles which emerge are those which entered the layer.

The disagreement between the two groups of research workers concerning the role of detachment seems to have been resolved. The geometric changes in deposits and scouring of deposited material have both been observed experimentally by those opposing the particular concepts (Anonymous, 1976). What is more Ives (1982) accepts the detachment phenomenon to some extent by stating that the effects of detachment mechanisms can be observed if there is an increase in filter flow rate, particularly during transients caused by control valves and that the presence of polyelectrolytes can reduce this effect.

4. FILTRATION VARIABLES AND REMOVAL EFFICIENCY

4.1 Particle size

It was noted earlier that the removal efficiency of a filter bed depends on the size of the particles being filtered. A critical particle size exists in the region of 1 μm. This particle size has the lowest opportunity for contact with the filter media and subsequent removal from suspension. Smaller particles are effectively transported by Brownian diffusion, larger ones by interception and settling. O'Melia and Ali (1978) point out that this effect of particle size on removal continues into the ripening period, i.e. one-micron particles show the poorest removal throughout the period of effective filtration.

4.2 Influent concentration

The removal efficiency of a clean filter is theoretically independent of influent concentration. As filtration proceeds removal efficiency improves with increasing influent concentration during the ripening period because retained particles act as collectors for other suspended particles. As expected low concentrations produce low head losses and low removal efficiencies. This is because removal by a packed bed filter can depend on the number of retained particles which act as collectors. When the influent concentration is low the rate at which new collectors accumulate in the bed is also low. The ripening process is lengthened but the removal efficiency is impaired. This indicates that filters treating low turbidity waters should be deep, while those treating waters with high suspended solids concentrations can be more shallow. Filters treating low turbidity waters, such as those operating in the direct filtration mode, must rely on the filter media to provide collectors. Those treating more concentrated suspensions remove solids primarily by contacts with previously retained particles. (O'Melia and Ali, 1978)

4.3 Media size

Concerning the significance of the media size Hedberg (1976) reveals that most workers seem to agree that the removal efficiency is improved in finer media. O'Melia and Ali (1978) point out that the effect on removal efficiency is less significant than those noted for suspended particle size and concentration. They indicate that the role of media size in filtration may be overstated both in the literature and in practice. According to the authors media size can easily be measured and readily controlled, but it is not of major importance.

4.4 Filtration velocity

The significance of an increased filtration rate v implies in general a decreasing removal efficiency (Conely and Hansen, 1982). However, the quantitative form of this dependence has not found an accepted expression. Hedberg (1976) suggests that the removal efficiency may vary with v^0 to v^{-4}.

4.5 Bed depth and filtration time

The effects of bed depth and filtration time have been studied by O'Melia and Ali (1978). Their results (figure 6) show that removal efficiency and head loss both increase significantly with time, as would be expected, but their distributuion with time is noteworthy. Removal is distributed throughout the bed at the onset of filtration but becomes localized in the upper region of the bed as filtration proceeds. Head loss follows a similar distribution.

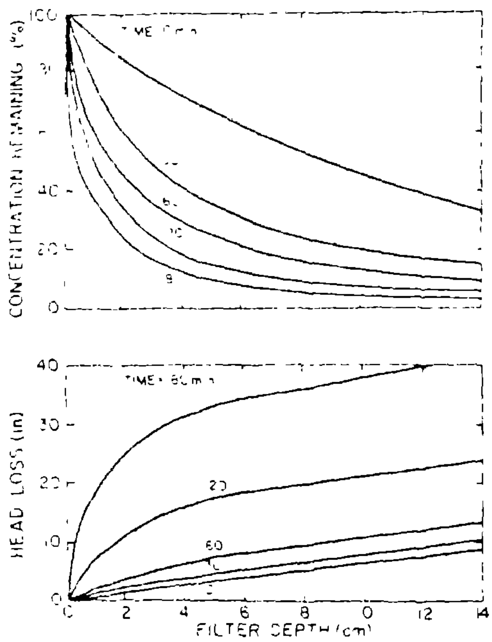


Figure 6. Variation of removal efficiency and head loss with time and depth

4.6 Temperature

A higher temperature will reduce the viscosity of the water to be filtered thereby facilitating removal. Hence in general the efficiencies are higher at high temperatures. According to Hedberg (1976) the removal efficiency may vary with T^{0} to T^{-2} due to temperature changes.

4.7 Chemical characteristics

In addition to the above variables the removal efficiency is also affected by the chemical characteristics of the suspended particles, the media as well as the water. Although several investigations have been carried out in which the importance of the pH-value, ionic strength, ionic species has been studied no empirical or theoretical relationships have been suggested to account quantitatively for these effects (Hedberg, 1976).

The overall effect of a change in any one of the above filtration variables is to modify the role of the mechanisms of filtration. A changed role of the mechanisms of filtration affects the filtration efficiency either favourably or adversely.

4.8 Parameters for evaluating filterability

The suitability of a particular type of raw water for treatment by filtration is commonly characterised with the parameters of turbidity and suspended solids. In the case of direct filtration colour also comes in as an additional parameter (Culp, 1977).

4.81 Turbidity and suspended solids

Turbidity may be identified as the lack of clarity of water. Suspended organic and inorganic matter gives rise to turbidity. Hence, the removal of turbidity from water involves the removal of a wide variety of polluting substances.

However, turbidity is not a direct measurement of the amount of suspended material in a given water but rather an arbitrary optical measurement based on the interference of light passing through the water. This is attributable to the fact that turbidity measurements are strongly influenced by the nature, size, concentration and refractive index of the particles in suspension. As a consequence there is no direct correlation between the amount of suspended material in a water sample and the turbidity of the sample.

The actual clogging of the filter occurs mainly due to suspended solids (O'Melia and Stumm, 1967). The evaluation of the acceptable quality of water by filtration is, therefore, undoubtedly more reliable with suspended solids as the parameter. Nonetheless, in practice, the determination of suspended solids content is too involving to the extent that turbidity is preferred as an indirect measure of the same. Using this approach, a better assessment can be ensured by establishing a correlation between turbidity and suspended solids concentration so that for any turbidity value measured, one is able to get the corresponding suspended solids concentration (Mbwette, 1983).

4.82 Filterability number

Judging the filterability of a given water by quantitative and qualitative measures such as suspended solids content, turbidity and colour only may be misleading. This is because the filterability of suspensions depends to an appreciable extent on the behaviour of the suspended particles in a filter media. The behaviour of such particles is influenced by the characteristics of the media, the size of the particles being filtered, influent concentration, filtration rates and temperature. It is also influenced by the chemical characteristics of the suspended particles as well as the media. Thus, the prediction of the filterability of a particular type of raw water by a filtration process involves two interactive elements: the suspension to be filtered and the filter. The concept of a filterability number evolved out of these considerations (Ives, 1978).

The apparates developed by Ives (1978) for determination of the filterability number is a simple small scale filter (figure 7).

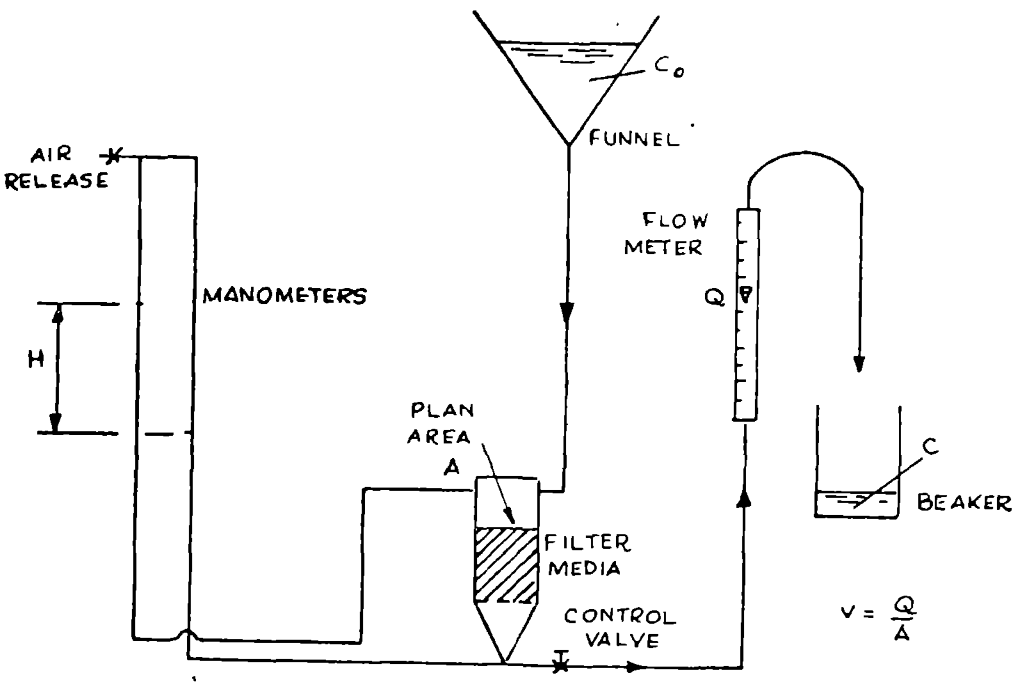


Figure 7. Diagram showing the principle of filterability apparatus (Ives, 1978)

The apparatus has facility for controlling and measuring flowrate (V), reading head loss (H) and sampling the inlet and filtrate suspensious for concentration measurements (C_0 and C).

The dimensionless filterability number is given by Ives (1982) as

$$F = \frac{HC}{VC_0 t} \tag{1}$$

No particular significance can be attached to the actual numerical values of F, but relative values of F indicate relative filterabilities. A minimum value of F from a number of tests would indicate an optimum filterability, even though nothing could be inferred from the numerical value attached to this minimum F.

Careful use of the filterability number test will enable fairly rapid screening of various pretreatment alternatives such as types and dosages of coagulants, coagulant aids, media and flow rates (Ives, 1978).

It is worth pointing out here that so far there is no standardized method for measuring filterability. Turbidity, suspended solids content and/or colour are still widely used for the same purpose.

5. DIRECT FILTRATION

5.1 General

The AWWA Water Quality Div. Com. on Coagulation-Filtration defines direct filtration as the treatment system in which filtration is not preceded by sedimentation (Culp, 1977). This definition includes flow sheets that utilize either flocculation basins or contact basins not equipped with sludge collection equipment and those which contain neither.

Direct filtration is not a new idea. Back in the early 1900's during the conversion period from slow sand to rapid sand filters there were several attempts at chemical treatment followed by rapid sand filtration without use of settling basins. These efforts failed because, in the fine-to-coarse single media filter beds that were used, most of the floc was removed in the top few centimeters of the bed and maximum head loss was reached rapidly. The development of coarse-to-fine dual-media and multi-media filters over the past fifteen years has made possible the storing of much larger quantities of floc in the beds without excessive head loss, thus making feasible the processing of a wide variety of raw waters by direct filtration. (Culp, 1977)

The direct filtration process differs from conventional sedimentation-filtration systems in that all solids - both those naturally occuring in the raw water and those added as part of the treatment process (alum, coagulant aid, filter aid and carbon for taste and odor control) - must be stored in the filter. It is sometimes known as contact-flocculation filtration. This is because flocs are formed inside the media. The rate of agglomeration of the destabilized floc occurs at a greatly accelerated speed because of the tremendous number of opportunities for contact afforded in the passage of the water through the granular bed. Shea et al (1971) state that there is ample experimental evidence to show the formation of large flocs within the filter pores. Whether these are formed in the liquid phase prior to their deposition or grow on the media after attachment is not clear. However, the authors guess that some aggregation does take place prior to deposition.

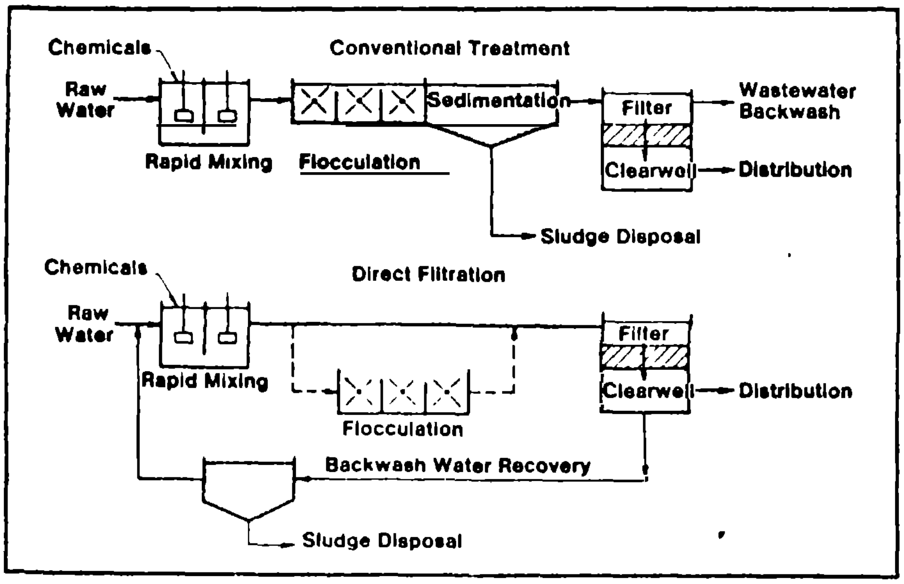


Figure 8. Comparison of conventional sedimentation filtration and direct filtration (Randtke, 1982)

5.2 Appropriate raw water quality

Dilute suspensions cannot be flocculated to a settleable size by the conventional flocculation process because the flocculator cannot provide the high velocity gradient necessary to induce adequate rate of particle contact (Adin and Rebhun, 1974). However, Habibian and O'Melia (1975) state that even such dilute suspensions can easily be filtered. The authors reason that this is due to the transport efficiency in the filter which is independent of particle concentration. After the onset of filtration the removal efficiency improves with increasing influent concentration during the ripening period because retained particles act as collectors for other suspended particles (O'Melia and Ali, 1978).

However, in terms of operational considerations, too high a suspended material load during direct filtration will result in impractically short filter runs before breakthrough. For this reason there is an upper limit to the water constituents beyond which the use of direct filtration is precluded. The following presents an initial guide to the appropriate water quality for the consideration of direct filtration.

According to Culp (1977) the possibilities of applying direct filtration to municipal plants are good if 1) the raw water turbidity and colour are each less than 25 units; 2) the colour is low and the maximum turbidity does not exceed 200 turbidity units; or 3) the turbidity is low and the maximum colour does not exceed 100 units. The presence of paper fiber or of diatoms in excess of 1000 areal standard units per millilitre (asu/ml) require that settling should be included in the treatment process chain. Diatom levels in excess of 200 asu/ml may require the use of special coarse coal on the top of the bed in order to extend filter runs. Coliform MPNS of 90 per 100 ml have been handled satisfactorily in direct filtration plants included in a recent AWWA survey and there appears to be no reason that substancially higher coliforms could not be removed in direct filtration.

Some workers referred to Bratby (1980), however, limit raw water turbidity to 5 to 10 turbidity units, stating that for average turbidities exceeding this value direct filtration may become inefficient due to short filter runs and breakthrough.

Voss and Gross (1981) refer to the experience of direct filtration in Guayana and report that direct filtration is qualified to cope with suspended solids concentration of up to 200 g/m^3.

This variety of values and parameters used to define the appropriate raw water quality understandably emanates from the lack of accepted standard measure for filterability of suspensions.

With this background in mind, it is needless to stress that the suitability of raw water for direct filtration cannot be determined from numerical values alone. In the words of Culp (1977) such values only provide preliminary indication. Pilot plant tests must be performed in each case to find out whether or not direct filtration will provide satisfactory treatment under the prevailing local circumstances of raw water quality.

Under appropriate raw water conditions and with proper engineering design direct filtration can produce water quality equal to that from plants that include settling and with equal reliability (Culp, 1977).

5.3 Filter media for direct filtration

Filter media considerations for direct filtration are basically the same as for filtration preceded by settling. The filter media is usually supported on a gravel bed. This is preferred to direct support on bottoms equipped with mechanical strainers or nozzles, which are not recommended (Culp, 1977).

5.31 Downflow filtration through heterogeneous media

When the effective grain size of a heterogeneous layer of sand is not uniform (when the uniformity coefficient high) throughout the whole depth, the fine sand comes up to the surface after backwashing. The filter media is thereby rendered unconducive to the use of the full bed. The retained impurities are arrested within the first few centimeters where they set up very large local head losses which are likely to shorten the filtration cycle and cause surface cracking of the media by lowering the pressure to below atmospheric level. Heterogeneous layer filters such as used in the conventional sand filtration have not been found adequate for direct filtration (Shea et al, 1971).

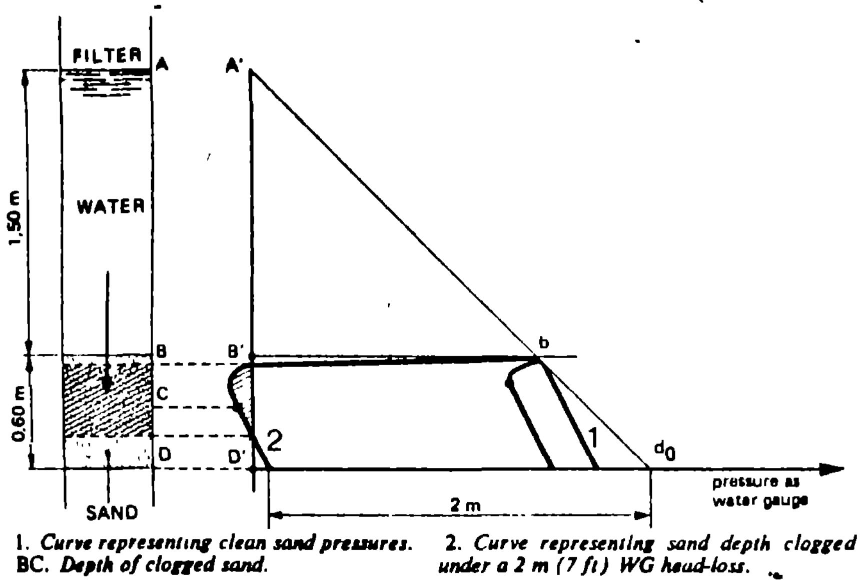


Figure 9. Downward filtration through a single heterogeneous layer-distribution of pressures. The cross-hatched portion of sand is under vaccum (Degremont, 1979)

5.32 Filtration through multi-media

Dual media and multi media techniques were developed in an effort to overcome the shortcomings inherent in the conventional non-uniform filter media. For a filter to give the highest possible throughput rate, an even deposition of material must be achieved throughout the full depth of the media, so that local development of head loss is avoided. As suggested by a number of researchers this can be achieved by making each succeeding layer more efficient so that each removes a greater proportion of the suspended matter into it. The best way to achieve this is to grade the filter media from coarse to fine in the flow direction (Shea et al, 1971; Hedberg, 1976; Degrêmont, 1979). With a heterogeneous media this can be accomplished only by upflow filtration. Another way to obtain coarse-to-fine filtration is by means of the multi-media filter, or by inducing a rate of flow through the media which decreases in the direction of flow. In both the latter alternatives suspended solids are forced further into the filter producing a more even distribution of head loss.

The multi-media filter bed is constructed of upper layers of coarse particles of low density and lower layers of fine particles of high density. The media are graded hydraulically during backwash with the coarser, less dense media being transported to the top of each layer; the more uniform the particle size distribution is in each layer, the less will be the fine-to-coarse gradation in each layer. As reported by Culp the multi-media is considerably more efficient than the traditional fine-to-coarse filter because its entire volume can be utilized for collecting suspended material without excessive head loss in any particular layer (Shea et al, 1971).

Nonetheless, dual and multi-media beds only partially meet the requirements of an ideal media because they are subject to mixing at the interfaces during backwashing. Furthermore anthracite which is normally used for the top layer is an expensive material whose homogeneity cannot be ensured (Ray, 1974).

5.33 Upflow filtration through heterogeneous media

Upward flow filtration is a logical development with respect to the requirement for filter media graded from coarse to fine in the direction of flow, since theoretically an ideally graded bed, shown diagramatically in figure 13, results from the natural disposition of uniform density grains after upflow backwashing. The relatively even distribution of silt arrestment throughout the filter results in even head loss development and consequently allows longer runs and/or higher flow rates than in comparable down flow filter. However, the higher flow rate which is feasible from other points of view is unfortunately usually sufficient to expand the bed to the extent that there is an increased tendency for premature breakthrough of retained material. The methods of combating expansion of the bed during filtration are beyond the scope of this paper.

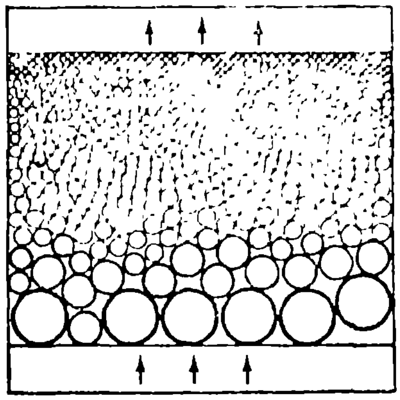
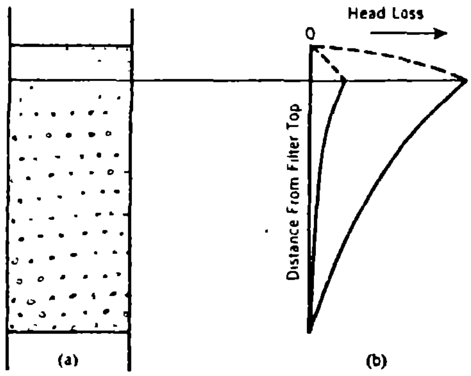


Figure 10. The ideal filter media

Design and operational problems make it unlikely that upflow filtration will find wide application as an alternative to rapid gravity or pressure filtration (Ray, 1974; Degremont, 1979). Shea et al (1971) also state that upflow filtration is a pattern which poses some technical problems.

5.34 Biflow filtration through heterogeneous media

Biflow filtration was developed in an effort to avoid the shortcomings of the upflow filter, i.e. the expansion or lifting of the bed during filtration. This is achieved by locating the filtrate pipe in the upper sand layer and filtering simultaneously from the top through a short depth of fine sand and from the bottom upwards through the bulk of the bed (figure 11). Such a filter is said to be self-regulating in proportioning the flow between the top layer and the main depth. It also maintains head losses towards the burried filtrate pipe, thus preventing any bed expansion or lifting. Because the whole depth of the filter is used in such arrangements, more efficient performance is the result. (Ives, 1964)



Principle of Biflow Filtration

About 20 per cent of the filtrate enters from the top of the column The other 80 per cent enters from the bottom. The sand at the top is fine, that at the bottom, coarse. The line running from 11(a) to 11(b) represents the point at which the filtrates coming from opposite directions meet. In 11 (b), the left-hand curves show initial head loss The right-hand lines show the loss after some time of operation.

Figure 11. Principle of biflow filtration (Ives, 1964)

5.35 Filtration through monograded media

The development of techniques such as multi-media, upflow and biflow filtration attempts to overcome the problem of fine to coarse stratification in a heterogeneous sand bed, but at the expense of a certain increase in complexity. An alternative approach is to retain downflow, single medium filtration but to use sand of relatively uniform coarse grading. For general water treatment application the sand has an effective grain size of 0,95 mm, a uniformity coefficient of less than 1,5 and is normally 0,85 to 1,0 m deep (Ray, 1974). In such media the effective grain size of the filtering material is more or less the same throughout the whole depth of the bed both initially and after backwashing. During filtration, therefore, the impurities penetrate deep into the sand instead of clogging the surface. In addition, the use of the coarser sand reduces the risk of formation of a vaccum. The media is washed simultaneously with air and water and is rinsed without expansion of the bed (Degrémont, 1979).

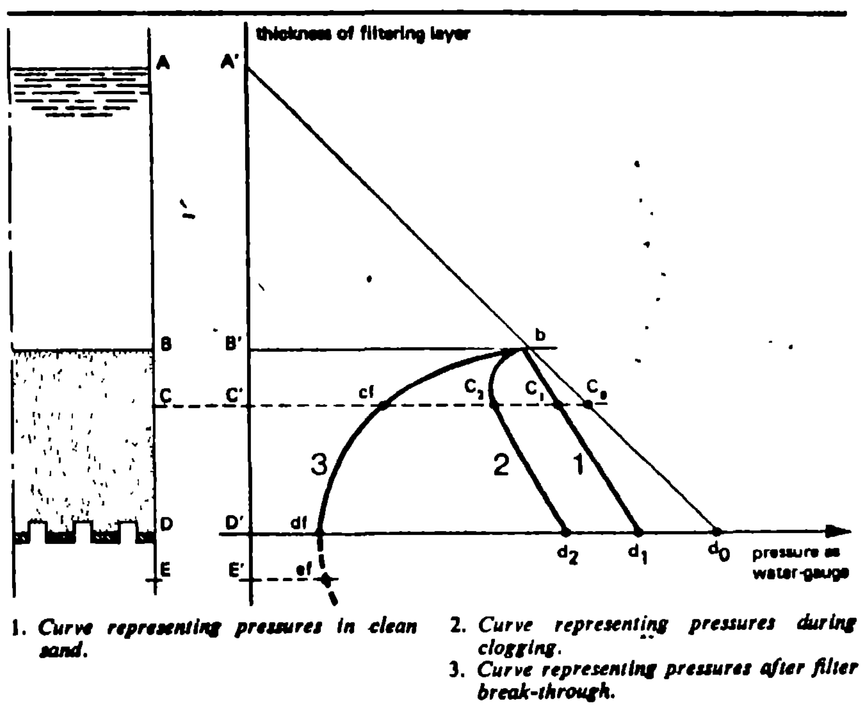


Figure 12. Distribution of pressure in monograded filter layer (Degrémont, 1979)

Ray (1974) reported that monograded sand filters are in use in France, South Africa, Britain and elsewhere. The Committee Report (1980) also indicates that more recent studies indicated good filtration results with these type of media in direct filtration.

Concerning the filtration rate Ray (1974) states that the media could be designed to operate between 7,5 and 30 m/h both in downflow or closed pressure form whereas Craft reported (Letterman, 1979) that single medium sand beds are inadequate for direct filtration when the filtration rates exceed 12 m/h.

5.36 Recommended media type

The use of downflow dual-media (e.g. anthracite and sand) or multi-media (e.g. anthracite, sand and garnet) has been recommended as more appropriate by Culp (1977).

In Guyana (Voss and Gross, 1981) the following layers of materials were used successfully in direct filtration

	Grain size (mm dia)	Density (g/cm^3)
Pumice stone	2 - 3	1,1
Hydroanthracite	1,2 - 2	1,5
Quartz sand	0,6 - 1	2,4

Other types of media were successfully applied to direct filtration in various developed countries (Ray, 1974).

O'Melia and Ali (1978) studied the role of retained particles in deep bed filtration and concluded it is plausible that the advantages of multi-media beds have been overstated. They argue that in most filtration processes the particles in the filter influent provide most of the removal after the run has begun. Retained particles accumulate in the upper regions of the bed and lead to the retention of more particles in that area. Hence removal and head loss tend to be localized in the upper regions of the bed during downflow filtration regardless of the size of media. According to the authors this effect is pronounced especially with concentrated suspensions.

Degrémont (1979) also point out some of the drawbacks connected with the use of multi-media. They write the following in favour of monograded media:

> Compared with the filter with a single uniform layer of sand, the advantages of the multi-layer equipment begin to dwindle. The fact of the matter is that a filter with a single uniform layer can operate at the same filtering rates, with the same cycle duration

> and with the same final head loss, and this by using a uniform and slightly finer grade of sand than anthracite of the double layer equipment combined with a greater depth to allow the retention in depth of the same quantity impurities per m^2 of filter area. Where washing is concerned, the advantage is on the side of the single-layer sand filter, washed with water and air, in which the existence of mud-balls is unknown and where the loss of sand is very slight.

Despite the numerous recommendations for the use of multi-media, the contentions of O'Melia and Ali (1978) and Degrémont (1979) should be given due consideration before choosing the type of media.

In selecting the media size, the finest media concomittant with appreciable filter runs should be selected from pilot plant trials. This serves to minimize polymer dosages. With coarse filter media the higher shear intensities require higher polymer dosages to increase the shear strength of floc (Bratby, 1980).

The ideal filter media is one which would result in the uniform floc and head loss over the depth of the bed under a wide variety of conditions, and still produce an acceptable effluent.

5.4 Backwashing of the media in direct filtration

Filtration consists of two closely interrelated sequential cycles, i.e. filtering and backwashing. During backwashing, a portion of the high quality water that was produced during the preceding filtering cycle is required for back-washing the filter media. The effectiveness of a backwashing operation has significant inpact on filter performance of the subsequent filter run.

The mechanism of sand cleaning has traditionally been based on the assumption that the deposited solids are disloged from the grains primarily by the abrasion of grains of filter medium rubbing against each other and colliding with each other. It was not until the early 1970's that the concept of hydrodynamic shear as the predominant cleaning mechanism evolved. Several investigators concluded from the literature that the effect of collisional interactions between particles in a fluidized state was relatively insignificant and that the principal mechanism is hydrodynamic shear (Huang, 1979).

The important conclusion for practice is that backwashing with water alone is an inherently weak cleaning process due to the limitations in particle collisions. Air scour and surface wash that promote interparticle abrasions during backwash are indespensible for effective cleaning (Amirtharajah, 1978).

Surface wash or air scour is a necessary part of filters used in direct filtration. With the increasing use of polyelectrolytes and in particular their use as sole coagulants, backwashing techniques become important (Committee Report, 1980).

A number of methods can be employed to achieve backwashing (Degrémont, 1979):

5.41 Washing with water alone with expansion of filter bed

The current of water must be sufficient to expand the filtering material, i.e. to bring about an apparent increase in its volume of at least 15 %. As the viscosity of water varies with temperature, it is desirable that a system should be provided for measuring this and for regulating the flow of wash-water so as to keep the degree of expansion constant over time. The expanded layer then becomes subject to convection currents. In certain zones the filtering material moves downwards and in others upwards, which means that portions of the compact layer of sludge encrusting the filtering material surface are carried deep down to form hard and bulky mud balls as a result of the whirling action of the currents. Thus this method calls for considerable care and is unsatisfactory for downward filtration.

5.42 Simultaneous air and water washing without expansion

The second method, now becoming widespread, is to use a backwash velocity which will not cause expansion of the bed, and at the same time to disturb the sand by air scour. The surface crust is completely broken up by the air. Mud balls are unknown with this method of backwashing process.

During air scour the wash-water flow-rate can be varied over wide range, but it must not fall below 5 m^3/h per m^2. The higher this flow-rate, the more rapid and effective will be the washing. The maximum figure will depend on the material and filter parameters.

Rinsing may be carried out by the following methods after the air scour has stopped

- continue the backwash at a constant rate of flow until the discharged water runs clear. The rate of flow must not drop below 12 m^3/h per m^2;

- increase the rate of flow of water during rinsing to at least 15 m^3/h per m^2.

5.43 Washing with air and water in succession

This method of washing is used when the nature of the filtering material is such that it is impossible to use air and water simultaneously without running the risk that the wash-water will carry off the filter media to the drain. This applies to multi-media filter beds which have low-density materials such as anthracite on top.

In the first stage of the washing operation air is used by itself to detach the retained impurities from the filtering material. In the second stage a backwash of water with a sufficiently high flow-rate to bring about the expansion of the bed enables the impurities detached during the first stage to be removed from the bed.

After a filter has been backwashed and restored to service, 10 - 20 minutes may be required before the desired effluent is produced. Prior to that time, the water should be filtered to waste. One way to cut the length of this filter period is to treat the filter backwash water with polymer (Culp, 1977).

6. COAGULATION WITH METAL SALTS

Coagulation is an essential part of the solids - liquids separation process. It dates from the early days of recorded history when various natural materials, such as crushed almonds and beans in Egypt, nuts in India and alum in China were used to clarify turbid water. Early studies showed the advantages of the addition of a coagulant (generally alum or iron sulfate) both to coloured and turbid waters. The first scientifically performed study was conducted by Austen and Wilbur in 1885, who suggested the use of alum prior to filtration (Committee Report, 1971).

When the abundant literature on the subject is consulted it becomes evident that the terms coagulation and flocculation are being used interchangeably. Bratby (1980) defines both terms as follows:

> Coagulation is the process whereby destabilization of a given suspension or solution is effected. That is, the function of coagulation is to overcome those factors which promote the stability of a given system by double layer compression and charge neutralization.
>
> Flocculation is the process whereby destabilized particles, or particles formed as a result of destabilization, are induced to come together, make contact and thereby form large(r) aggregates. With polyelectrolytes flocculation takes place through bridging as well.

In direct filtration coagulation is the single most unavoidable pretreatment. It could only be avoided in the removal of iron and manganese.

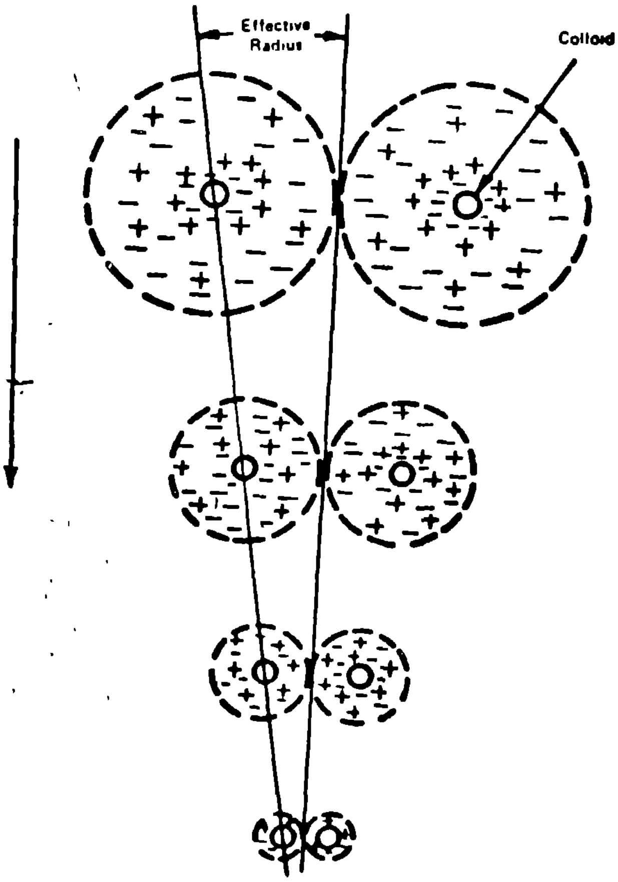


Figure 13. Coagulation (double layer compression and charge neutralization)
Arrow indicates increased addition of coagulant (Beardsley, 1973)

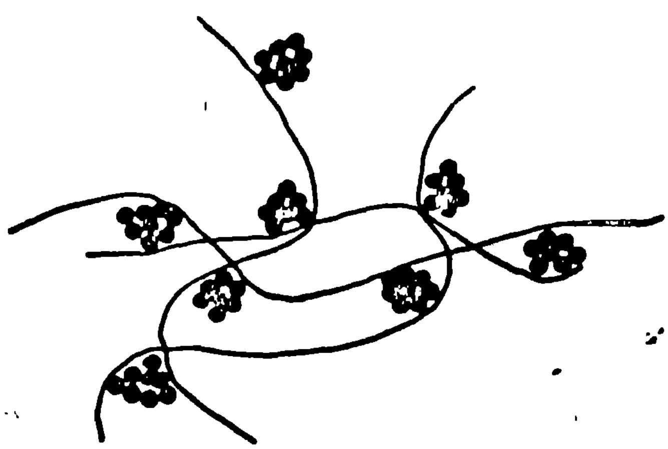


Figure 14. Flocculation (agglomeration and bridging) (Beardsley, 1973)

The commonly used metal coagulants fall into two general categories: those based on aluminium and those based on iron. The aluminium coagulants include aluminium sulphate, aluminium chloride, polyaluminium chloride and sodium aluminate. The iron coagulants include ferric sulphate, ferrous sulphate, chlorinated copperas and ferric chloride. Other chemicals used as coagulants include hydrated lime and magnesium carbonate. The popularity of aluminium and iron coagulants arises not only from their effectiveness as coagulants but also from their ready availability and relatively low cost. The efficay of these coagulants arises principally from their ability to form multi-charged poly-nuclear complexes in solution with enhanced adsorption characteristics. The nature of the complexes formed may be controlled by the pH of the system. (Bratby, 1980)

6.1 Theoretical considerations

In the field of colloid science, at least two different approaches have been advanced historically to explain the basic mechanisms involved in the stability and instability of colloid systems. The first theory is the so called chemical theory, which assumes that colloids are aggregates of definite chemical structural units and emphasizes specific chemical interactions between the coagulant and the colloids. According to this theory, the coagulation of colloids is the result of a principitation of insoluble complexes that are formed by specific chemical interactions.

The second theory - the physical or double-layer theory emphasizes the importance of the electrical double layers surrounding the colloidal particles in the solutions and the effects of counter-ion adsorption and zeta-potential reduction in the destabilization of colloidal systems. These two theories may appear to be contradictory, but they are not mutually exclusive. As a matter of fact, both mechanisms must be employed in a comprehensive understanding and in effective control of colloid stability and instability (Committee Report, 1971). The term stability refers to the capacity of colloidal or smaller dimensions to remain as independent entities within a dispersion. Bratby (1980) states that particles smaller than of the order of 10^{-3} mm are referred to as colloids whereas material smaller than approximately 10^{-6} mm are referred to as comprising solutions.

6.11 Physical mechanisms

The destabilization of a colloidal system by the reduction of the repulsive zeta potential in the following two ways has been discussed in the section on mechanisms of filtration.

1) The compression of the double-layer thickness due to the incorporation of simple nonhydrolysed ions (such as sodium and calcium ions) into the diffuse double layer.

2) The specific adsorption of the counter ions onto the particle surface, with a concurrent reduction in the surface potential of the colloidal particles.

Thus, the possibility of colloidal particles having the same sign of charge to approach each other, i.e. the possibility of their coagulation will depend on the difference in their resultant interaction energy and kinetic energy. The interaction energy can be enhanced by reducing the resultant interaction energy, which is the net value of the coulombic electrostatic repulsive energy and the Van der Waal's attractive energy (figure 15). Reduction of the net interaction energy can be effected by the introduction of a coagulant capable of providing the necessary counterion into the stable system. The kinetic energy of the colloid can be supplied by either Brownian movement or turbulent mixing, depending on on the colloid size. Turbulent mixing, which creates enough driving force for destabilizing a colloidal system of larger particle size, is necessary for rapid coagulation results.

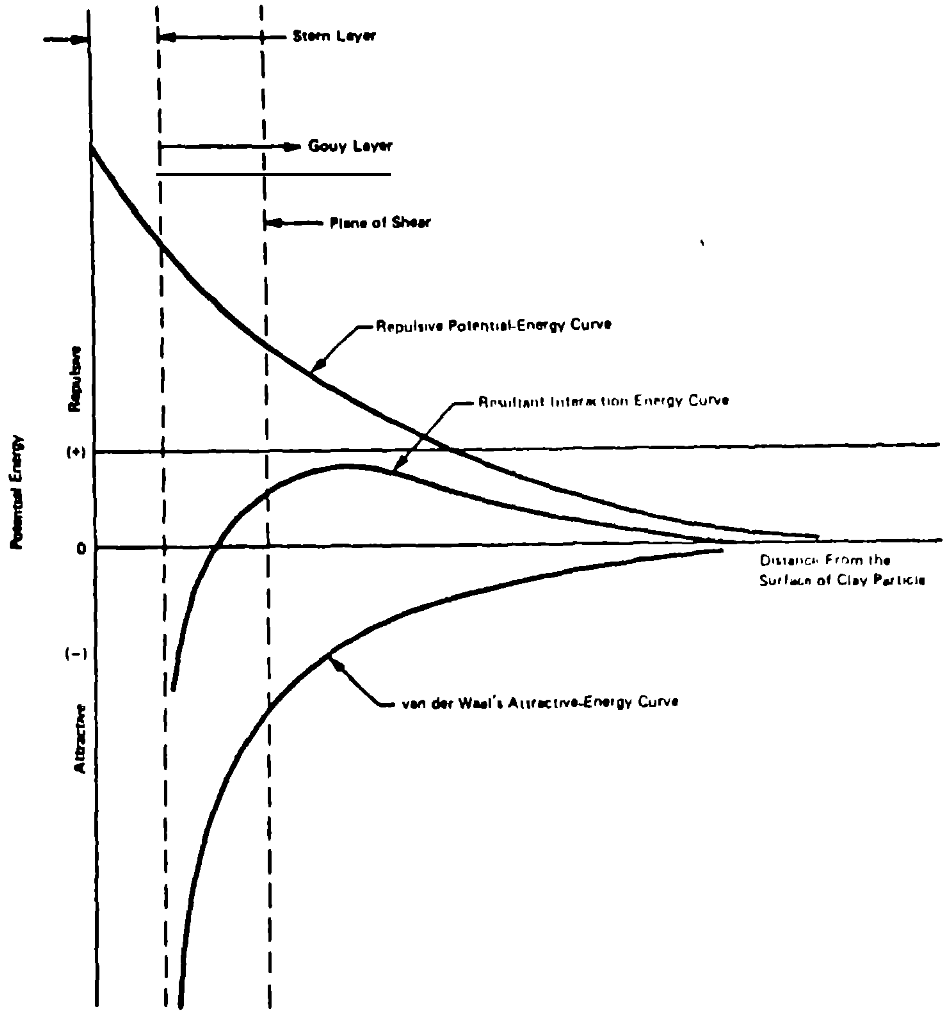


Figure 15. Potential energy of interaction of colloidal particles (Committee Report, 1971)

6.12 Chemical mechanisms

Destabilization reactions of colloids in aqueous dispersion are complex and arise not only from the electrical double layer compression and charge neutralization but also from precipitation of insoluble complexes that are formed by specific chemical interactions as noted earlier. The first studies to show a general stoichiometric relationship between the coagulant dosage and a measurable property of the colloidal system - the colour reported by Black, Singley et al in 1963 (Committee Report, 1971), reaffirmed the importance of a chemical mechanism of colloidal destabilization.

Further substantiation of the chemical nature of the destabilization of organic colour by iron (3^+) was presented by Singley, Maulding and Harris (Committee Report, 1971) when they showed that the optimum conditions for colour removal rarely coincided with the conditions required for the reduction of the repulsive zeta potential to zero. A purely physical model would certainly predict optimum coagulation under these circumstances. The Committee Report (1971) points out further that the charge could effectively be neutralized by using an organic polyelectrolyte in conjunction with ferric sulfate without producing colour removal was demonstrated.

The predominant specific chemical interaction depends very largely on two factors (Bratby, 1980): 1) The nature of the colloidal dispersion: whether hydrophobic (water repellant) or hydrophilic (a strong affinity for water molecules in the surface layers of the colloid) particles are prevalent; the surface nature of the colloid; the intensity of surface charge carried by the colloid and so on. 2) The type of coagulant added to the colloidal dispersion: whether coagulant species are charged or uncharged; the intensity of charge in the former instance; the adsorptive capacity of the species; the capacity for bridge formation between adjacent colloids etc.

Under appropriate conditions of coagulant concentration and pH, metal coagulants in aqueous solution form metal-hydroxide precipitates. Such species serve to enmesh particulate material thus effecting destabilization essentially by a "sweep action". This chemical mechanism of destabilization is that of precipitate enmeshment.

Precipitation mechanisms are also of importance during destabilization of hydrophilic colloids. Here, because of the extent of hydration, electrostatic effects are relatively unimportant. Coordination reactions occur between metal ions and certain functional groups on the particle surface. Destabilization in such cases is visualized as being the result of metal ion - functional group - hydroxide precipitate formation.

When metal coagulants are dissolved in water, the metal ions react with water or hydrolyse. Different species of the hydrolysis products are formed. As the extent of hydrolysis increases, progressively higher polynuclear species form. On adsorption of such polymeric species to particles a coagulant bridge spanning between adjacent particles is formed thereby promoting destabilization. Because of its relative significance this will be discussed further.

6.121 Extent of hydrolysis and adsorption

The hydrolysis of polyvalent metal ions in aqueous solution has reportedly been studied in detail by many investigators (Stumm and O'Melia, 1968; Committee Report, 1971 and Bratby, 1980), with considerable attention devoted to those ions of interest in coagulation.

Aluminium and ferric salts, when in solution, immediately dissociate to form hydrated reaction products. The metal ions form coordination compounds (Bratby, 1980) with water molecules to give $\left\{Al(H_2O)_6\right\}^{3+}$ and $\left\{Fe(H_2O)_6\right\}^{3+}$. These species, referred to as the trivalent ions of aluminium and iron, are often presented as Al^{3+} and Fe^{3+} for reasons of convinience in presentation.

Stumm and Morgan (1962) emphasized that the effects of ferric and aluminium salts upon coagulation are not brought about by the simple aquo-metal ions themselves ($Fe(H_2O)_6^{3+}$ and $Al(H_2O)_6^{3+}$), but by their hydrolysis products. These hydrolysis products are multinuclear hydroxo-metal complexes that may be highly charged. Complex formation of these ions can occur not only with OH^-, but also with other bases and with ionizable groups on colloids.

The term hydrolysis refers to the general reaction in which a proton is transferred from an acid to water, or from water to a base. The hydrolysis of metal ion is a stepwise replacement of coordinated molecules of "water of hydration" by hydroxyl ions. The replacement occurs by the transfer of protons from waters of hydration to free water molecules to form a hydronium ion.

The hydrolysis of iron (3+) and aluminium ions to yield a variety of hydrolysis products may be represented as follows:

$$Al(H_2O)_6^{3+} + H_2O \rightleftharpoons Al(H_2O)_5OH^{2+} + H_3O^+$$

$$Al(H_2O)_5OH^{2+} + H_2O \rightleftharpoons Al(H_2O)_4^{+} + H_3O^+$$

$$Al(H_2O)_4OH_2^{+} + H_2O \rightleftharpoons Al(H_2O)_3(OH)_3^{(s)} + H_3O^+$$

$$Al(H_2O)_3OH_{3(s)} + H_2O \rightleftharpoons Al(H_2O)_2(OH)_4^{-} + H_3O^+$$

(Committee Report, 1971)

The extent of this substitution depends on the concentration of the substituted OH^- ions. In other words,

> the extent to which OH^- ions are bound to the metal complex is dependent primarily on the pH of the solution.

A wide variety of such soluble species has been reported by various investigators, including

$Fe_2(OH)_2^{4+}$	$Fe_2(OH)_4^{4+}$
$Al_2(OH)_2^{4+}$	$Al_4(OH)_8^{4+}$
$Al_2(OH)_5^{+}$	$Al_6(OH)_{15}^{3+}$
$Al_7(OH)_{17}^{4+}$	$Al_8(OH)_{20}^{4+}$
$Al_{13}(OH)_{34}^{4+}$	

(Stumm and O'Melia, 1968)

According to Stumm and O'Melia (1968) multimeric hydroxo-metal complexes of the form Meq $(OH)p^{z+}$ are almost of universal occurence in the water solvent system.

Under equilibrium conditions both iron (3^+) and aluminium exist primarily as the insoluble solids $Fe(OH)_3$ or $Al(OH)_3$ as shown in figure 16.

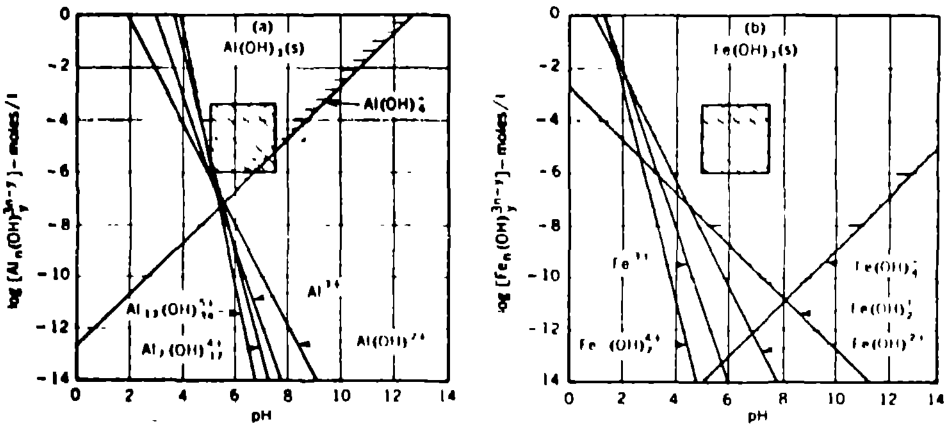


Figure 16. Equilibrium composition of solutions in contact with freshly precipitated $Al(OH)_3$ and $Fe(OH)_3$ (Stumm and O'Melia, 1968)

Singley and co-workers (1968) have shown that, under the nonequilibrium conditions existing in water treatment plant coagulation, the predominant species may not be the simple insoluble species, except for iron (3^+) in solutions of 1×10^{-4} Moles or less, where the predominant species are negatively charged at pH values above about 4,5. Distribution diagrams for iron (3^+) and aluminium are shown in figure 17 from the studies of Singley and co-workers (1968). There is a general agreement that the settleable or filterable precipitates formed are hydrolysed iron or aluminium complexes that have adsorbed or chemically combined with impurities to be removed (Committee Report, 1971).

Under favourable solution conditions (pH, temperature, applied metal ion concentration, time of aging), the hydrolysis products of iron (3^+) and aluminium have a different charge than the metal ions themselves, and are adsorbed more readily at particle water interfaces than nonhydrolysed metal ions. This tendency to be adsorbed is especially pronounced for polynuclear polyhydroxo species. Stumm and O'Melia (1968) point out that no adequate theory for this enhanced adsorption by hydrolysis is available. The authors, however, give two likely qualitative reasons. First, hydrolyzed species are larger and less hydrated than non-hydrolysed species. Second, the enhancement of adsorption is apparently due to the presence of a coordinated hydroxide group.

Thus it is evident from the foregoing discussion that by adjusting the pH the types of the hydrolysed species can be controlled for the best adsorption destabilization for direct filtration purposes.

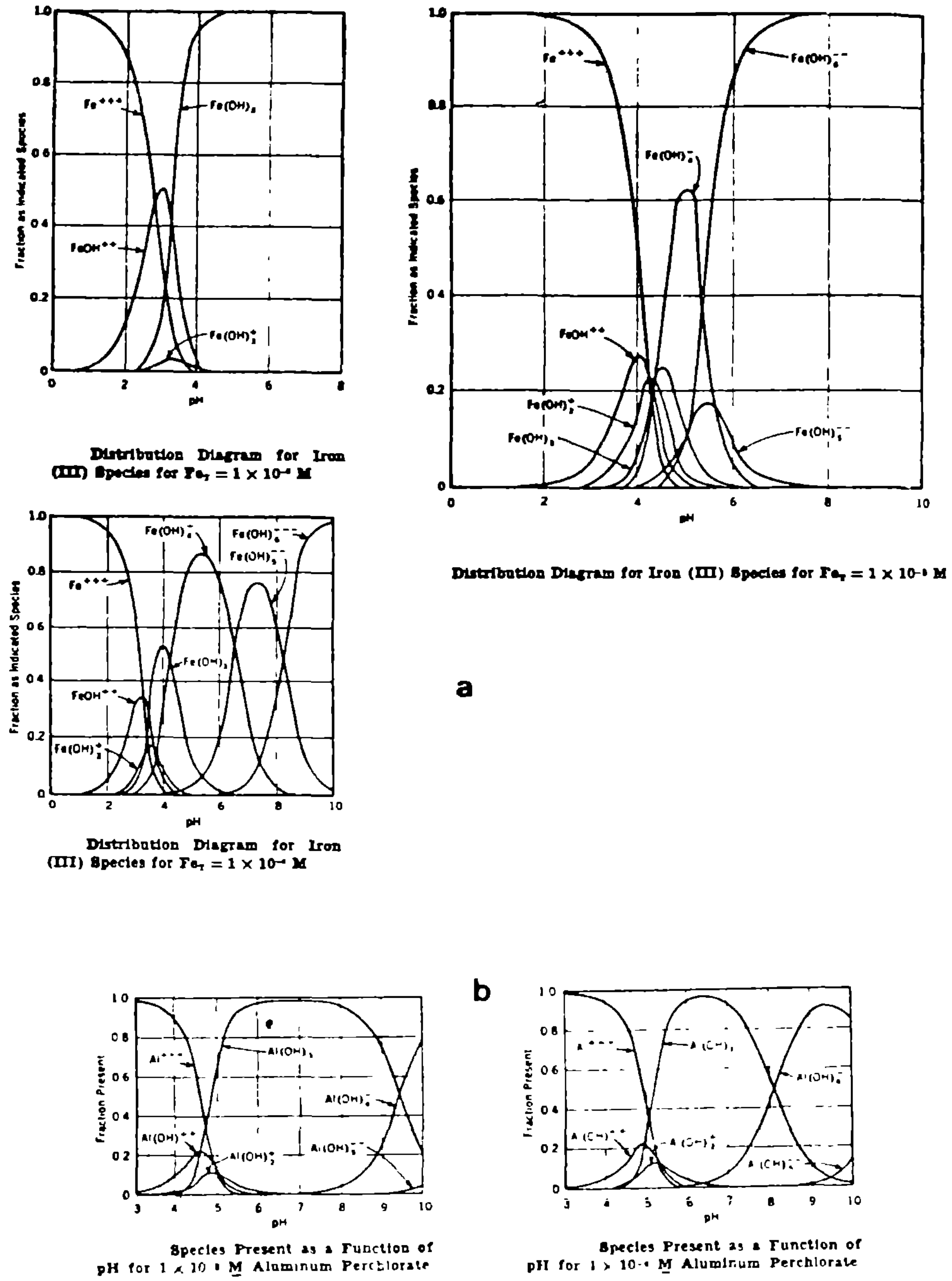


Figure 17. a) Species distribution diagrams for iron (3^{+}) during destabilization. b) As for a) but for aluminium (3^{+}) (Bratby, 1980)

6.2 Practical considerations

O'Melia (1982) states that turbid waters can be classified into the following four types:

1. High turbidity, high alkalinity:
   Alum and ferric salts generally prove effective. The use of these metal salts with such waters does not frequently necessitate the use of coagulant aids or addition of base for pH control.

2. High turbidity, low alkalinity:
   With these types of waters addition of base may be needed to prevent pH from falling below levels at which aluminium or ferric polymers are formed.

3. Low turbidity, high alkalinity:
   Alum and ferric salts are effective in relatively large doses, so that $Al(OH)_3^{(s)}$ or $Fe(OH)_3^{(s)}$ is precipitated. Addition of clay or activated silica beforehand may reduce the dosage requirements.

4. Low turbidity, low alkalinity:
   These are reportedly the most difficult waters to coagulate. Clays or other targets may be added. Alum and iron (3^+) salts used alone are usually ineffective since the pH can be lowered below the neutral range where $Al(OH)_3^{(s)}$ and $Fe(OH)_3^{(s)}$ are produced and sweep coagulation is achieved. The addition of lime or other base can bring it to water type 3.

6.21 Effect of coagulant dose on turbidity

The classical "residual turbidity" versus "alum dose" plot (Snoeyink and Jenkins, 1980), figure 18, illustrates how alum functions as a coagulant at constant pH. At low alum doses there is no reduction in turbidity. At these low doses there is insufficient hydroxoaluminium (3^+) species to provide effective destabilization. The final turbidity remains constant or even increased slightly with increasing coagulant dose; this increase being due to the formation of hydrolysis products of aluminium.

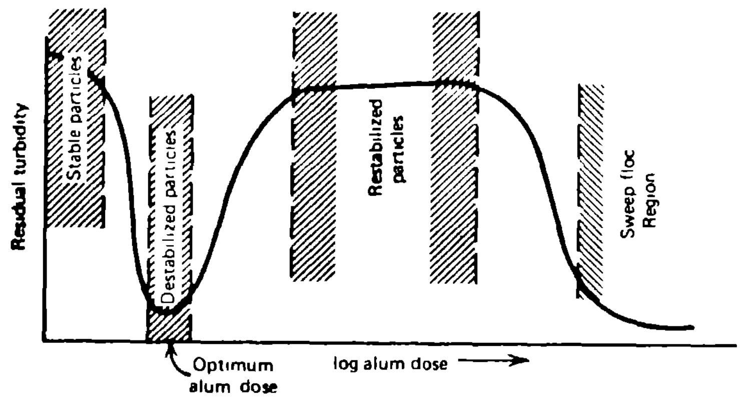


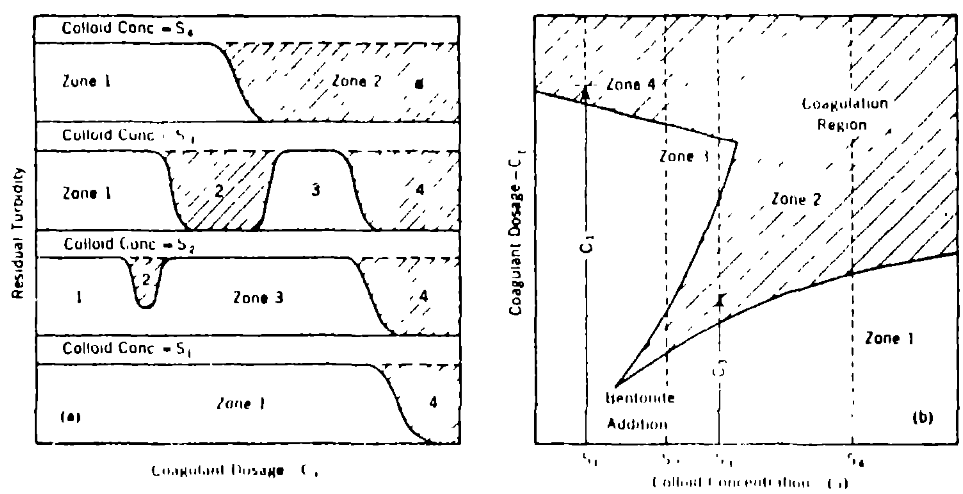
Figure 18. **Alum dose versus residual turbidity for water coagulation/flocculation. The residual turbidity is that which remains in a test solution to which alum was added After mixing to simulate that which occurs in a water treatment plant, the sample is allowed to settle for 30 min before turbidity is measured**

(Snoeyink and Jenkins, 1980)

The further addition of alum to the point at which complete destabilization occurs causes a reduction in turbidity to a minimum value. Further increase in alum dose will result in restabilization of the particles because of a near complete coverage of the particle with aluminium hydrolysis products. More addition of alum to very high doses results in the formation of a precipitate of $Al(OH)_3^{(s)}$ because the solubility product of $Al(OH)_3^{(s)}$ is exceeded (Snoeyink and Jenkins, 1980). This bulky precipitate enmeshes particles in it and settles rapidly to form the so called "sweep floc" region for the aggregation of colloidal suspensions. At most water treatment plants coagulation/flocculation takes place in the "sweep floc" range because it is very difficult to vary the coagulant dose to correspond to the varying influent conditions as required to operate in the range of complete particle destabilization (Snoeyink and Jenkins, 1980).

In direct filtration practice, however, the zone of destabilization is recognized to be the most useful. Research work carried out in connection with mix design for mechanisms of alum coagulation indicates that at dosage levels below 3,0 mg/l there is a zone in which adsorption destabilization occurs, but "sweep flocculation" does not take place (Wagner and Hudson, 1982). In the destabilization zone, excellent coagulation occurs but flocculation does not take place. With direct filtration the destabilized particles can be removed by being adsorbed to the filter media. As filter clogging is related directly to coagulant dose (Wagner and Hudson, 1982), the reduced dosage requirement increases the chances of successful treatment by direct filtration resulting in longer runs. This is probably why the afore-mentioned researchers suggest that the use of the adsorption-destabilization process should be well-suited to direct filtration.

It is worthwhile to note, however, that for any particular pH the dosage required for bringing about the zones of adsorption destabilization and "sweep floc" is dependent upon the colloid concentration in the water to be treated (Stumm and O'Melia, 1968).



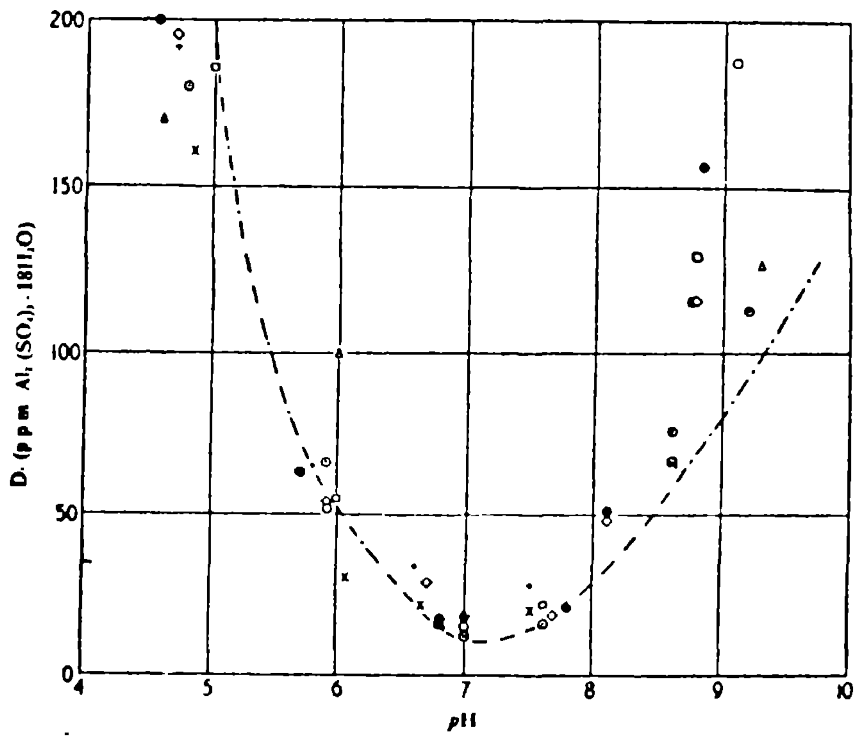
Destabilization (Zone 2), Restabilization (Zone 3), and Precipitation ("Sweep Floc," Zone 4) Regions for the Aggregation of Colloidal Suspensions by Al(III) or Fe(III)

(a) schematic coagulation curves at constant pH for four colloid concentrations, (b) effect of colloid concentration (expressed as concentration of surface, typical concentration units are square meters per liter) on the dosage of coagulant required to produce destabilization, restabilization, and precipitation regions. pH is constant. Curves in (a) are sections through (b) at the surface concentrations indicated by the vertical dashed lines in (b). Shaded areas in (a) and (b) denote regions in which coagulation occurs.

Figure 19. (Stumm and O'Melia, 1968)

6.22 Effect of type of turbidity on coagulant dosage

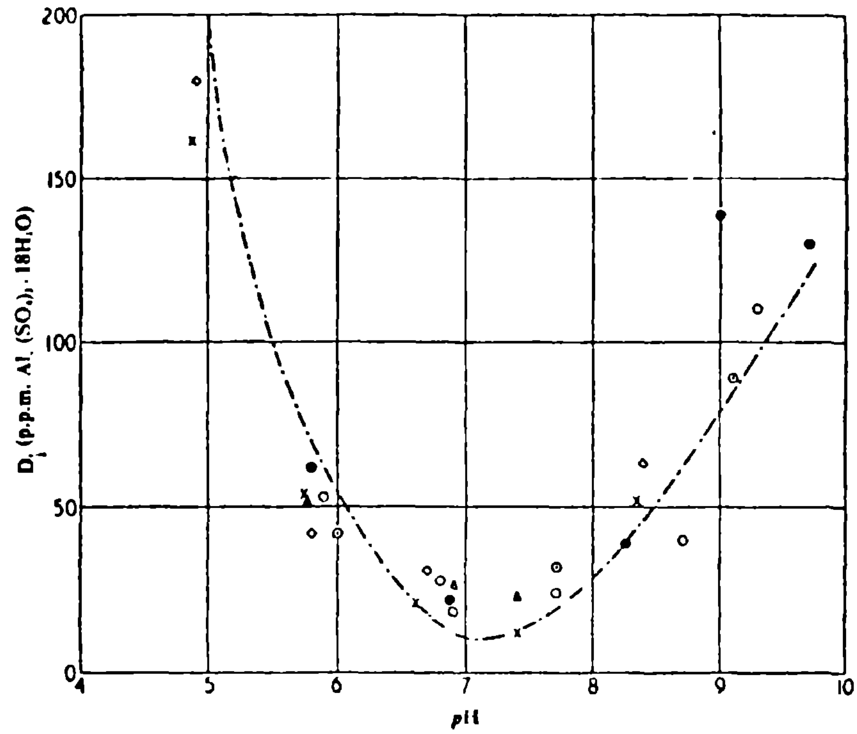
Works carried out by Packham (1963) show that the dose of aluminium sulphate required for coagulation is markedly affected by the pH but is largely independent of the nature of the material in suspension. The studies indicate that minerals with and without organic matter follow the same general pattern of behaviour as clay minerals with the exception of montmorillonite. Figures 19 and 20 illustrate this pattern of behaviour. The results for a pure clay mineral, kaolinite, are shown as a dotted curve for comparison.



Mole ✓, Severn ○, Kennet △, Great Ouse ○, Nar ⌐, Itchen ◇, Chelmer ⊕, Kaolinite (Speswhite)

The effect of pH on the coagulation of 50 ppm suspensions of minerals isolated from various rivers (organic matter present)

Figure 20. (Packham, 1963)



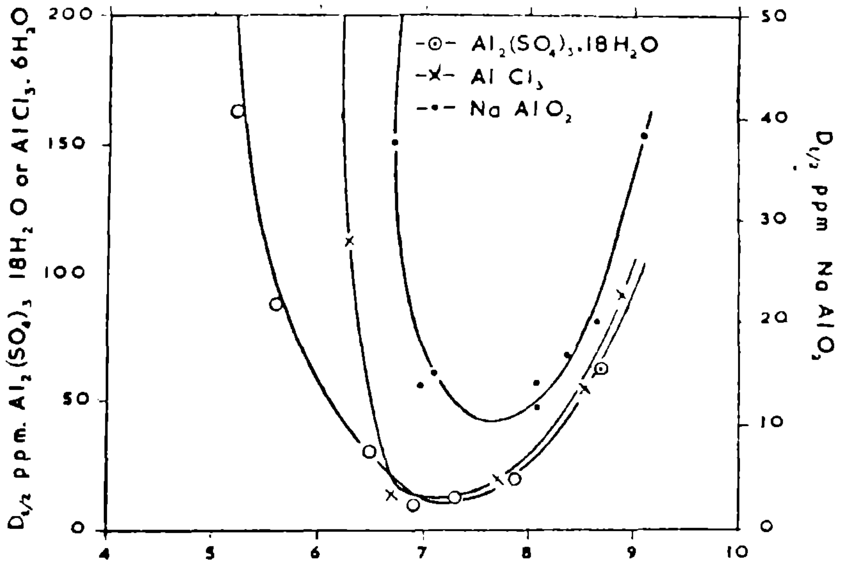
Mole ×, Severn ○, Kennet △, Great Ouse ○, Itchen ◇,
Kaolinite (Speswhite)

The effect of pH on the coagulation of 50 ppm suspensions of minerals isolated from various rivers (organic matter removed)

Figure 21. (Packham, 1963). The pH of optimum destabilization is between 6,8 and 7,8, and on each side of this range the coagulant dose required increases rapidly with change of pH.

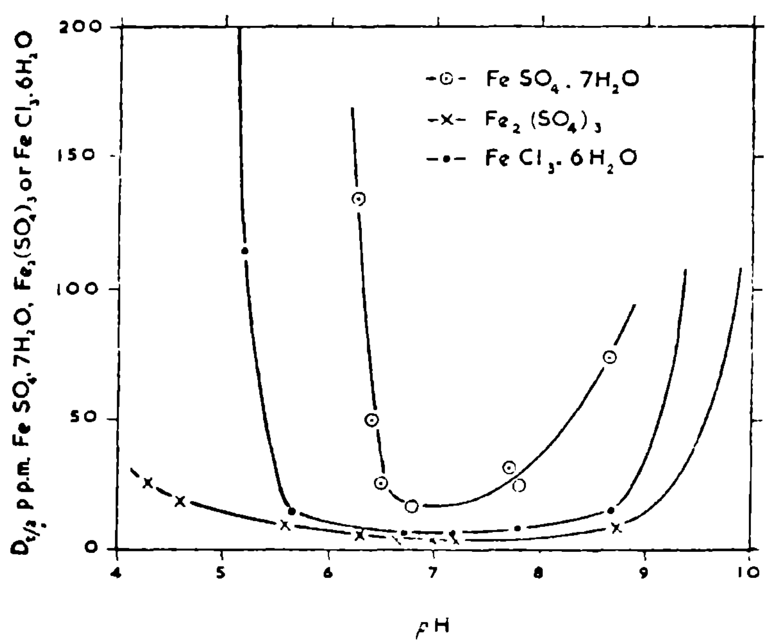
6.23 Effect of type of coagulant on pH of optimum destabilization

The optimum pH range for coagulation of turbidity varies with the type of coagulant used. The results of coagulation tests on 50 mg/l kaoline suspensions using aluminium sulphate, aluminium chloride, sodium aluminate, ferrous sulphate, ferric sulphate and ferric chloride are shown in figure 21 and figure 22.



Coagulation of 50 ppm kaolin with aluminium compounds

Figure 22. (Packham, 1963)



Coagulation of 50 ppm kaolin with iron compounds

Figure 23. (Packham, 1963).

For aluminium salts and sodium aluminate the pH range is quite narrow, whereas the range for ferric salts it is from about 5,5 to 8,8.

6.24 Metal coagulants and rapid mixing

The hydrolysis and adsorption of metal coagulants are extremely rapid and essentially irreversible. Furthermore, the rates of formation and the types of species that develop are undoubtedly dependent, among other things, upon local concentrations of metal and hydroxide ion (Jorden and Vrale, 1971).

That the hydrolysis-adsorption reactions are extremely rapid and that the rates of formation of species depend on local concentrations suggest that rapid mixing is necessary to ensure non-equilibrium conditions. Non-equilibrium conditions, as pointed out earlier, discourage the formation of the insoluble $Al(OH)_3$ and promote the polymerization process. It also ensures the homogenization of the destabilizing chemical and the water.

For direct filtration the rapid mixing process does not usually differ from that of the conventional system. However, a hydraulic jump or parshall flume structures have been reportedly used in direct filtration with good results (Culp, 1977). These have features that make them suitable for rapid mixing the metal coagulant with the raw water much more efficiently than the conventional backmix reactors (Jorden and Vrale, 1971).

7. COAGULATION WITH POLYELECTROLYTES

A polymer molecule is defined as a series of repeating chemical units held together by covalent bonds. Polyelectrolytes are special classes of polymers, distinguished from ordinary polymer molecules by the possession of ionizable functional groups along the polymer chain. When these groups dissociate the polymer molecules become charged either positively or negatively, depending on the specific functional groups present, and are referred to as cationic and anionic polyelectrolytes, respectively. Polyelectrolytes that possess both positively and negatively charged sites are called polyampholytes, whereas those that possess no ionizable functional groups are termed nonionic polyelectrolytes. (Committee Report, 1971)

Polyelectrolytes are effective in enhancing the rate of orthokinetic flocculation when added to a system already distabilized with, say, metal coagulants. Polyelectrolytes may also be effectively applied as primary coagulants to satble colloid system. Furthermore, there are instances where polyelectrolytes are effective in precipitating substances dissolved in solution. From the above comments, it is appreciated that the destabilization mechanism operative with polyelectrolytes is complex. For a given system, there may be a dominance of charge effects, or adsorption or chemical reactions at the functional groups (Bratby, 1980).

7.1 Mechanisms of destabilization

However, according to Bratby (1980) it is possible to set down two principal mechanisms based on 1) a bridging model, where polyelectrolyte segments are adsorbed on the surfaces of adjacent colloids thereby binding them together and 2) a model whereby ionic polyelectrolytes, bearing a charge of opposite sign to the suspended material, are adsorbed and thereby reduce the potential energy of repulsion between adjacent colloids. The two mechanisms introduced above are referred to as the bridging mechanism and the electrostatic patch mechanism.

7.11 The bridging mechanism

The principal phenomenon for acceptance of the bridging mechanism lies in the ability of charged polyelectrolytes to destabilize particles bearing the same charge. Furthermore, Bratby (1980) points out that direct evidence is available whereby electron micrographs have indentified polyelectrolyte bridge between particles.

Stages in the bridging mechanism of destabilization with polyelectrolytes include 1) dispersion, 2) adsorption, 3) compression or settling down and 4) collision.

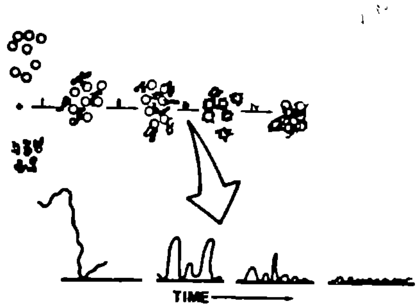


Figure 24 stages in the bridging mechanism of destabilization with polyelectrolytes: I dispersion, II adsorption, III compression or settling down and IV collision (Bratby, 1980).

7.12 The electrostatic patch mechanism

For the case of non-ionic and anionic polyelectrolytes applied to a negatively charged colloidal dispersion, a destabilization mechanism described by the bridging model adequately accounts for the phenomena taking place. However, for the case of charged polyelectrolytes applied to dispersions with particles carrying surface charges of opposite sign, the bridging model is often inadequate. Such systems include cationic polyelectrolytes applied to a negative colloidal dispersion. It could also include anionic polyelectrolytes applied to dispersions destabilized with metal coagulants i.e. as flocculant aids to particle-metal hydrolysis product aggregates, which may be positively charged. An electrostatic patch mechanism has been proposed for the afore-mentioned types of systems where a strong electrostatic attraction between polyelectrolyte and particle surface exists. Rather than adsorption of the polyelectrolyte at only a few sites, with the remainder of the chain extending into solution, virtually complete adsorption of it onto the particle surface takes place. The adsorbed polyelectrolyte chains thus form a charge mosaic with alternating regions of positive and negative charge (figure 25) (Bratby, 1980).

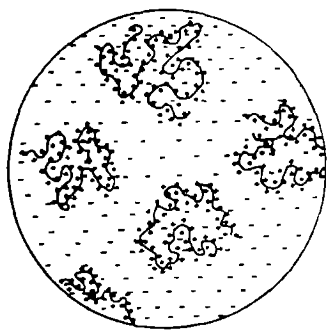


Figure 25. Possible arrangement of adsorbed polycations on a particle with low negative surface charge density (Bratby, 1980)

7.2 Polyelectrolytes as primary coagulants

There are many instances where polyelectrolytes have been employed as primary coagulants, effectively replacing the use of metal coagulants. Such applications include (Bratby, 1980) treatment of waters predominantly turbid or coloured with humic substances. Shea et al (1971) applied a catonic polyelectrolyte to direct filtration of turbid waters and found it to be superior in performance to alum in terms of both technical (filter runs, effluent quality etc.) and economic (cost per unit volume treated) aspects. Adin and Rebhun (1974) also found cationic polyelectrolyte better suited to high rate direct filtration than alum.

Molecular weight seems to be a significant factor when cationic polymers are used as primary coagulants. According to Stump and Novak (1979) very small cationic polymers (< 10 000 mol wt) tend to perform very poorly with regard to turbidity removal while very large cationic polymers (> 1 million mol wt) can cause excessive head losses. Therefore selection of polymers for direct filtration can be limited to cationic polymers with molecular weights between 10 000 and 200 000. Other researchers working with molecular weights of 600 to 100 000 found that a markedly better filter effluent was obtained with high molecular weight polymer, although head losses were appreciably higher and depth of penetration of floc decreased.

The other parameters widely held responsible for polymer performance are the rapid mix velocity Gradient $\bar{G}$ and the the detention time. In general, the polymers in the higher molecular weight range (> 100 000) perform best with intense mixing at 600 - 1 000 s^{-1}. Polymers of lower molecular weight do well with $\bar{G}$ values in the 300 s^{-1} range. Flocculation almost always improves polymer performance and a detention time of about 20 minutes can be expected to yield good results. (Stump and Novak, 1979)

Dosages of 0,1 - 5 mg/l for cationic polymer are usually required for direct filtration purposes (Culp, 1977).

7.3 Polyelectrolytes as flocculent aids

The essential function of polyelectrolytes as flocculent aids is not primarily of destabilization (this is effected by the metal coagulants) but rather of supplementing the orthokinetic flocculation process by altering floc characteristics. Filter aids should produce aggregates that will be large enough to be captured and strong enough to withstand shear in the filter voids (Bratby, 1980).

The reason for the inferior results often evident using metal coagulant alone during direct filtration is that the metal hydroxide flocks suffer lack of floc compressibility within the filter bed. Moreover, the flocs formed are too weak to withstand high shear forces resulting in early breakthrough of turbidity and shorter filter runs. Adin and Rebhun (1974) report that efficient filtration with alum alone was achieved only at filtration velocities of 5 - 10 m/h and with media of up to 0,6 mm grain size thus indicating that contact filtration with alum alone may not be efficient at high rates with coarse media. Furthermore, there is a proportional relationship between floc volume and metal coagulant dosage. Hutchison and Foley (1974) report that there is an almost inversely proportional relationship between the length of filter runs and metal coagulant dose.

Flocculation with polyelectrolytes as coagulant aids is characterised with extremely low doses, stronger attachment to the grains of the bed and a lower turbidity of the filtrate. Dosages may range from 0,05 to 0,5 mg/l (Culp, 1977). Table 1 shows the dosage range of some of the metal coagulants and the polymers when used together. The effect of the treatment in terms of length of filter runs and filtered water turbidity can be seen from the same table.

Table 1. Low dosage direct filtration (Wagner and Hudson, 1982)

Page	Plant Number	Coagulant Name	Coagulant Dosage mg/L	Polymer Name	Polymer Dosage mg/L	Filter Runs hours	Filtered Water Turbidity ntu
1	2	Alum	3 4-60			13 0	1 0
	3	$FeCl_3$	3 6			21 8	0 2
	1	Alum	3.7	Nalco	0.3-1	8 48	0 39
2	5	$FeCl_3$	3-6			15+	0 3
	6	Alum	5-6	Nalro C	1 3	5+	0 03-0 95
	15	Alum	3-35	Cat-T	0-1 5	20+	0 02-0 06
3	27	Alum	8-10	990N	0-0.5	8+	0 2-0 4
4	42	Alum	5-15 *			30+	1 0
	45			Nalco 607	0.5	240+	0 1-0 3
5	48	Alum	8.0	Seperan	0 1	5-10	
	52	Alum	3			24	0 35
	53			Nalco 607		24	0 19
	55	Alum	2	Nalco 607	0 5	48	0 2
6	57	Alum	4	Poly	1 0	72	0 5
	41	Alum	5-20	Polymer		12 36	
7	46	Alum	5-20	Cat T	1-2	244	0 35
	19	Alum	5-10			48+	0 2
8	65	Alum	2-9			2 5-8	0 13
	66	Alum	0 5-12	Poly	0 3-1 5	3 6-65	0 39
	57	Alum	0-24			15-56	0 15-0 6
9	22			Cat-Poly	2-3	16-36	0 2-0 6
	12			Cat-T	1-2	2 5-10	0.1-4
	32	Alum	6-25	990N	0-0 05	24-30	0 2
	36			Cat	2-3	8-12	<1
	38	Alum	0-100	Seperan	0 1	3-48	0 2-1
10	49			Cat T	0 1	50	
11	35			AmCyCat	2-3	8-12	< 1
12	39	Alum	1-15				
13	16			Nalco 607	0 6-10	19-72	0 25-0 7
16	61	$FeCl_3$	1 75			20-28	0 13-0 16
	60	Alum	3 6			39	[illegible]
				Nalco 607	0.4	21	[illegible]
17	20			Cat 7	1 0	10-20	0 25-0 6
18	64			AmCyCat	2-3	8-12	<1

*Data are from 1980 AWWA Subcommittee Report

Most laboratory studies that compare the effectiveness of cationic, anionic and non ionic polymers as coagulant aids indicate that of the three polymer groups tested cationic polymers achieve superior turbidity removal (Stump and Novak, 1979).

Further, Bratby (1980) remarks that the combined cost of metal coagulant plus flocculent aid is usually less than using cationic (primary coagulant) polymer alone although this depends upon the water quality.

7.4 Assessment

7.41 Advantages

The use of polyelectrolytes provides the following advantages:

1. Much lower dosages of polyelectrolyte would be required when it is used as a primary coagulant (figure 26).

2. It would appreciably reduce the dosage of the primary metal coagulant when used as a flocculant aid.

3. Improved filtration performance can be achieved, in both cases, after backwash operations (figure 27).

4. Reduction of the amount of soluble anions added with the coagulant is possible.

5. Carry-through into the distribution system associated with residual alumina (which causes fouling and "dirty water" complaints) can be avoided.

6. Volume of sludge produced will be reduced as it is more dense.

7. Perhaps the most appealing advantage of polyelectrolytes over metal coagulants is their effect on the cycle output at higher filtration rates. For instance, it has been reported that changing the approach velocity from 5 to 10 m/h decreased the cycle output by about 1/3 with alum and only by about 1/12 with polyelectrolytes. The need for backwash with alum was doubled. Changing the velocity from 10 to 20 m/h caused alum unsuitable by the accepted criteria. On the other hand, with polymer doubling the velocity increased the cycle output by 25 per cent and decreased the need for backwash by about 20 per cent. (Adin and Rebhun, 1974)

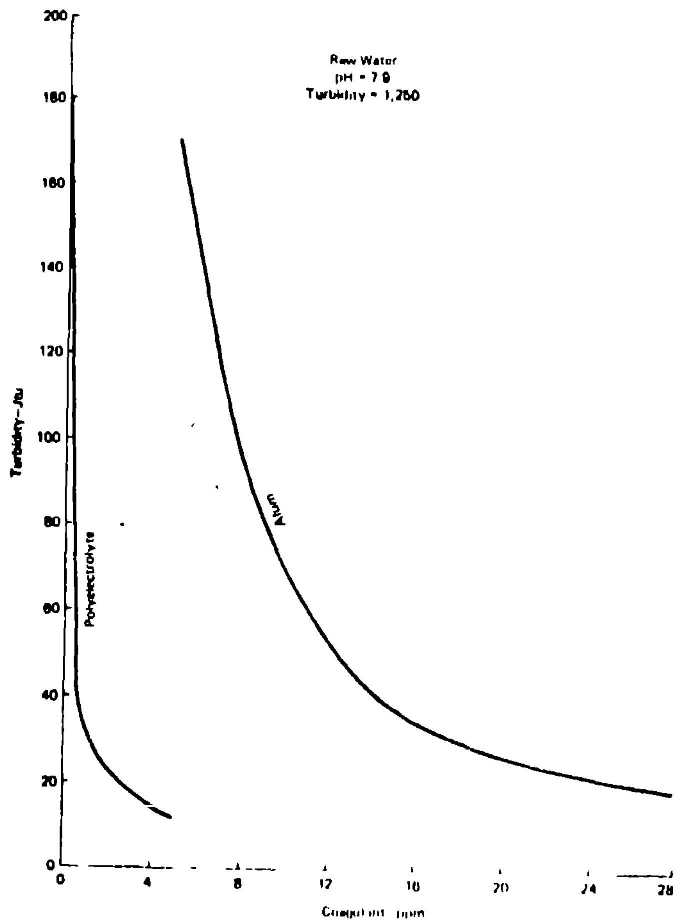


Figure 26. Comparison data of polyelectrolyte and alum dosages on turbidity removal (Beardsley, 1973)

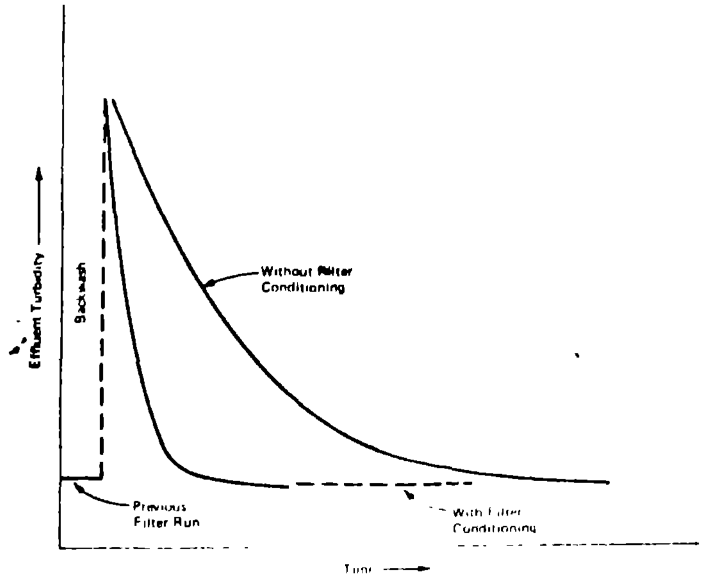


Figure 27. Effectiveness of polyelectrolyte in bringing filter "on line" after backwash (Beardsley, 1973)

8. Finally, it is noteworthy that the use of cationic polyelectrolytes as primary coagulants, in many cases, is less expensive than the use of alum. According to Shea et al (1971) 1,0 mg/l of polymer is equivalent in cost to 10 mg/l of alum. However, in terms of flocculating power, 1,0 mg/l of cationic flocculant may be equivalent to more than 10 mg/l of alum. The investigators found in their study the use of 1,0 mg/l Cat-floc resulted in filter performance far exceeding those obtained with 20 mg/l of alum. Also, additional savings can be expected in smaller chemical storage and application facilities according to the same researchers.

7.42 Drawbacks

Although in general an overall improvement in quality and economy of the process could be achieved with them, polyelectrolytes have some drawbacks. The following are a few of them:

1. The type and charasteristics of these polymers are so diverse (Bratby, 1980) that they cannot be indiscreminately applied. Testing of the intended polymer type in connection with the particular suspension to be treated is imperative for assessing its suitability.

2. These polymers may not be as readily and as cheaply available as metal coagulants (Bratby, 1980).

3. The dosing of polymers requires extra care. Overdosing of polyelectrolytes would result in surface clogging of the media and in an increase of operational costs. Since the dosage requirement is very small the optimization of the dosage of polyelectrolytes is bound to be a difficult one (Bratby, 1980).

4. The efficiency of destabilization of polyelectrolytes depends upon the velocity gradient of the rapid mixing operation. The velocity gradient in turn depends upon the molecular weight of the polymer. As the properties of the polymers are diverse, so also are the corresponding velocity gradient requirements. (Stump and Novak, 1979)

5. With synthetic products, although there is no evidence that polymerised species are of high toxicity the unpolymerised monomer species are (Bratby, 1980).

6. Since most products are defined virtually solely by brand name or number, valuable information concerning the type of polymer, molecular weight etc. linked to the particular water constituents is seriously limited (Bratby, 1980).

In the light of the above reasons, serious and careful consideration of the usage of polyelectrolytes is necessary.

8. ALTERNATIVE CONFIGURATIONS OF COAGULANT ADDITION FOR DIRECT FILTRATION

There are essentially three alternative configurations of coagulant addition for direct filtration (Bratby, 1980). The first is metal coagulant alone, added during rapid mixing and passed either directly or via a contact basin to the filter. The second is metal coagulant added as above but with a flocculant aid added (e.g. non-ionic polyelectrolyte or activated silica) just before filtration and the third is metal coagulant eliminated and replaced with, usually, a cationic polyelectrolyte serving as primary coagulant. Some of the common configurations are presented in figure 28.

In a discussion to the relevant merits of the above alternatives, it is necessary to review the removal process during filtration. The removal process comprises three stages: a working-in stage, a working stage and a breakthrough stage (Adin and Rebhun, 1974).

During the working-in stage the turbidity of the filtered water decreases rapidly until it reaches a stable low value. It seems that for efficient attachment the flocculent particles must provide an initial coating of the bed grains. The coat evidently increases the chance of efficient attachment. This stage is shorter with alum than with polymer (Adin et al, 1974).

The working stage, considered to commence when the working-in stage produces a stable low value, is the main phase during filtration. Adin and Rebhun (1974) found that during this stage polymers gave better effluent quality for all depths than was obtained with alum.

In the breakthrough stage the quality of the filtrate turbidity begins to rise beyond an acceptable preset effluent quality value (figure 29). The onset of breakthrough is usually faster using metal coagulants than polyelectrolytes (Adin and Rebhun, 1974).

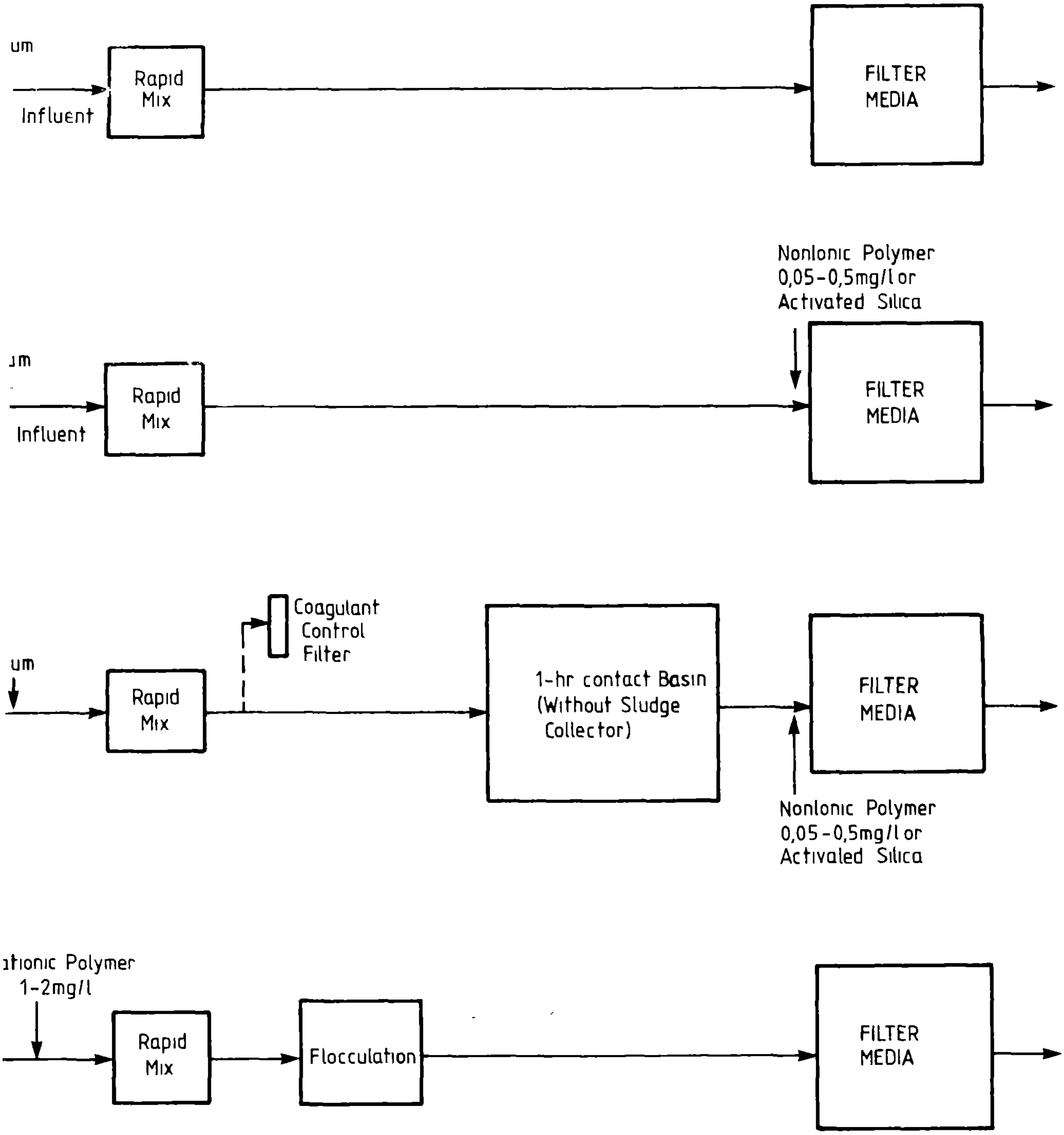


Figure 28. Some common configurations of coagulant addition (adapted from Culp, 1977).

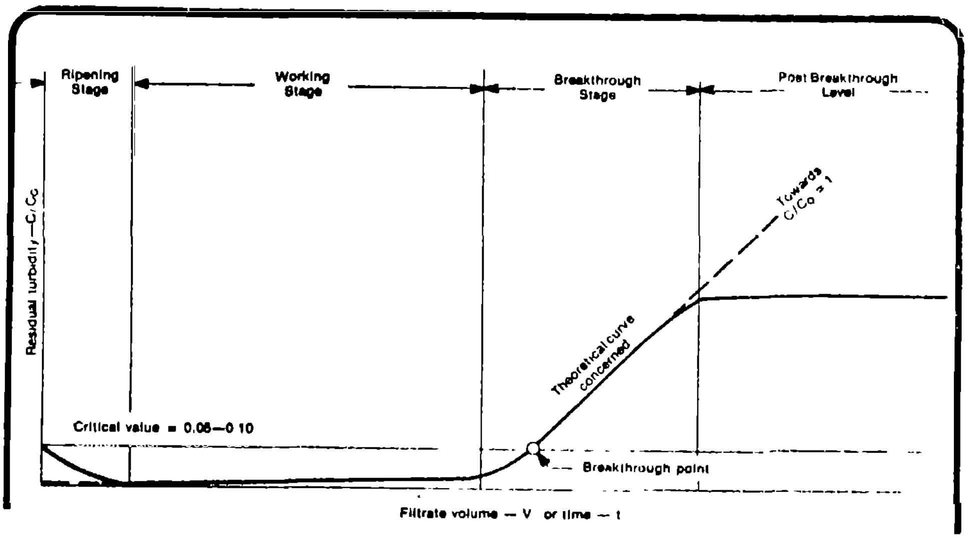


Figure 29. Graphic representation of filtration stages (Adin and Rebhun, 1977)

The work of the bed may be described through a frontal advancement of the working layer (figure 30) in which effective filtration is taking place. "Saturation" of one layer is followed by continued activity in the next. The "saturated" layer removes at a constant and poor efficiency a certain small part of the solids. The main removal process takes place in the "working layer". The "working layer" is characterized by a high removal coefficient and steep concentration gradient. This front advances more rapidly with alum than with cationic polymer. Increasing the grain size causes a faster advance of the front (penetration) and a decrease in head loss. These phenomena were observed and explained by Adin and Rebhun (1974).

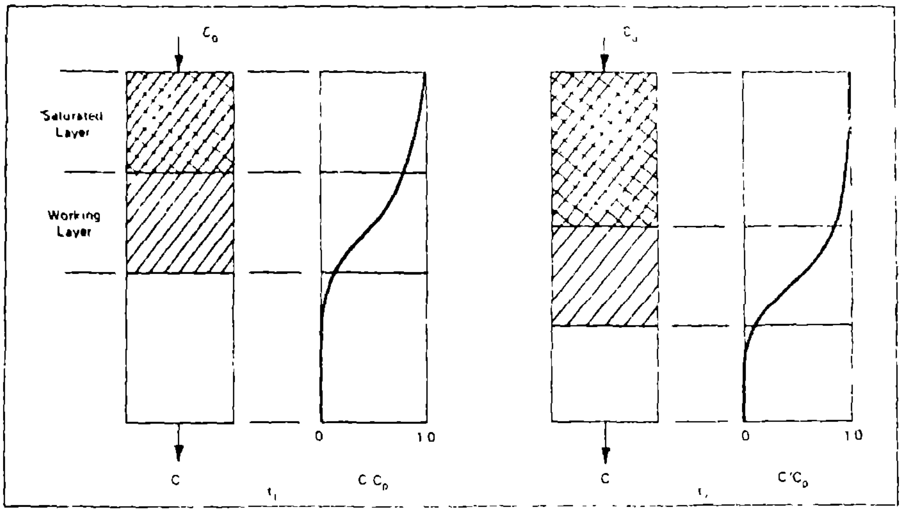


Figure 30. Progression of working layer through a filter (Adin and Rebhun, 1974)

8.1 Flocculation and direct filtration

Flocculation as defined in the section on coagulation is the term which describes the subsequent process where the small coagulated particles are built into larger aggregates or flocs which may easily be removed by sedimentation and/or filtration processes. In flocculation, collisions or very near approaches between particles are necessary for adhesion to occur and aggregates to form.

Habibian and O'Melia (1975) state that particle transport in flocculation and in filtration processes can be considered separately in two distinct regions - the perkinetic and orthokinetic regions.

In perkinetic flocculation Brownian motion is the prime factor controlling the transport and eventually the aggregation rate of particles smaller than about 1 μm. Diffusion is promoted by temperature and concentration gradients. This is a naturally random process.

When the particles have aggregated to be larger than 1 μm, perkinetic flocculation ceases to be significant, and particle collisions must be induced by velocity gradients. This enhances the relative motion of the particles thereby increasing the opportunities of contact by means of hydrodynamic transport mechanisms such as interception and sedimentation. This latter type of flocculation is known as orthokinetic flocculation.

From the foregoing discussion it is apparent that the transport mechanisms in flocculation and filtration have remarkable similarities. In both processes particle transport can be considered in the perkinetic region for particles of less than 1 μm and in the orthokinetic regions for those over 1 μm.

However, an important difference arises in comparing volume flocculation and flocculation in filtration. This difference lies in the detention time. Following the rapid mixing of coagulants with raw water, destabilization takes place in less than 1 sec (Culp, 1977). At this point in the process the colloids are susceptible to stick together after collision. The rate of agglomeration of these microscopic destabilized particles to form remarkable floc is dependent principally upon the number of opportunities for contact that are afforded.

In a still body of water agglomeration takes place at a slow, almost imperceptible rate. Perkinetic flocculation is probably significant in this case. The rate can be increased by aggitation or orthokinetic flocculation. In conventional volume flocculation the slow mixing device is responsible for promoting particle contact. In a well designed flocculation basin agglomeration may take from 5 to 45 minutes (Culp, 1977). The detention time which must be provided to achieve a given degree of aggregation by volume flocculation increases as the concentration of particles to be aggregated decreases.

8.11 Flocculation unit

The need for a slow mixing or flocculation basin and the related detention time before direct filtration has been a matter of disagreement. Theoretically (O'Melia et al, 1975; Adin et al, 1979) filtration is capable of effectively removing particles of any size if the attachment step is effective. Particles smaller than 1 μm would be removed efficiently by Brownian diffusion and those bigger than 1 μm by hydrodynamic and gravity forces. Furthermore, flocculation takes place in the filter media at an accelerated rate due to the tremendous number of opportunities afforded during the transport step. This formation of floc within the interestices of the bed has been experimentally detected (Shea et al, 1971). Thus removal of particles of even 1 μm size is inevitable during the transport step if attachment is effective. This theory implies that no prior flocculation is necessary before direct filtration. Laboratory experiments (Adin and Rebhun, 1974) investigating a scheme in which hydraulic rapid mixing alone was used gave good results. Culp (1977) points out the redundancy of the flocculation basin for direct filtration. Ghosh et al (1981) state explicitly that for most direct filtration operations slow mixing following coagulation may not be needed if the suspended solids concentration is 30 mg per litre or higher.

Monscvitz et al (1978) and Treweek (1979), however, indicated that the flocculation basin is necessary to achieve the desired level of treatment. Concerning the detention time for flocculation, Letterman et al (1979) state that a short period of flocculation (2 to 10 minutes mean detention time) results in good filtration efficiency. Treweek (1979) found that flocculation time shorter than 7 minutes was not sufficient to produce aggregates removable in filter media. Hutchison and Foley (1974) reported that the flocculation times should be greater than 3,5 minutes to prevent breakthrough and rapid head loss developments. They indicated that with water temperatures of less than 3,3 $^{\circ}$C, flocculation times longer than 10 minutes might be in order to prevent after floc formation. Other workers (McCormick et al, 1982) recommended that flocculation time should be varied from 10 minutes during hot weather to 30 minutes during cold weather. Still other researchers working with temperatures of 9 ^{o}C reported (McCormick et al, 1982) that increasing flocculation time from 13 to 26 minutes was not accompanied by improved water quality.

It is not easy to make generalizations about the need for a flocculation basin or the corresponding detention time. The findings and relevant recommendations pertain to the individual circumstances under which the investigations were carried out. Nonetheless, the need for a flocculation basin with the use of polyelectrolytes seems likely. In fact, the recommendations for a flocculation basin have been made in connection with experimental works that employed polyelectrolytes either as main coagulants or coagulant aids. This is possibly because of the slow diffusion rate of polyelectrolytes. Currently available means of rapid mixing do not probably ensure uniform and fast enough dispersion of the polyelectrolytes into the water to be treated. Thus a slow mixing or flocculation basin may be required to provide the appropriate detention time and ensure adequate dispersion and satisfactory destabilization. On the contrary, diffusion and destabilization take place relatively faster with metal coagulants. Therefore flocculation units may be dispensed with for direct filtration operations using metal coagulants alone.

8.12 Contact basin

Some direct filtaration plants that draw the raw water from sources of erratic quality changes especially with regard to suspended solids content or high concentrations of coliform organisms reportedly (Culp, 1977) use a contact basin. In such cases the reliability of the treatment process is improved by incorporating a coagulant-control filter with a recording turbidimeter monitoring the control-filter effluent. The advantage of this system is that the coagulant requirements are determined 1 h in advance of the water reaching the plant filters. This provides much more time for dosage adjustments without possible adverse effects on product water quality. Thus plant operations are kept abreast of changes in raw water quality.

With the coagulant-control filter, the contact basin and monitoring of filter effluent turbidity, the reliability of the direct filtration process is at least as good as that of conventional water treatment plants within the limits of raw water quality appropriate for application of direct filtration. There are cost savings even with the use of a contact basin in direct filtration because the costs for flocculation basin and equipment and sludge collection equipment are eliminated; the size of the contact basin is much smaller than that of a basin required to accomplish settling (Culp, 1977).

9. EVALUATION OF DIRECT FILTRATION

9.1 Advantages

Tredgett (1974), Culp (1977) and Bratby (1980) state that the chief advantage of direct filtration is the potential for capital cost savings up to 30 - 35 % in cases where the raw water quality is appropriate, as compared to the conventional sedimentation-filtration process. The cost saving results from the elimination of sludge-collecting equipment, settling basin structures, flocculation equipment and sometimes from flocculation-basin structures.

With direct filtration there may also be savings of 10 - 30 % in chemical costs (Culp, 1977) because generally less alum is required to produce a filterable floc than to produce a settleable floc. Although flocculant aid dosages (if used) may entail costs higher than in conventional plants, the higher associated costs are more than offset by the lower costs for coagulant.

Further advantages are that operation and maintenance costs are reduced because there is no equipment to operate and maintain.

Direct filtration produces less sludge than conventional treatment, and the sludge is more dense. The collection of waste solids is simplified because all waste solids are all contained in a single stream, the waste filter-backwash water.

9.2 Limitations

The direct filtration process may not be applicable to raw waters having turbidity greater than 100 - 200 tu, colour greater than 100 units, colour and turbidity each greater than 25 units, plankton exceeding 500 - 1 000 asu/ml, or appreciable amounts of paper fiber (Culp, 1977).

Filter runs are generally shorter in direct filtration than in filtration preceded by settling. The cost consequences of this may not be too significant, but the ability to handle suspended solids in direct filtration is limited. There is a point where operational problems may result in some cases such that it would be better to reduce the load to the filters by introducing settling or simple roughing filtration in the process chain.

Perhaps the main potential disadvantage with direct filtration is the short time lag between coagulant addition and filtration requiring a higher standard of control or operator vigilance (Spink and Monscvitz, 1974). The chances of operator error may be increased. In the treatment of raw waters containing high concentrations of coliform organisms, the reliability of public health protection may be reduced. In the great majority of cases, however, Culp (1977) mentions that this is not a factor.

Washwater usage in direct filtration plants may be as high as 6 per cent as compared to 2 per cent for backwash plus 2 per cent for sludge wasting, a total of 4 per cent in a conventional plant treating similar raw water. This difference, however, is not a significant item in overall treatment-plant operating costs (Culp, 1977).

10. HORIZONTAL ROUGHING FILTRATION

With proper grading of the filter media, rapid filters perform as deep-bed filters. Such filters allow deeper penetration of the suspended matter and therefore provide some silt storage capacity. Though rapid roughing sand filters are advantageous in this respect by providing long filter-runs, they nonetheless require expensive and sophisticated underdrain and backwash facilities. This is because with the rapid roughing sand filters the filter bed needs to be fluidized by a backwash process using high wash-water rates and even compressed air to support the scour of the sand grains.

Horizontal roughing filtration, on the other hand, obviates the need of complicated underdrain and backwash facilities. An additional advantage afforded by horizontal roughing filtration is the possibility of using much more coarser media than what can be used in rapid roughing sand filters. The use of such coarser media eventually results in considerably more silt storage capacity as well as extended filter runs. Moreover, while the height of upward and downflow rapid roughing filter structures is limited due to structural and economic constraints, horizontal roughing filtration gives a practically unlimited filter length. Horizontal roughing filters are simple in construction and need no highly skilled attendance for their operation. As such horizontal roughing filtration appears qualified for an appropriate and self reliant pretreatment of surface waters prior to slow sand filtration (Wegelin, 1981).

10.1 Experience with HRF

There are reportedly (Wegelin, 1981) several water treatment plants in Europe using horizontal roughing filters. Although the roughing filters are being run at high filtration rates (5 - 10 m/h), the suspended solids contents are on the average less than 10 mg/l (Kuntschik, 1976; Wegelin, 1981).

Unlike the rivers in moderate climates the rivers in tropical countries mostly carry much higher loads of suspended solids. Investigations have been initiated and carried out in the developing countries in an effort to adopt horizontal roughing filtration to such different raw water qualities and to meet other local conditions.

The prominent examples are the investigations carried out at the Asian Institute of Technology in Bangkok, Thailand and at the University of Dar es Salaam, Tanzania. The studies were made in connection with the reduction of suspended solids for subsequent slow sand filtration with HRF as pretreatment (Wegelin, 1981).

Of particular interest here are the findings of the tests in Tanzania. They show that the removal efficiency of the HRF in respect of turbidity removal is not much influenced by the gravel surface (Wegelin, 1981). They also show that the efficiency is related to the Reynolds number (Riti, 1981; Wegelin, 1981).

10.2 Mechanisms of horizontal filtration

There is no indication that the mechanisms of horizontal flow filtration are in any way different from the mechanisms of filtration discussed already. It seems that both the transport and attachment forces are responsible in the removal process. Of interest is, however, the role of mechanical straining. Although deep bed filters are specifically designed to avoid mechanical straining, this inevitably takes place in horizontal roughing filters (Riti, 1981) due to the fact that floating solids which have escaped the coarse screen at the intake get strained in the filter. According to Huisman (1977) with grain size D1 (figure 31) the pore size d is given by d = 0,155 D. The minimum diameter of particles to be strained by the coarse gravel, for instance for grain sizes of 18 - 32 mm, is about 3,9 mm. Whereas the finest gravel of say 4 - 8 mm grain size straines particles bigger than 0,93 mm. However, in most cases natural waters contain only colloidal particles which are much smaller in size (table 2). Under such circumstances the role of mechanical straining may be insignificant and deep bed filtration dominates.

Table 2. Sizes of materials involved in water treatment (Riti, 1981)

Material	Particle diameter u (1 u = 0,001 mm)
1. Sand	500
2. Soil (clay, silt, loam)	1 - 100
3. Bacteria	0,3 - 3,0
4. Viruses	0,005 - 0,01
5. Floc particles	100 - 2000

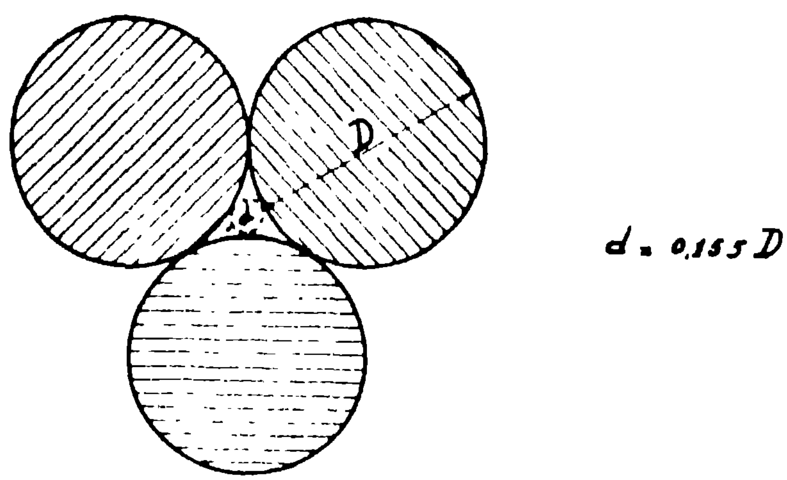


FIG. 31 RELATION BETWEEN GRAIN SIZE AND PORE SIZE (Huisman, 1977)

10.3 HRF as pretreatment for direct filtration

Horizontal coarse gravel filtration as a pretreatment for various purposes is probably not a new idea. But its application as a pretreatment unit operation with regard to direct filtration has not been indicated in the literature surveyed. Nonetheless, its use in this respect is anticipated to yield good results in the reduction of chemical costs, elimination of the conventional settling units, producing a more or less constant quality water to the filters thereby controlling shockloads, and last but not least, lengthening the filter runs of the subsequent filter units.

The HRF itself takes a very long time to clog. The observations made with low turbidity waters indicate that filter runs of as long as 3 - 5 years could be expected (Riti, 1981; Wegelin, 1981). If not as long, at least reasonably long filter runs could be expected with high turbidity waters as well.

In any case, the cleaning of the HRF could easily and cheaply be carried out using manual labour. In the case of their use in connection with direct filtration the wash water pumps of the downflow filters could be employed to flush out the accumulated dirt from the HRF at higher hydraulic loading rates using clean water. It is preferable for this purpose if the bottom of the HRF were constructed sloping at about 1 : 100. This would also provide easy drainage when the filter has to be taken out of service for cleaning (Riti, 1981).

The applicability of horizontal flow coarse material pre-filtration of highly turbid surface waters especially in areas where the geologic formation does not encourage the use of infiltration galleries is bound to be of practical relevance and economic significance.

11. INVESTIGATIONS ON A PILOT HRF

The investigations were carried out on a pilot HRF of about 1,0 m width, by 1,0 m height and 9,0 m length. The pilot HRF was located in Oulu (North Finland). It was planned and constructed by Hanhimäki (1983). The pilot HRF was needed for studying the use of HRF in limiting suspended solids load that arises during excavation works on rivers which would otherwise pollute the waters downstream. The photograph of the pilot HRF is presented in figure 32.

11.1 Aims of the tests

The investigations on the pilot HRF were carried out to study the possibility of using horizontal coarse gravel filters with highly turbid waters as a pretreatment preceeding direct filtration. The purification efficiency under tropical conditions of such filters especially their capacity to reduce high turbidity and suspended solids concentration remains to be proved (Wegelin, 1981). Due to this the tests were aimed at investigating the variables that have important bearing on the removal efficiency especially with regard to suspended solids content. The variables investigated included the following:

- type of media
- Reynold's number
- variation of suspended solids load
- variation of hydraulic loading rates
- length of filter.

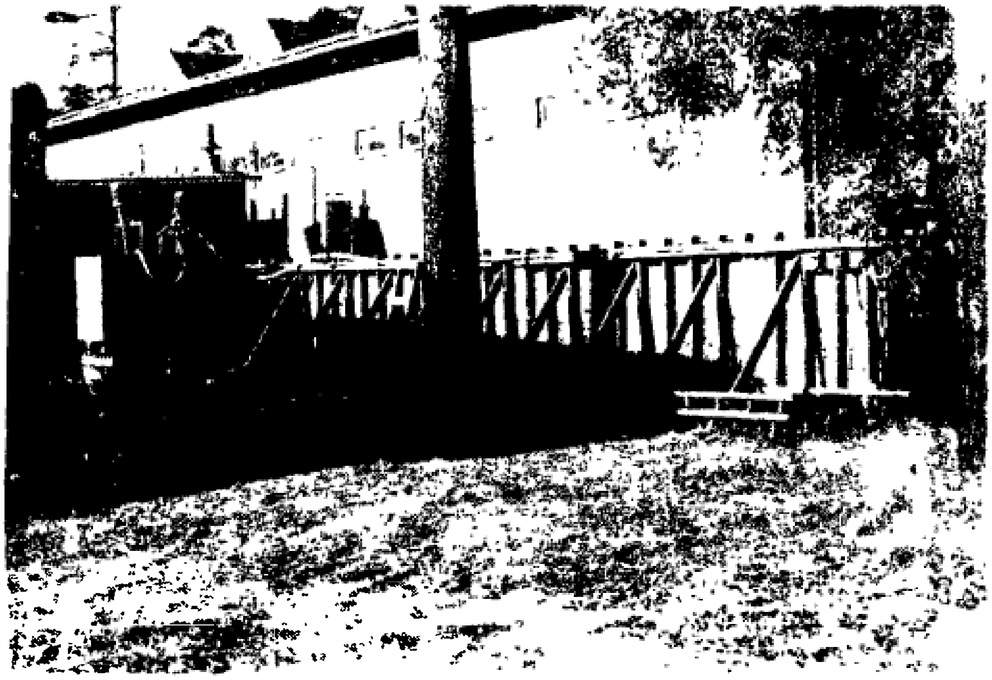


Figure 32. The pilot HRF

11.2 Description of the pilot HRF set up

The general layout of the pilot HRF consists of the filter box, the mixımg basin for preparing clayey water, outlet pipe and pumping units for recirculating the effluent and for providing backwash water. There is in addition a submersıble pump for pumping out filter washings from the ınlet end of the filter box structure so as to attain effective washing results. Figure 33 shows the layout plan and cross-section of the pilot filter.

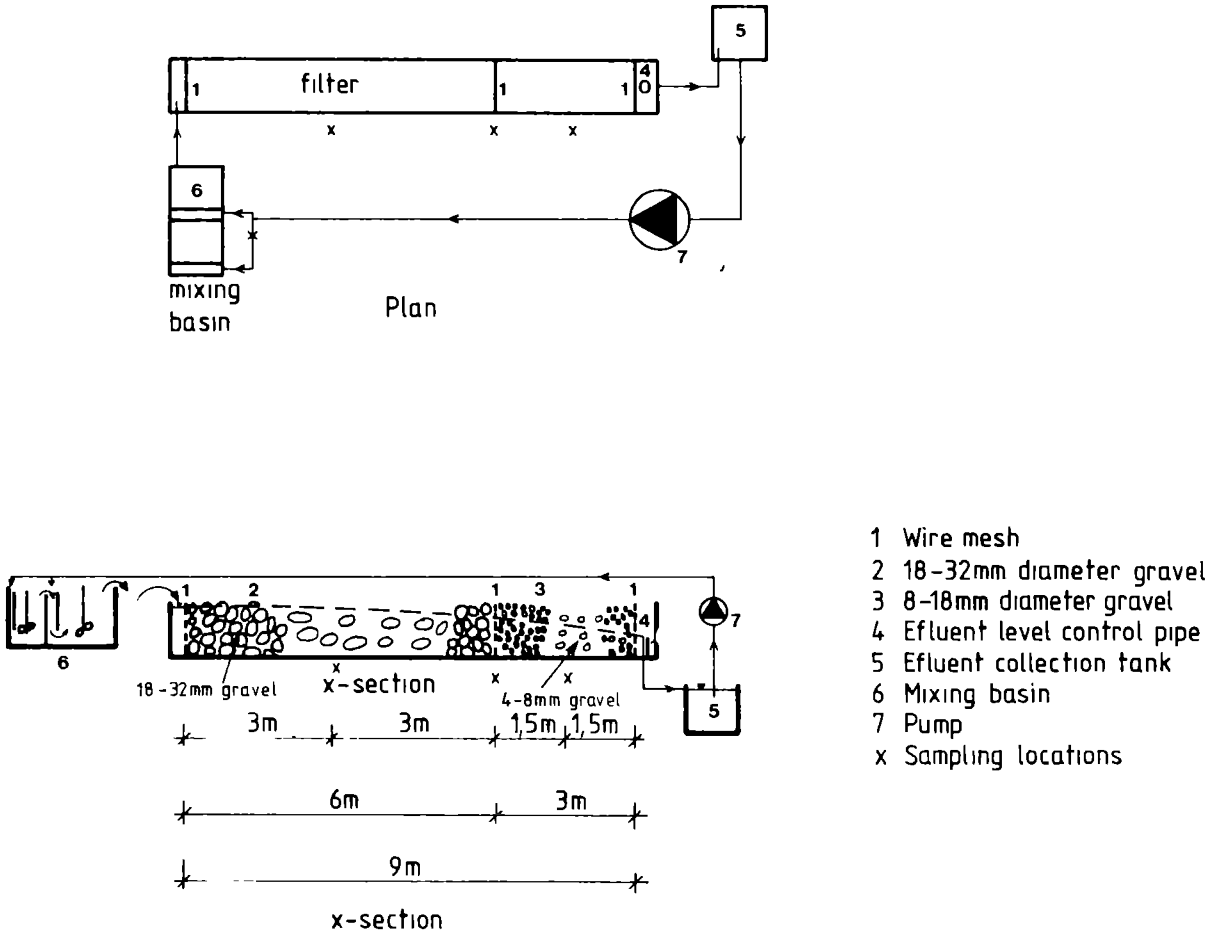


Figure 33. Schematic diagram for pilot filter layout and cross-section

The first 6 m of the horizontal filter box is filled with coarse natural gravel of size 18 - 32 mm whereas the last 3 m section consists of fine crushed gravel of 4 - 8 mm grain size. The grading of both media has been presented in figure 34. The uniformity coefficient d_{60}/d_{10} of the coarse gravel is about 1,3 and that of the fine gravel about 2.

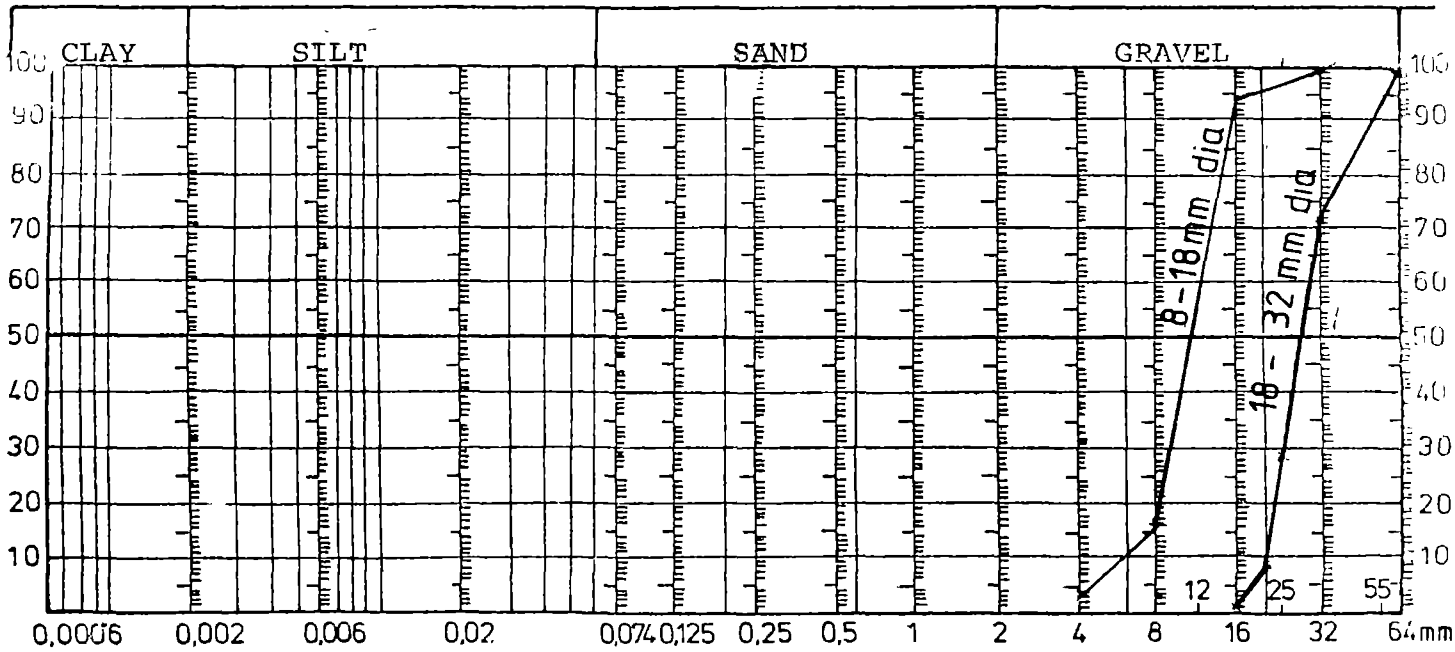


Figure 34. Grading of the coarse and fine gravel used in the horizontal filter during the investigation

The picture on figure 35 shows the coarse and fine gravel as placed in the filter. The porosity of the coarse gravel has been measured to be 0,38.

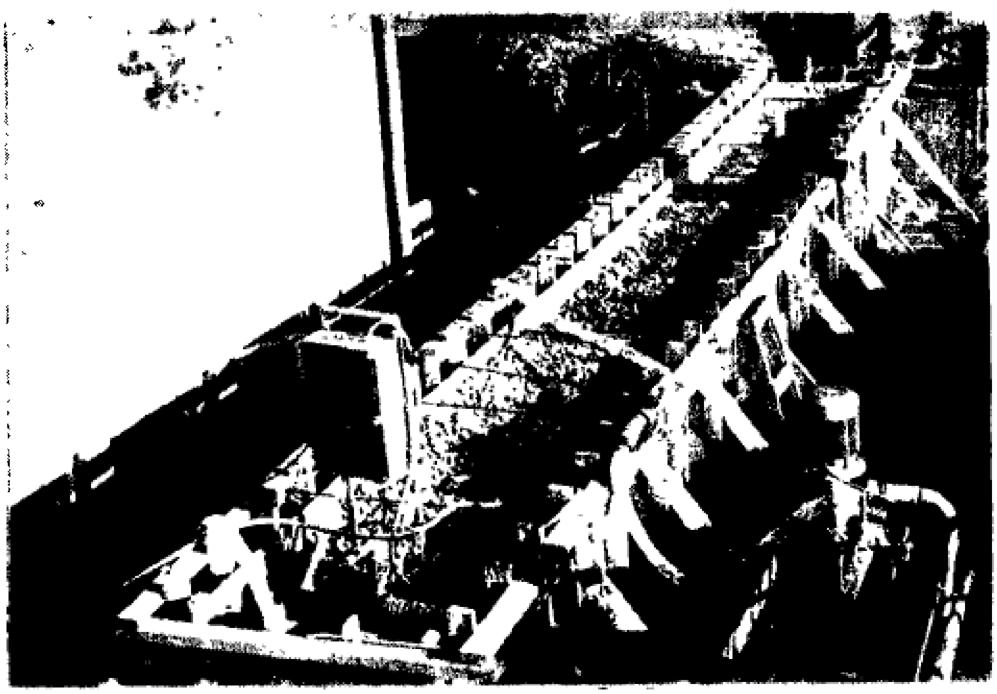


Figure 35. Filter media in place

Figure 36. Close-up view of the coarse gravel (18 - 32 mm size)

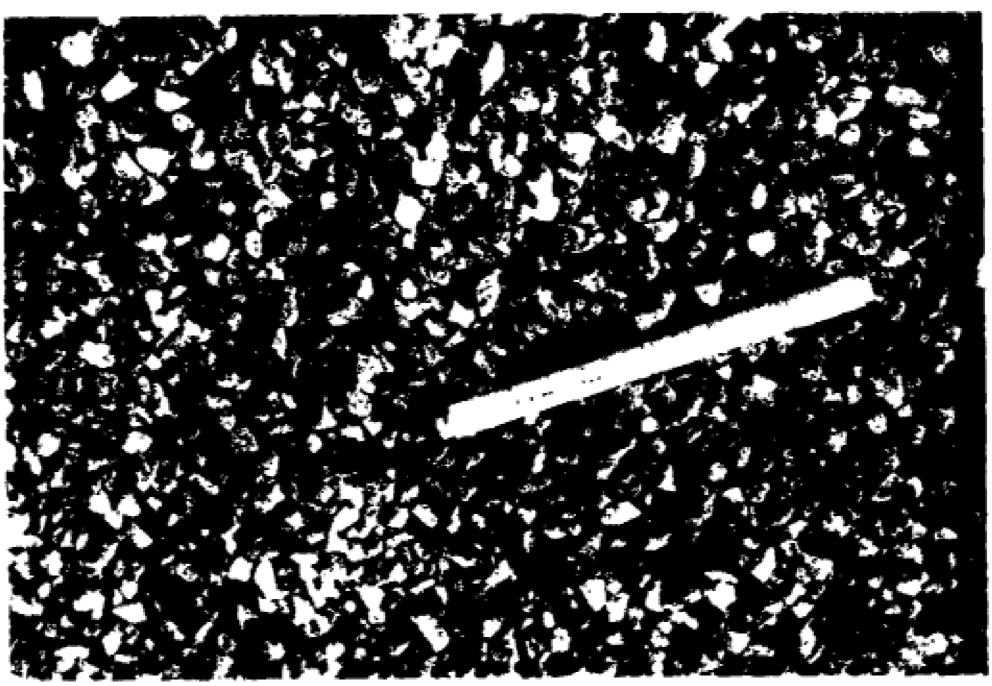


Figure 37. Close-up view of the fine gravel (4 - 8 mm size)

The inlet zone, the coarse gravel, the fine gravel and the outlet zone are respectively separated from each other by wire mesh partitions.

The filter box in addition consists of inlet, outlet, overflow and backwash piping arrangements. The hydraulic gradient in the filter is monitored by eight piezometric tubes coming from four different locations and placed againts the wall over a millimeter paper. Each pair of piezometers is placed on opposite side of the filter partitions.

The mixing basin consists of four distinct zones. The first zone is an inlet zone for part of the recirculated water. The second zone is the mixing unit for mixing water and clay. It contains one of the mixers. The third zone is an inlet zone for most of the recirculated water and the made up water. The fourth zone is an addtional mixing unit. In this zone some of the made up water that overflows into the third zone in controlled proportions is thoroughly mixed with the relatively clean water until the required turbidity is met. The stirrer in the last zone keeps the concentration of particles more or less constant. The piping system that leads the recirculated water is equipped with a gate valve on the branch which goes to the first zone. The opening of the gate valve increases the overflow rate into the third basin thereby reducing the flow of relatively clean water into the same as a result of which the concentration of the made up water in the fourth zone increases accordingly. A sketch of the cross-section of the mixing basin is shown on figure 38.

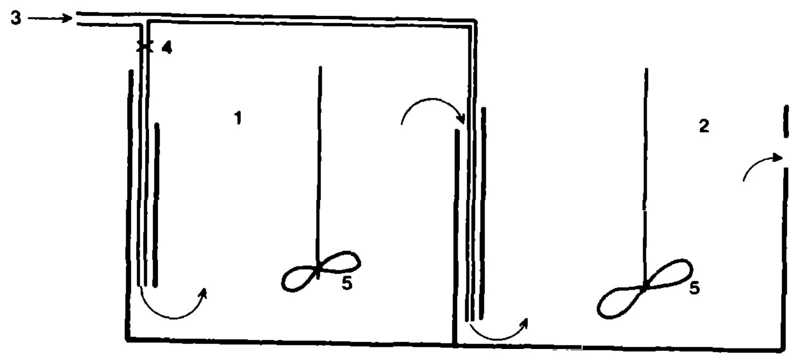


Figure 38. Sketch of the cross-section of the mixing basin

On the wall of the fourth basin has been placed a graduated scale for controlling the hydraulic loading rate. The scale readings corresponding to expected hydraulic loading rates were calibrated by Hanhimäki (1983).

The pumps for backwashing and recirculating the effluent are capable of providing a maximum of about 5 litres per second.

11.3 Procedures for testing the HRF

Before carrying out any tests on the HRF it was backwashed clean. The filter was then run overnight with clean water at rates of about 15 m/h. The overflow rate was then adjusted as necessary and run until the piezometric readings stabilized. The amount of clayey water overflowing into the third unit was controlled until the resulting mixture in the fourth unit attained the required degree of turbidity. Next, the temperature, scale reading and piezometric readings were recorded. The test samples from the different sampling locations were taken at one hour intervals. Samples taken from all locations except the inlet and outlet points were collected in one bottle from three points in the same sampling location. The points deliver samples from the bottom, the middle and top sections of the filter.

The samples were finally analysed for content of suspended solids and turbidity. No colour analysis was necessary since the tap water used contained less colour than 5 mg pt/l and the clay contained only 2,82 % of humus. The grading of the clay used for made up water is shown in figure 39.

The results of the investigations on the horizontal roughing filter are presented on table 3 and the results pertaining to the best performance have been plotted in figures 40, 41 and 42. Turbidity values for samples containing extremely high suspended solids (over 1000 mg/l) have been left out. This is due to the unreliability of the readings for such cases. The plotting of the water levels for the different hydraulic loading rates according to the corresponding piezometric readings have been shown on figure 57.

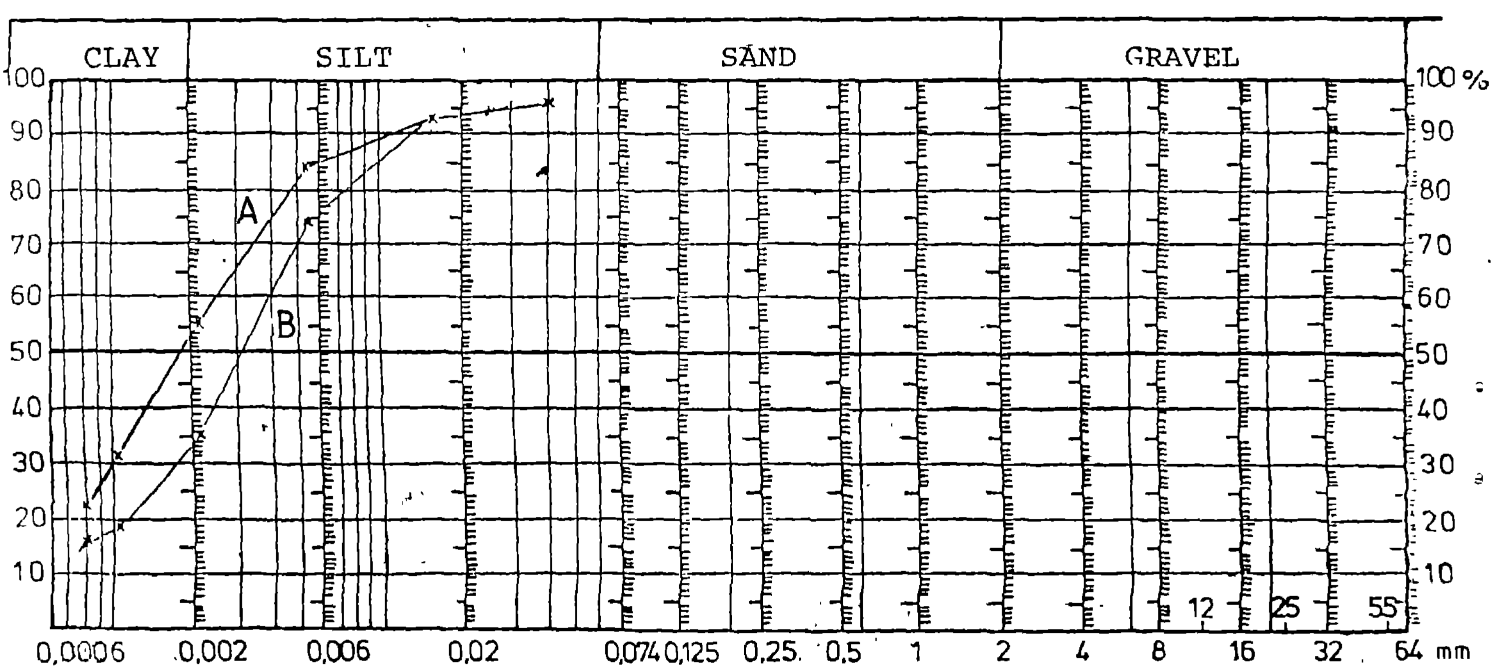


Figure 39. Grading of the clay used for made up water. Two samples of the same clay A and B were taken.

Table 3. Test results from the HRF

Date	Temp. (^{o}C)	Nominal loading rate ($m^3/m^2/h$)	Influent		3 m from influent (Gravel Ø 18-32 mm)		6 m from influent (Gravel Ø 18-32 mm)		7,5 m from influent (Fine gr. 4-8 mm)		Effluent at 9 m		Reduction in suspended solids		
			SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	Total (%)	Coarse (%)	Fine (%)
2.10.83	+7	5	208	180	163	150	138	130	108	125	90	120	56	33	23
			755	350	492	310	400	290	307	265	252	240	66	47	19
			1813	-	1340	-	1011	-	843	-	820	-	54	44	10
			2568	-	2033	-	1625	-	1520	-	1315	-	48	36	12
			5010	-	3261	-	2842	-	2700	-	2166	-	56	43	13
3.10.83	+3 - +4	10	571	280	486	280	369	260	330	250	333	270	41	35	6
			2565	-	2095	-	1595	-	1604	-	1619	-	37	37	-
			2311	-	2172	-	1714	-	1666	-	1547	-	33	26	7
			2735	-	2577	-	2247	-	2135	-	1933	-	29	18	11
			2266	-	3320	-	2712	-	2460	-	2533	-	-	-	-
4.10.83	+2 - +4	15	284	130	458	200	456	215	236	130	135	100	53	-60	113
			1688	-	1731	-	1296	-	1332	-	1367	-	19	23	-4
			4455	-	4019	-	3365	-	3675	-	3978	-	11	25	-14
			6092	-	3446	-	5260	-	5235	-	4653	-	24	14	10
			5428	-	5084	-	4731	-	2793	-	4566	-	16	-13	3

Table 3. Cont'd

Date	Temp. (°C)	Nominal loading rate ($m^3/m^2/h$)	Influent		3 m from influent (Gravel Ø 18-32 mm)		6 m from influent (Gravel Ø 18-32 mm)		7,5 m from influent (Fine gr. 4-8 mm)		Effluent at 9 m		Reduction in suspended solids		
			SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	Total (%)	Coarse (%)	Fine (%)
5.10.83	+6	15	970	-	1186	-	1097	-	1175	-	1241	-	-28	-13	-15
		10	438	270	482	270	490	285	436	270	285	235	35	-12	47
		5	412	265	489	300	452	300	392	280	315	240	23	-9	32
6.10.83	+6	5	622	270	363	230	392	240	283	210	217	160	65	37	28
		10	634	370	460	370	570	310	479	270	503	280	20	10	10
		10	972	400	1022	380	982	380	808	390	685	370	30	-1	31
		15	1702	-	1560	-	1574	-	1613	-	1955	-	-15	7	-22

Table 3. Cont'd

Date	Temp. (°C)	Nominal loading rate ($m^3/m^2/h$)	Influent		3 m from influent (Gravel Ø 18-32 mm)		6 m from influent (Gravel Ø 18-32 mm)		7,5 m from influent (Fine gr. 4-8 mm)		Effluent at 9 m		Reduction in suspended solids		
			SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	SS (mg/l)	Turb. (NTU)	Total (%)	Coarse (%)	Fine (%)
8.10.83	+4	15	116	70	110	70	101	56	126	70	176	68	-52	13	-39
			555	300	369	290	508	270	470	250	443	250	20	8	12
			1373	-	1273	-	1015	-	1069	-	1016	-	26	26	0
			1684	-	1644	-	1496	-	1507	-	1446	-	14	11	3
			1736	-	1587	-	1616	-	1550	-	1500	-	13	7	6
			1700	-	2009	-	1768	-	1700	-	1544	-	9	0	9
9.10.83	+1	5	134	100	138	100	177	110	91	79	56	68	57	-32	89
			112	83	104	87	183	100	78	70	35	57	68	-64	132
			221	130	223	130	215	130	144	120	99	100	55	3	52
			270	160	255	160	193	140	166	130	109	120	59	28	31
			394	240	368	220	343	210	193	170	168	170	57	13	44
			446	280	361	220	334	220	254	180	198	170	55	25	30
			752	320	495	300	514	280	396	260	300	230	60	32	28

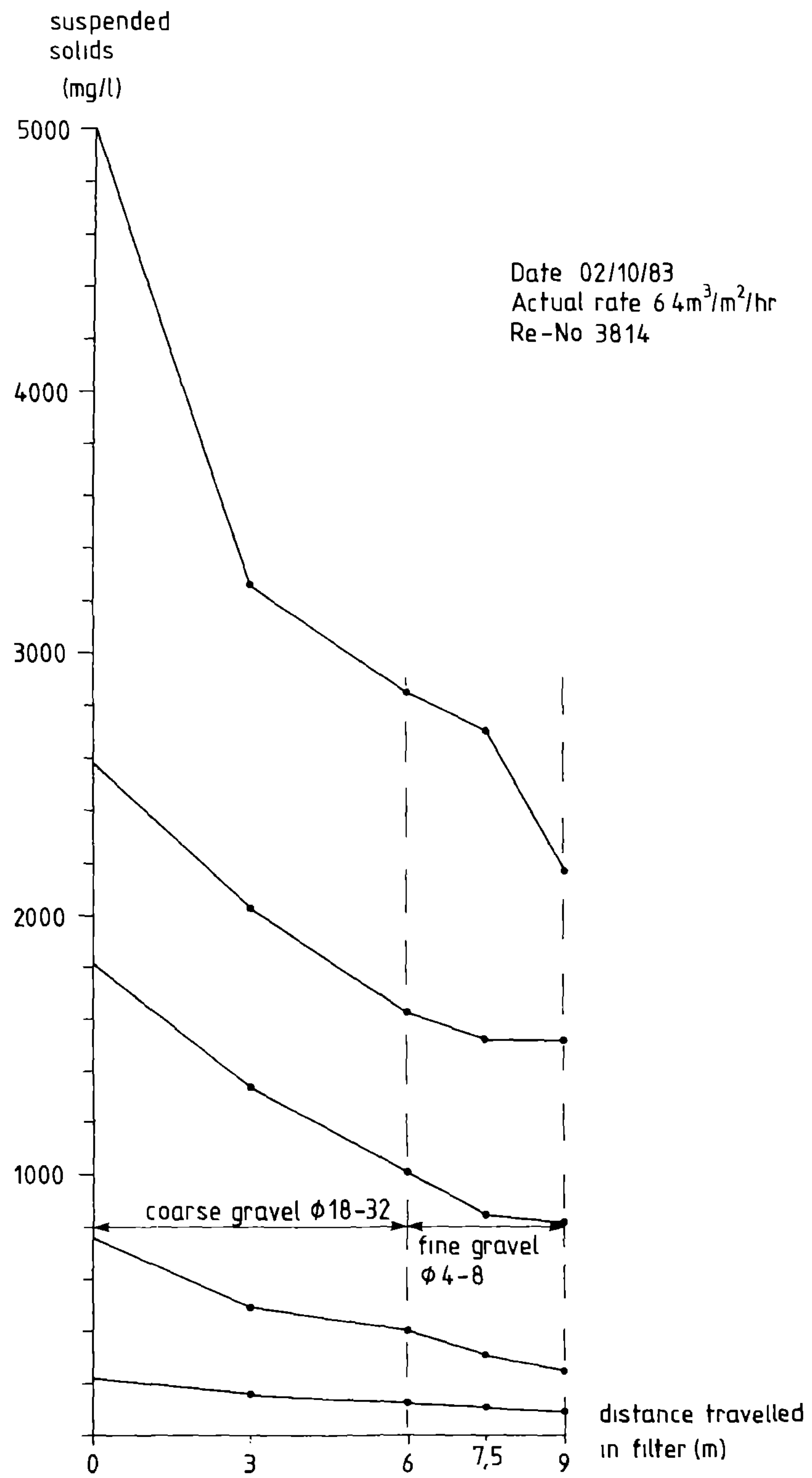


Figure 40. Pilot filter test. Suspended solids removal

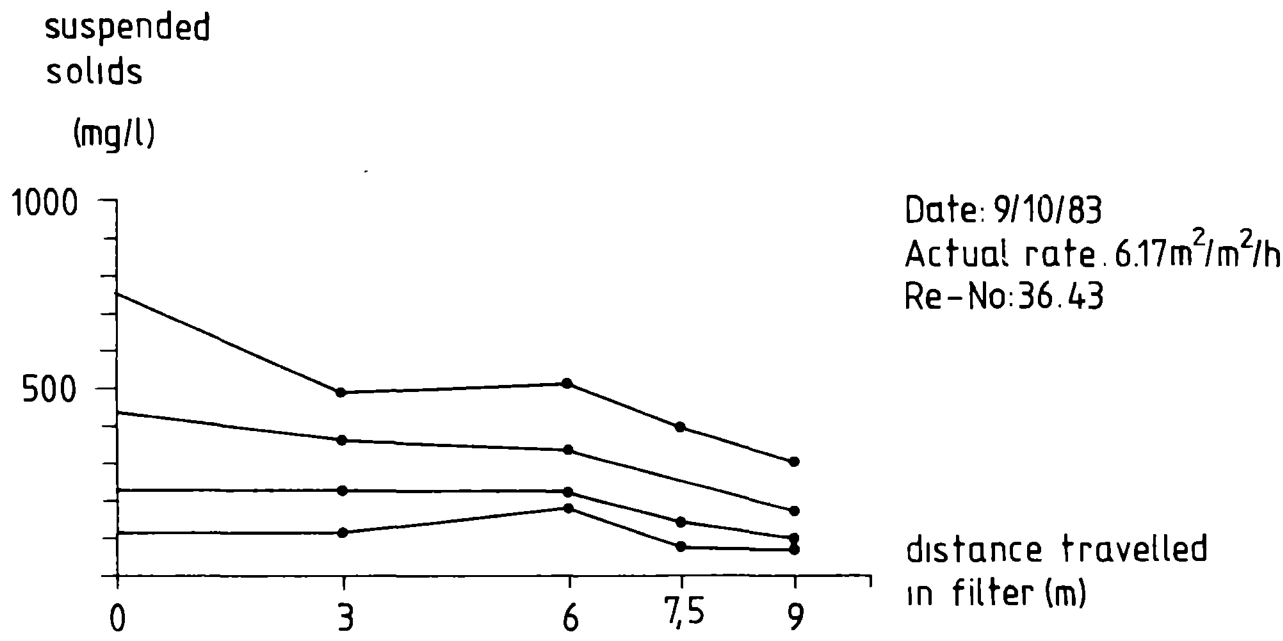


Figure 41. Pilot filter test. Suspended solids removal

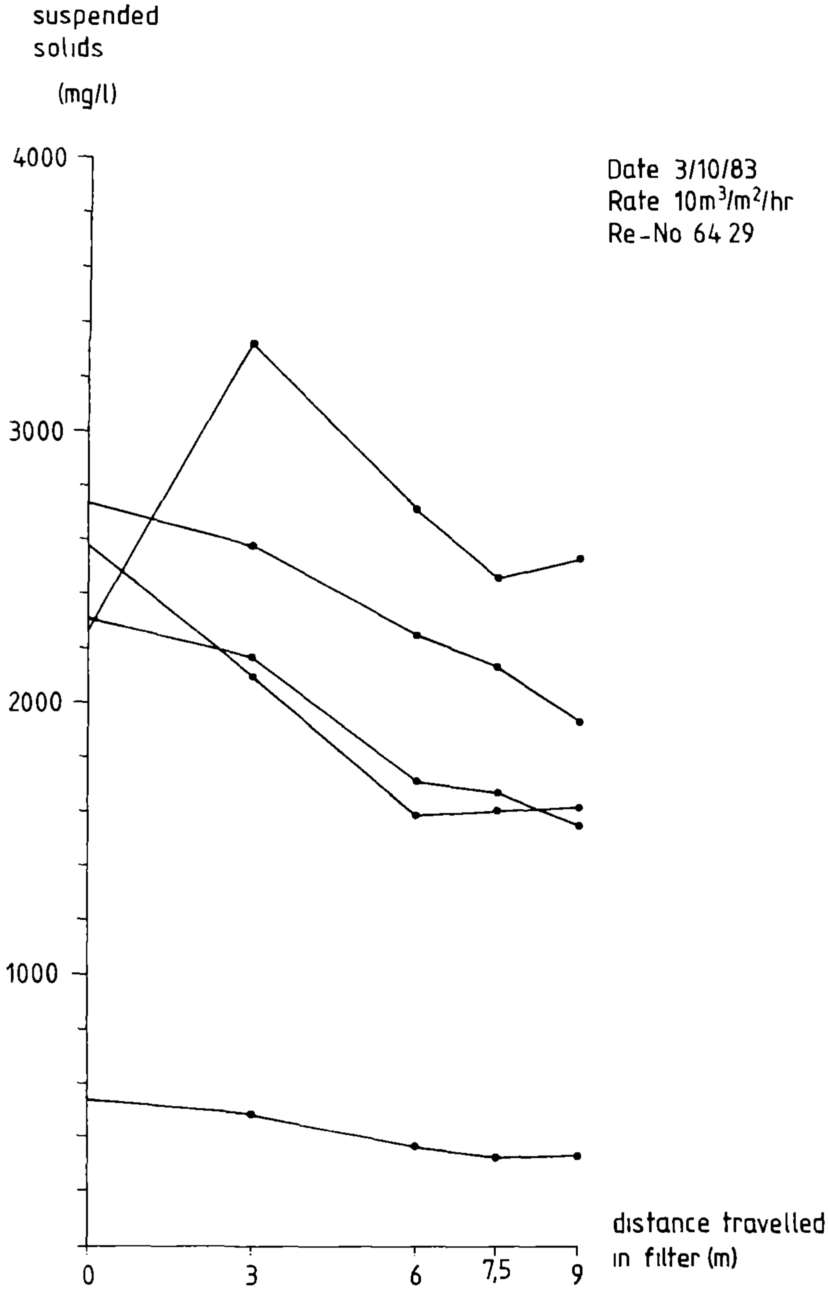


Figure 42. Pilot filter test. Suspended solids removal

11.4 Evaluation of results

11.41 Effect of media selection and Reynold's number

In Oulu the investigations on the HRF were carried out with four media graded from coarse to fine in the order 18 - 32 mm, 8 - 18 mm, 4 - 8 mm and 1,5 - 5 mm respectively. Working with this media Hanhimäki (1983) found that the finest gravel (1,5 - 5 mm) produced faster clogging without any more significantly notable reductions than the preceding 4 - 8 mm gravel. Thus the investigator studied the filter performance with the remaining triple media. The total reduction percentages obtained by this triple media are appreciably lower than the one obtained from the dual media investigated in connection with the subject study.

To find the reasons for the enhanced removal efficiency the Reynold's numbers were computed according to Airaksinen (1978) only for results pertaining to the hydraulic loading rates that gave good total reduction percentages.

$$Re = \rho \frac{VD50}{\nu}$$

where
- V = approach velocity Q/A in m/sec
- $D50$ = diameter of 50 % passing gravel
- ν = kinematic viscosity of the water
- ρ = density of water

The reference computations and the corresponding Re-numbers have been shown on table 4. The area for computing the approach velocity is taken from piezometer readings in figure 43.

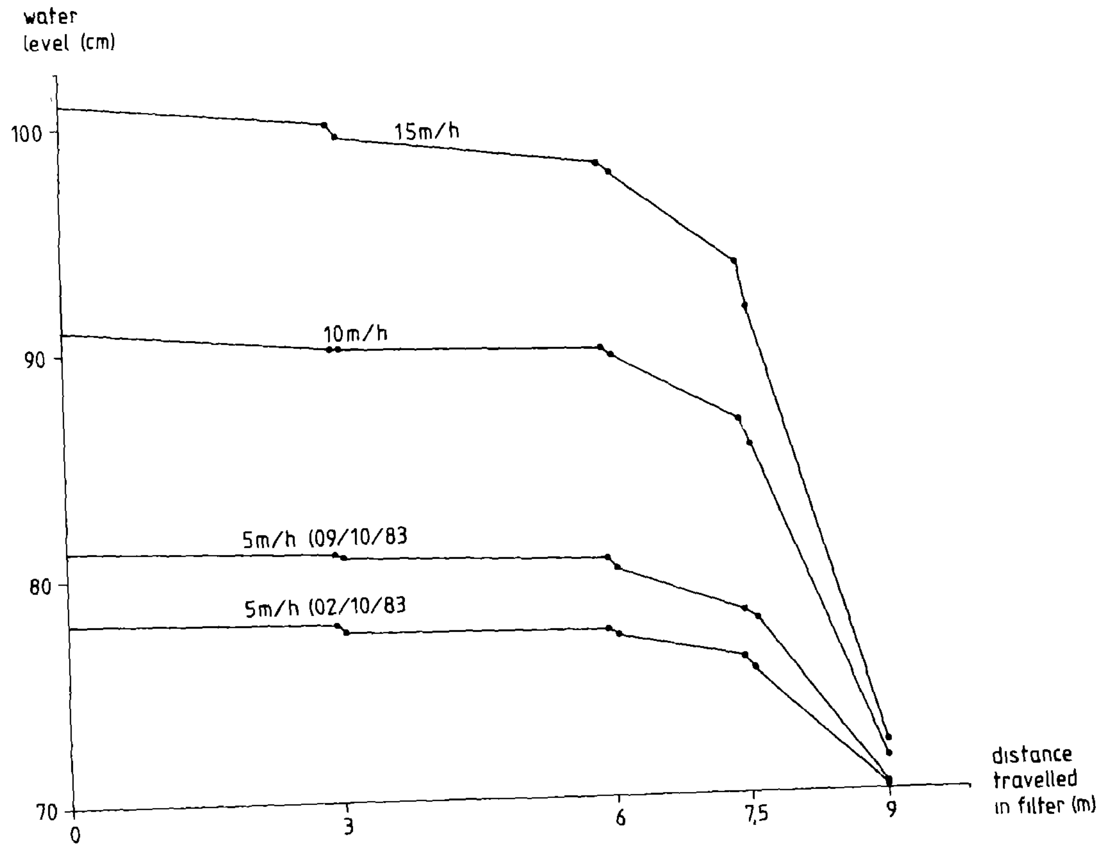


Figure 43. Piezometric readings for the various flow rates

Table 4. Reference computations for the Re-numbers

Measured porosity of 18 - 32 mm gravel = 0,38

Test number	Nominal loading rate (Q) ($m^3/m^2/h$)	Area loaded (A) (m^2)	Approach velocity = Q/A (m/sec)	Re-number = $\frac{VD_{50}}{\nu}$	Remarks
1	5 (2.10.83)	0,78	0,0018	38,00	D_{50} is 30 mm in all cases
2	5 (Hanhimäki)	0,66	0,0020	45,00	= $1,4 \times 10^{-6}$ m^2/sec
3	5 (9.10.83)	0,81	0,0017	36,00	Re-number is calculated for the inlet face of the gravel
4	10 (3.10.83)	0,91	0,0030	64,00	
5	5 (Hanhimäki)	0,85	0,0016	34,00	

Todd (1959) had presented a diagram (figure 44) showing the regions of laminar flow, the transition and turbulent flow in relation to ground water movement. The diagram depicts that turbulent flow commences when the Reynold's number calculated as before is slightly over 10. The best performance results of the HRF were obtained in all cases at Re-numbers of over 10 and less than 100 (pls. refer to table 4). The performance of the filter for total suspended solids reduction as well as for proportionately more removals in the coarse (18 - 32 mm) media than in the fine media was obtained at Reynold's number of 38,00 in this investigation. Thus effective filtration has been evidenced at filtration velocities corresponding to Reynold's numbers at about the onset of turbulence as noted also by other researchers. Kuntschik (1976), Wegelin (1981) and Riti (1981) observed that there is an increase in removal rate in the range of velocities corresponding to the onset of turbulence, which may multiply the chances of contact between grain surfaces and the suspended particles. This is in keeping with the theory of filtration that enhanced transport mechanisms promote removal of impurities.

The significance of the Reynold's number yielding best performance cannot be overemphasized. It indicates that good performance can be expected even with higher loading rates so long as the Reynold's number is maintained at about 38. This means that the cross-section of the filter can also be designed accordingly.

For about the same Reynold's numbers, however, the performance of the dual media was found to be slightly better than that of the triple media. This could only be attributed to the grading. As has been pointed out by O'Melia and Ali (1978) and Degremont (1979) and confirmed by this investigation a monograded media could be expected to perform as well as media graded from coarse to fine in the direction of flow.

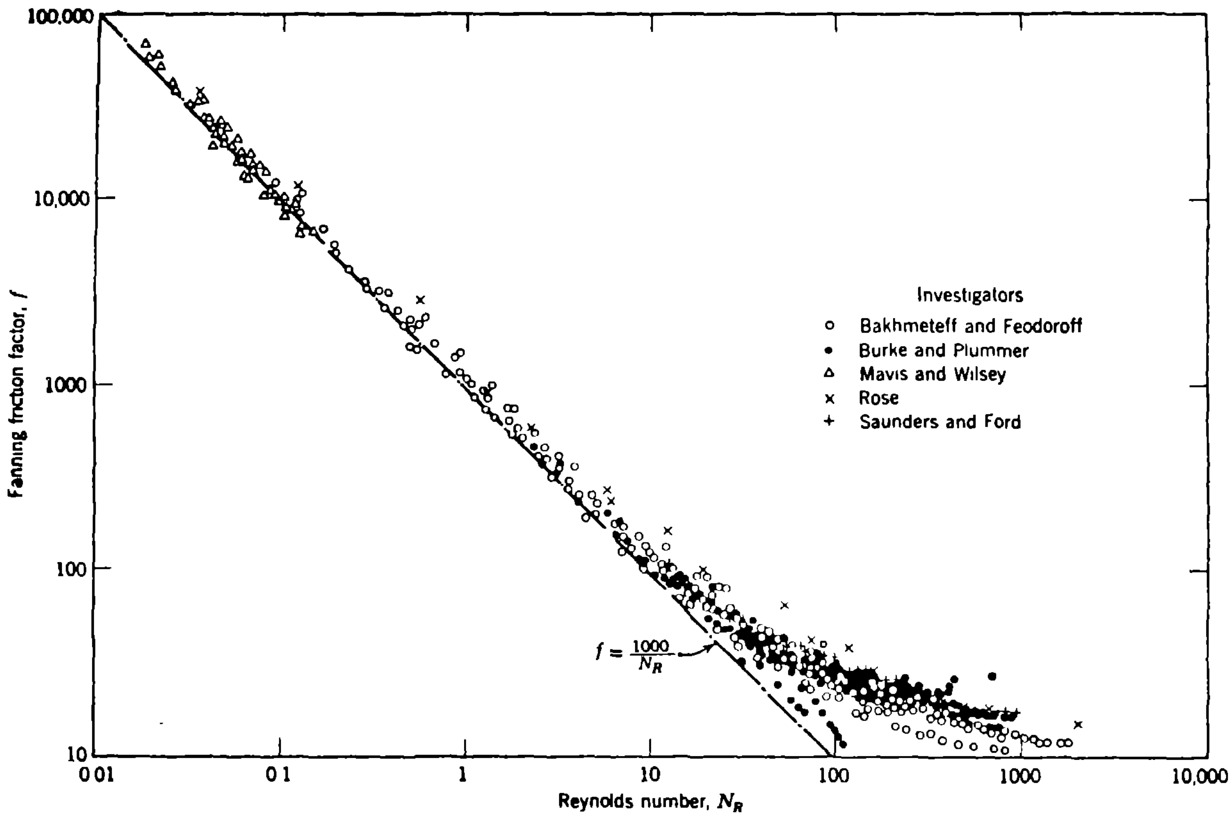


Figure 44. Relation of Fanning friction factor to Reynold's number for flow through granular porous media. Deviation from laminar flow occurs at Reynold's number = 10 (Todd, 1959)

11.42 Effect of variation of suspended solids load

From table 3 and figures 40, 41 and 42 it is apparent that the effluent quality is dependant on the initial load in terms of suspended solids. This is contrary to what has been concluded by Riti (1981). The explanation for this is that Riti (1979 - 81) worked with low turbidity water and as a result the effluent qualities seemingly converge for low values. Here, however, the dependance of the effluent quality on the initial load was marked. Studies made by Hanhimäki (1983) and Knutschik (1976) confirm the dependence of the effluent quality on the initial load. The higher the suspended solids load the higher is the suspended solids content in the effluent.

One other trend observed with the increase in suspended solids content is that the filtration efficiency increases to a maximum and then decreases. The optimum suspended solids content with the hydraulic loading rate of 5 m/h is about 700 mg/l. Two such typical trends have been plotted in figure 45.

On the other hand, the removal of the filter did not follow the usual continuously decreasing suspended solids pattern for low turbidity waters. Figure 46 shows the plotting of typical results. The erratic removal behaviour of the filter for the suspended solids content of 100 mg/l is remarkable. This pattern with low turbidity waters had been observed by Hanhimäki (1983) also. (In this context low turbidity refers to suspended solids content of about 150 mg/l or less.)

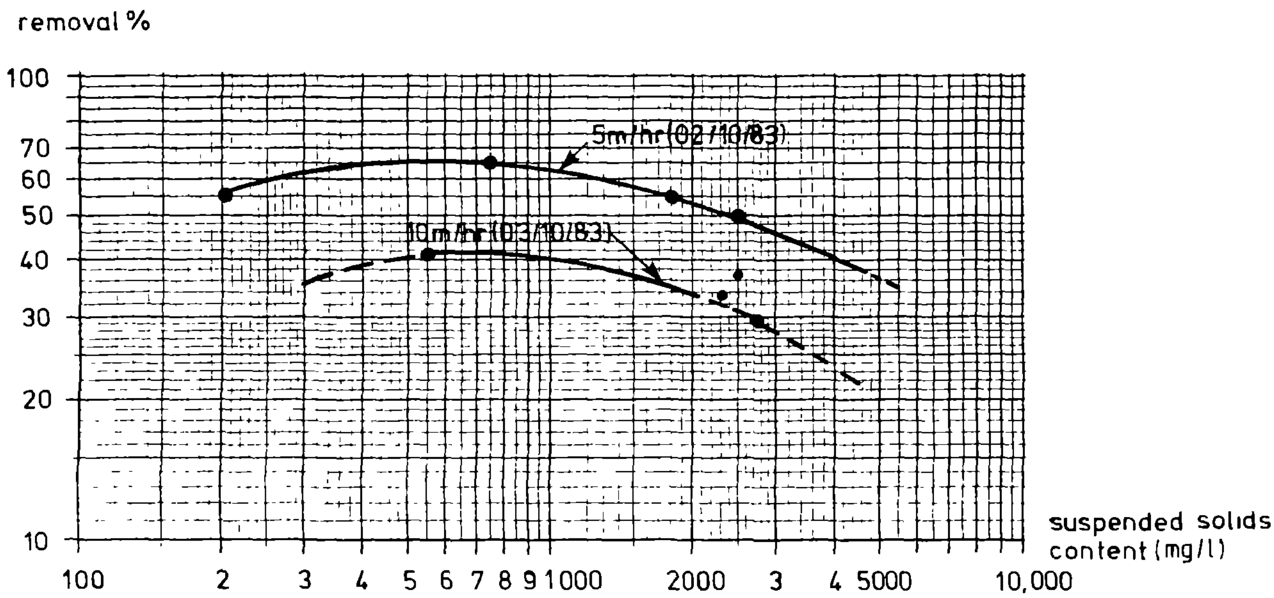


Figure 45. Efficiency of filter in removing increasing content of suspended solids

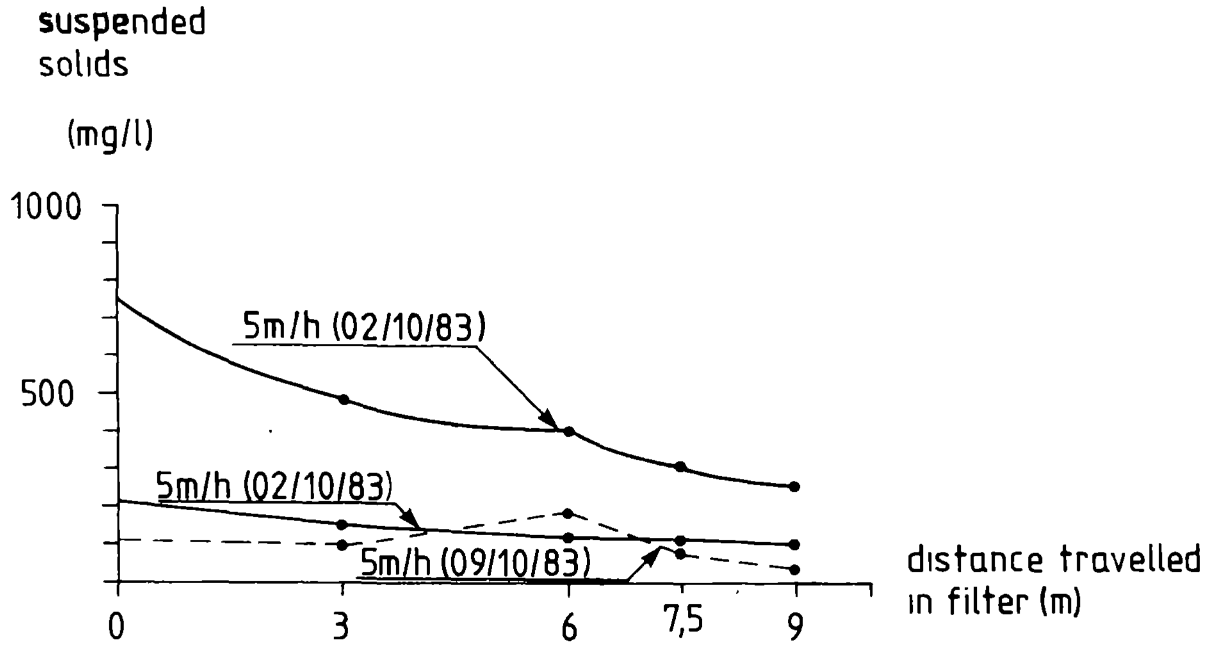


Figure 46. Plotting of typical results

11.43 Effect of variation of hydraulic loading rates

The results of the tests in table 3 show that the performance of the filter was poor under changing hydraulic loading rates. Good removals were obtained only when the filter was run at stable filtration rates. The hydraulic loading rate of 15 m/h gave consistently erratic and poor results under all circumstances. The results also indicate slightly better performance when the filtration rates were increased from 5 m/h to 15 m/h than the opposite operation. Hence, based on the above observations it seems appropriate to draw the conclusion that horizontal filters should not be subjected to erratic hydraulic loading rate changes if they have to perform satisfactorily.

11.44 Prediction of length of the HRF

Iwaski (1937) postulated that the quantity of suspension particles removed by a layer of filter media is proportional to the concentration of suspension entering that layer. The mathematical representation of this theory is the following:

$$-\frac{\partial c}{\partial L} = \lambda c \qquad (2)$$

in which c = volumetric concentration of material entering a unit volume of filter

L = filter depth

λ = filter coefficient

A rational basis for the above assumption has been provided by investigators in aerosol filtration together with supporting experimental evidence (O'Melia and Stumm, 1967).

According to equation (2) the rate at which the suspension concentration diminishes with respect to distance is proportional to the local concentration in the filter. In a uniform filter the reduction in concentration will be logarithmic with filter depth. A consequence of this fact is that in a uniform filter, layers of the media farther from the surface remove progressively less suspended particles (Hedberg, 1976; Ives, 1982).

It has been noted in this experiment that while working at good removal efficiencies, the removal of the 9 m HRF progressively decreases until it reaches the effluent quality value. From figures 40 and 46 it is evident that except for the low turbidity water, the removal pattern traces a logarithmic curve.

If this removal curve were plotted on a semi-logarithmic paper, it should plot as a straight line. Indeed, figure 47 shows this to be the case except for that of the low turbidity water again. This is in conformity with the theory discussed under the subject topic.

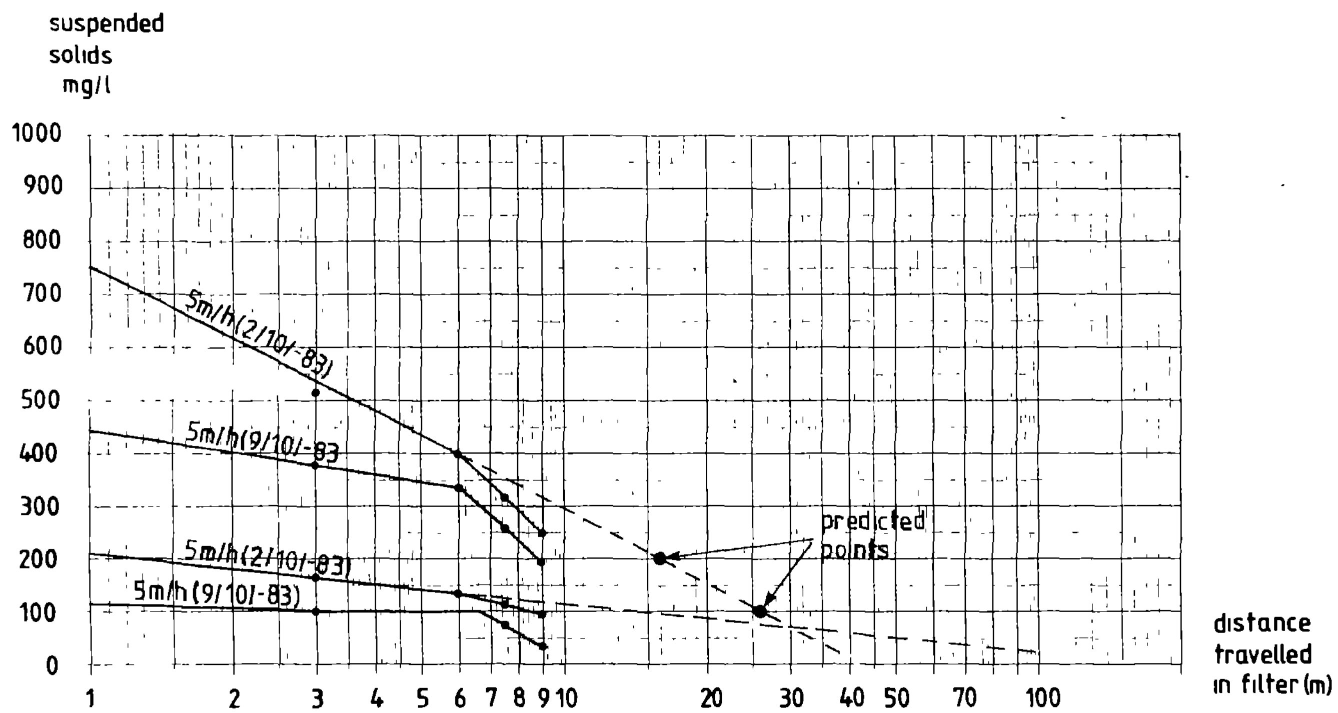


Figure 47. Prediction of filter length

Thus extrapolation of the straight line should theoretically give the length of filter for a required effluent quality from an influent of known concentration. For direct filtration practices it has already been stated that 200 mg/l has satisfactorily been treated in Guyana (Voss and Gross, 1981). Thus assuming that the maximum amount of suspended solids content to be encountered with river waters in tropical countries is 750 mg/l, extrapolation of the line for the data of 5 m/h (2.10.83) gives a minimum filter length of 16 m for a reduction of suspended solids content to 200 mg/l. This relates to removal in coarse gravel alone. The fine gravel can be expected to give at least 10 % polishing removal which would bring the total suspended solids content well below 200 mg/l. However, to ensure longer filter runs the length of the filter could be extended beyond the minimum 19 m (16 m for the coarse gravel and 3 m for the fine gravel) as required. For instance a 26 m filter (coarse gravel alone) could be expected to produce an effluent much less than 100 mg/l suspended solids content consistently from an influent containing even up to 750 mg/l.

It is important that the prediction is made on the basis of the removal curve for the coarse gravel. This is in order to ensure that the clogging potential (storage capacity for suspended solids) of the coarse gravel is utilized to the maximum before the run is terminated due to the premature clogging of the fine gravel. Besides this, the pattern of clogging in the coarse gravel probably follows the clogging front advancement noted with downflow filters. Since most of the removal takes place in the first few meters of the HRF this part would evidently clog faster and more removal would continue in the next layer and so on until final breakthrough. Therefore the longer the coarser media the longer the filter run to be expected. Thus it should be noted that the determination of the actual length of the HRF would also depend on the length of filter run required. In Germany for an average influent turbidity of 8 mg/l (Wegelin, 1981) a length as long as 50 m had been provided and operates five years between washings (Kuntschik, 1976). In Switzerland for a mean suspended solids concentration of about 7 mg/l 15 m long HRFs operated for four years without the need for cleaning them (Wegelin, 1981). Filtration rates of 10 m/h in the case of the former and 5 m/h in the case of the latter were used (Wegelin, 1981).

Further, since the removal curve would depend very much on the raw water conditions, the media selected, the filtration rate etc. it is necessary to get adequate data from pilot plant tests pertaining to the actual circumstances so as to make a similar but reliable prediction for the minimum length of the coarse gravel in the HRF. That filters treating low turbidity waters should be long, while those treating waters with high suspended solids concentrations could be made shorter has been discussed in relation to filtration variables and removal efficiency. Hence, predictions should be made based on local conditions and requirements.

In conclusion of the subject topic it is recommended that some sort of simple presedimentation unit be included ahead of the HRF if the suspended solids content of the raw water source exceeds about 600 mg/l which is the optimum level of efficiency for the HRF operation. It is remarkable that this value does not deviate considerably from the limit of 500 mg/l for the efficient and economic performance of deep bed filters pointed out by Tien et al (1979).

12. DIRECT FILTRATION TESTS

12.1 Purposes of the test runs

Test runs were carried out on down-flow rapid experimental filters (figure 48) using made up water. The objectives of the experiments were

- to find out if direct filtration of the effluent from the HRF can provide a final effluent quality in terms of turbidity for a reasonable filter run period,

- to assess the effect of loading rate, variation of turbidity and aluminium sulphate dosage on filter performance.

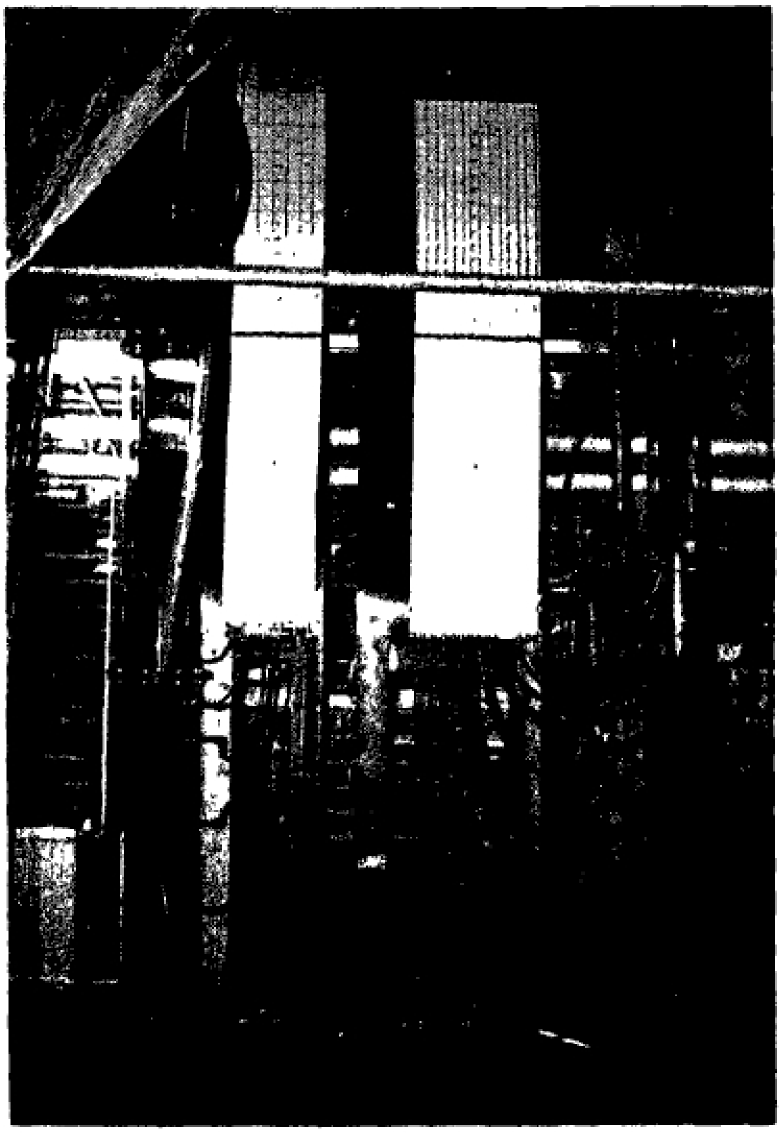


Figure 48. The photograph of the down-flow experimental filters

12.2 Experimental setup and testing procedures

Figure 49 is a self explanatory schematic diagram of the experimental setup. It consisted of a suspension preparation system, a feeding system, a dosing system as well as pressure development and effluent quality monitoring system. Figures 50 a, b and c show the mixing tank, the dosing equipment and the monitoring devices.

The mixing tank where the suspension is prepared is the same one used before the HRF.

The made up water prepared in the mixing tank is led to the constant head feed pipe. This pipe as its name implies keeps the level of water at the same level so that a constant rate of flow is maintained during the filtration process. Calibration of the rate is achieved by raising or lowering the annular v-notch weirs. The excess water continuously overflows to the drain.

The dosage system consists of a tank of known volume containing the chemical solution. The chemical is kept in solution by a mixer. Volumetric dosage pumps are used for dosing the solution. The rapid mixing of the alum with the suspension is achieved by the hydraulic energy of contact between the jet from the dosing tubes and their passing through an elbow bend which had been tapered so as to provide a plug-flow action. There is no mechanical rapid mixing. The contact time of the coagulant and suspension before they reach the bed is only few minutes.

1 Mixing tank first compartment
2 " " second "
3 " " third "
4 " " fourth "
5 Pipe conveying regulated flow
6 Pipe " " most of the flow
7 Valve for regulating the flow
8 Baffles for head control
9 Weirs for rate control
10 Filters
11 Piezometer board
12 Sampling points
13 Piezometer connections
13 Piezometer connections
14 Pressure transducers
15 On-line turbidimeters
16 Automatic plotter
17 Chemical solution tank
18 Dosing pumps
19 Stirrer

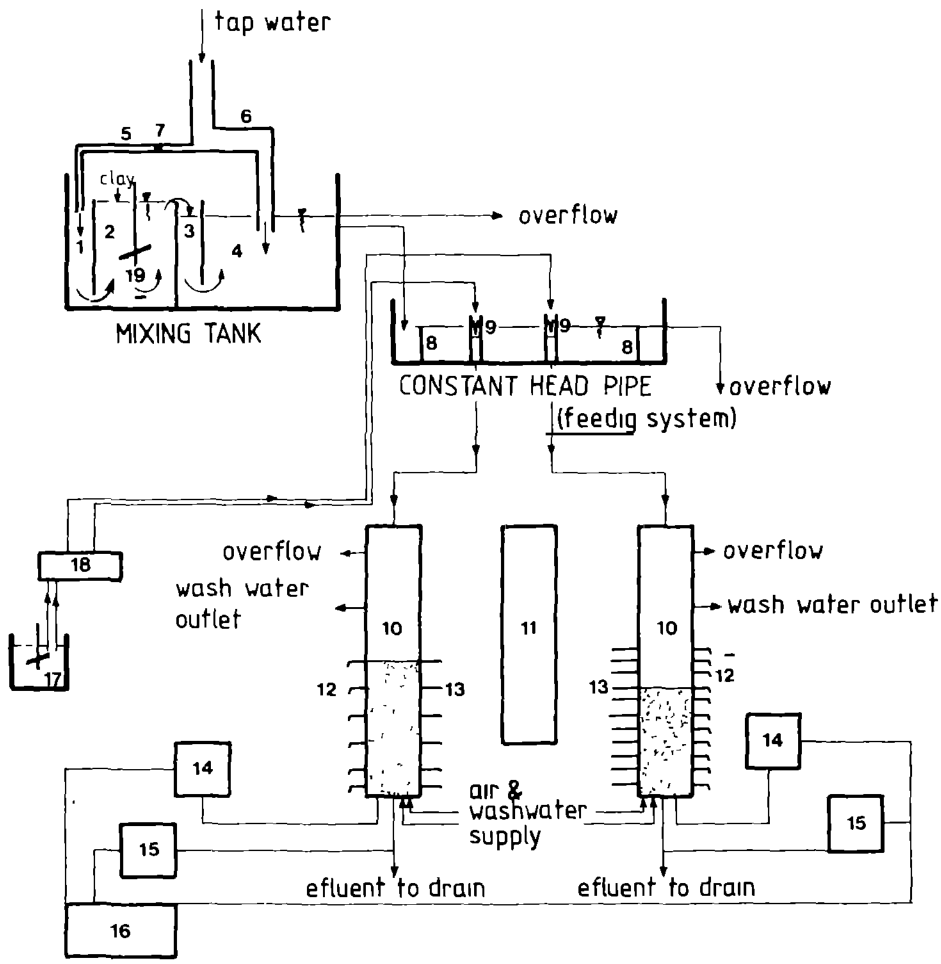


Figure 49. Schematic diagram of the experimental setup for direct filtration.

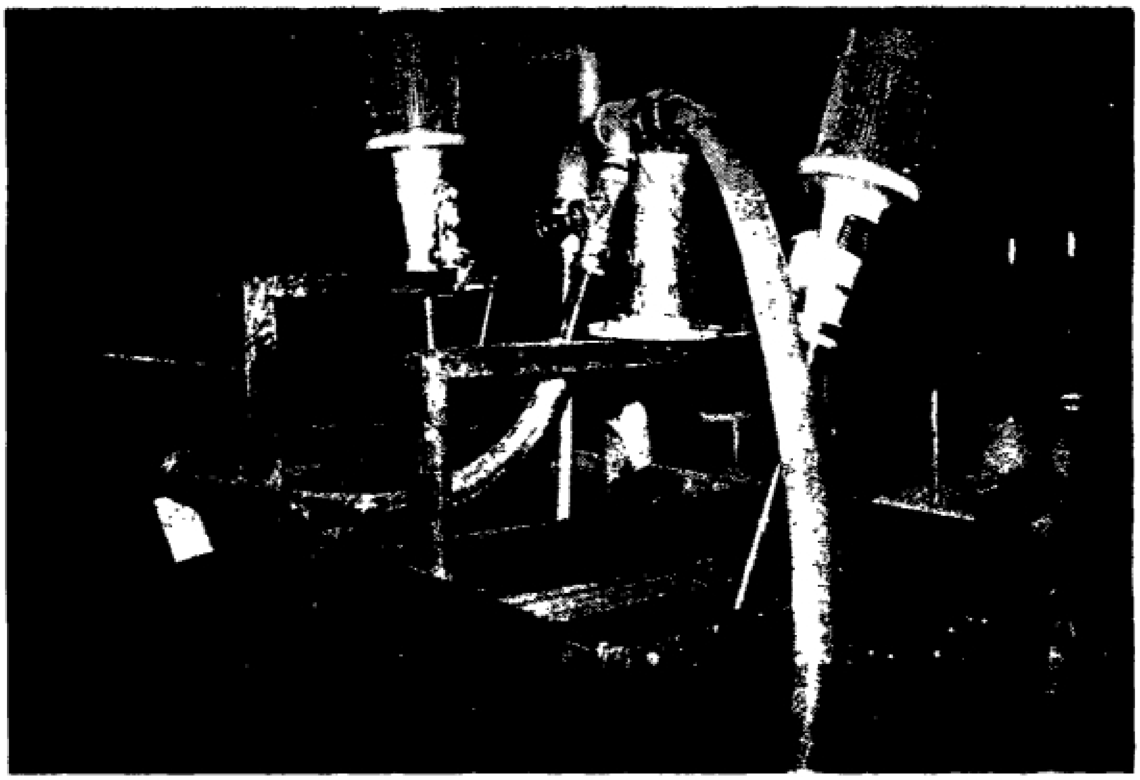


Figure 50 a) Mixing tank for preparation of the made up water

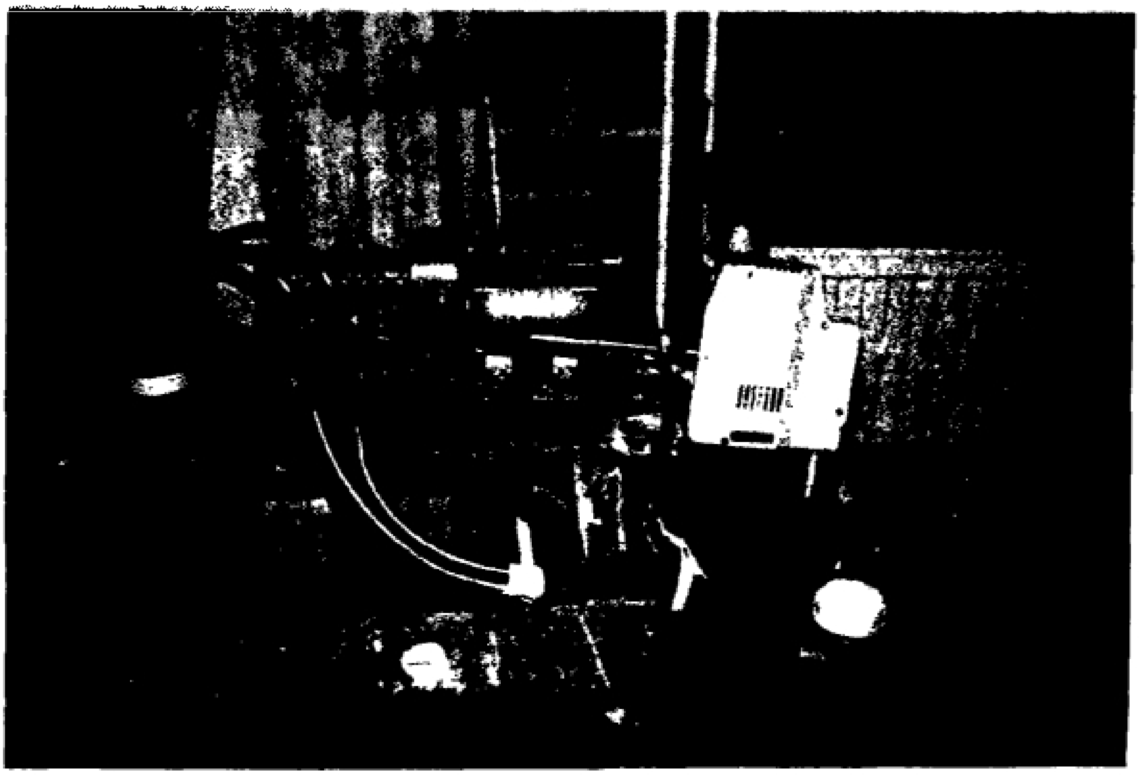


Figure 50 b) Chemical dosing equipment

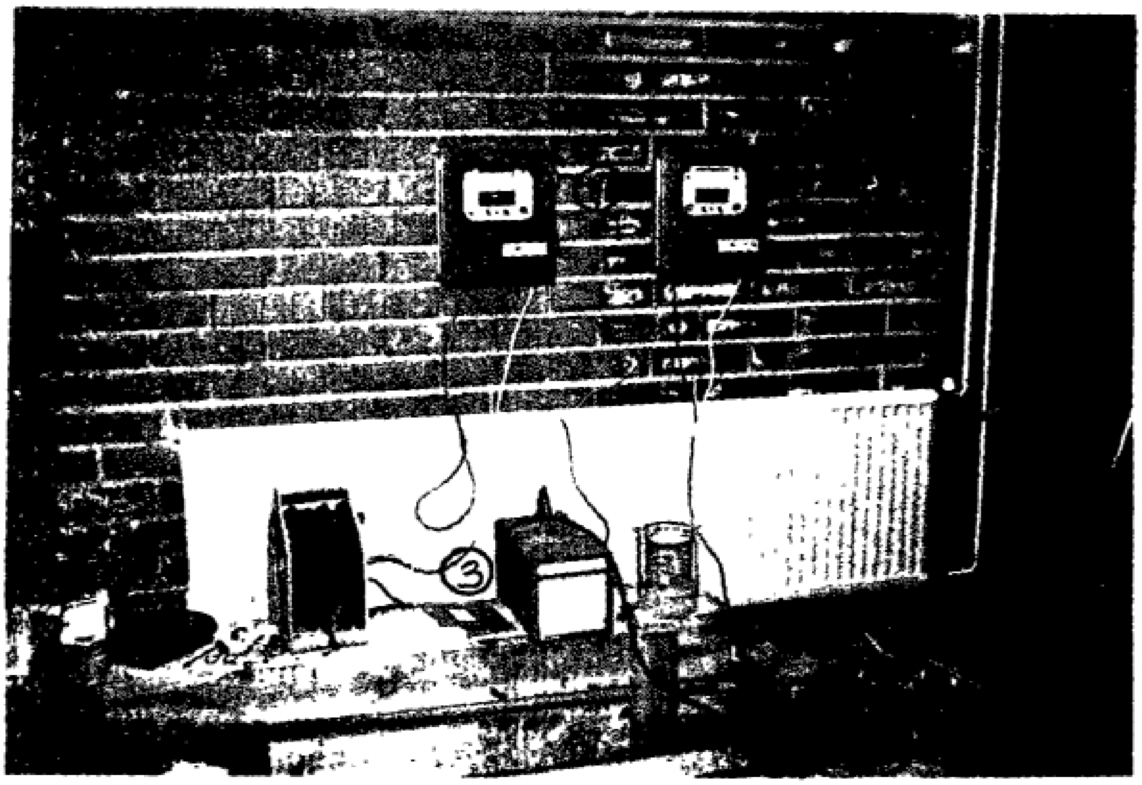


Figure 50 c) Pressure development and effluent quality monitoring system consisting of

1. on-line turbidimeters (on the wall)
2. pressure transducers (on the floor)
3. automatic plotter (on the wooden platform)

The test filters are two pyrex columns about 3 m high and 15 cm in diameter. The bed depths were at the start of the experiment 1,40 m in one of the filters and 1,0 in the other. Connections for piezometers and for sampling are located at short intervals along each column. The supernatant water level is constantly kept above the bed by raising the effluent discharge pipe to the required level. The filters have been connected to air and clean backwash water supply lines. The filter bottoms have also been connected to pressure transducers. The effluent pipe of the filters has a branch which connects it to the on-line turbidimeters. Both the pressure transducers and the on-line turbidimeters are connected to an automatic plotter.

The filter media was sand of grain size 0,8 to 1,2 mm and uniformity coefficient (d 60/d 10) of about 1,25. Figure 51 shows the grading curve for the sand. This type of sand was chosen so as to avoid rapid surface clogging as well as to ensure ample storage. Surface clogging is avoided because with this type of sand the media remains homogeneous after backwashing and fine to coarse stratification is avoided. The filter media was supported by 10 cm thick coarse sand of 3 mm diameter grain size. Below this coarse sand is a 0,35 mm strainer which collects the filtrate. Before every filter-run the media is washed with water aided by air scour and rinsed at slightly higher backwash rates with water alone.

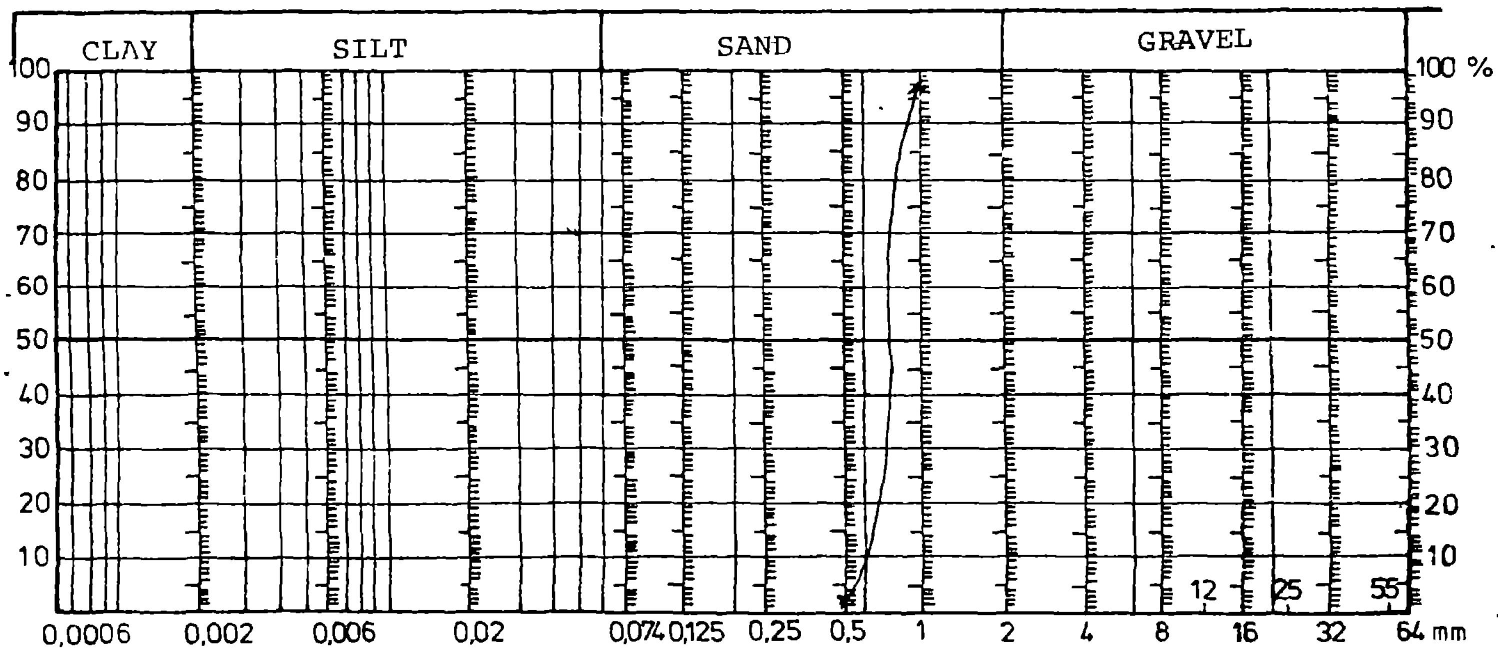


Figure 51. Grading curve for the filter sand

The effluent quality from the filters was monitored by the on-line Great-Lakes turbidimeters and a separate Hach turbidimeter. The Hach turbidimeter was also used to monitor the influent turbidity. Pressure developments were read from the piezometers in addition to the plottings from the automatic plotter. It is appropriate to point out here that the on-line turbidimeters were initially giving unreliable and unreasonable readings. This may have been due to the fact that the meters have not been used for a long time. Nonetheless the plottings gave a good indication of the performance of the filter especially for the periods when it run unattended.

The tap water used for making the made up water had more or less similar properties as the raw water used for the roughing filter in Oulu. A summary of the major characteristics is presented in table 5 below. The values are averages for the testing period.

Table 5. Properties of the tap waters

Characteristics	Tap water from Rusko	Tap water from Oulu
Temperature	1,6 °C	4,4 °C
Colour	5 pt mg/l	8 pt mg/l
KMNO4	9,8 mg/l	8,5 mg/l
pH	8,8	8,7
Residual Alum	0,11 mg/l	0,24 mg/l
Fe	< 0,05 mg/l	0,17 mg/l
Mn	< 0,02 mg/l	< 0,005 mg/l
Residual Chlorine	0,28 mg/l	0,035 mg/l
Residual Turbidity	0,08 FTU	0,67 FTU
Hardness	3 °dH	2 °dH
Alkalinity	0,64 Mval/l	0,2 Mval/l

Because of the close similarity in properties it was not necessary to modify the characteristics of the tap water at Rusko in any way. Remarkable is the low temperature at which the experiments at Oulu and Rusko were carried out.

The concentration of suspended solids of the made up water for the direct filtration tests were judged by its degree of turbidity. The reasons for the decision to use turbidity as a measure of evaluation of the treatment efficiency were the following. For one thing, the criteria adopted for the break-through condition is in terms of turbidity units. For another in view of the number of the samples required for the tests, the determination of suspended solids content would be too involved, time consuming and costly. However, from the results of the experiments on the roughing filter (table 3) it is evident that there is a close relationship between the turbidity values and the suspended solids content for low concentrations (i.e. below about 200 mg/l). For instance, the results of 2.10.83 show a more or less direct relationship. Thus, eventhough suspended solids concentrations for the direct filtration tests were not made, it is thus still possible to get some idea of the suspended solids content of the influent to the downflow filters from the turbidity values measured.

The coagulant chosen for the experiment was aluminium sulphate $[Al_2(SO_4)_3\ 14H_2O]$ without any coagulant aid. Although initial screening of possible dosages was done using the jar test, the actual requirements were determined by optimizing it on the filters. Dosages that yielded acceptable effluent values within a reasonably short break-in time (less than half an hour in most cases) were maintained up to the end of the respective filter runs.

It has become a common practice in the operation of granular deep-bed downflow filters to terminate the filter run according to two criteria:

a) The effluent quality criterion, usually expressed in terms of maximum permissible filter effluent turbidity or suspended solids concentration.

b) The head loss criterion, which is the maximum head loss allowed to develop.

The operation is optimized if the breakthrough and head loss limits are reached simultaneously. (Adin and Rebhun, 1974; Bratby, 1980; Ives, 1982) For this investigation the limit for the acceptable effluent quality value for turbidity is taken as 5 NTU and for colour 20 PCU. These values are the WHO (World Health Organization) standards that are widely accepted according to Wagner and Hudson (1982). Concerning the allowable head loss, Bratby (1980) reports that many water treatment plants employing direct filtration permit a maximum head loss of 1 - 2 m. The maximum available head loss provided in connection with this experiment of 1,0 m is thus in keeping with the above.

12.3 Evaluation of the results

The results of the tests have been presented in table 6 and figures 52 to 58. The plottings in the figures are to be read and interpreted in conjunction with table 5. This is because there is marked difference between the readings of the Hach 2100 and the on-line Great Lakes turbidimeters. For instance although the on-line meter shows that the effluent quality is consistently above 5 NTU, the Hach 2100 readings indicate that the effluent turbidity is on the average $\leq$ 5 NTU for the attended 10 hrs for Run No. 4 at a dosage of 2,5 mg/l. The reliability of the on-line meters is questionable because they gave consistently higher readings most of the time. On the other hand the Hach 2100 meter gave reliable readings when cross-checked against another meter of the same manufacturer whereas the on-line meters did not. It is suspected that the malfunction of the on-line meters may be due to their being out of use for an extended period. However, for all practical purposes the values give a good indication to the performance of the filter.

Table 6. Observation results from the direct filtration tests

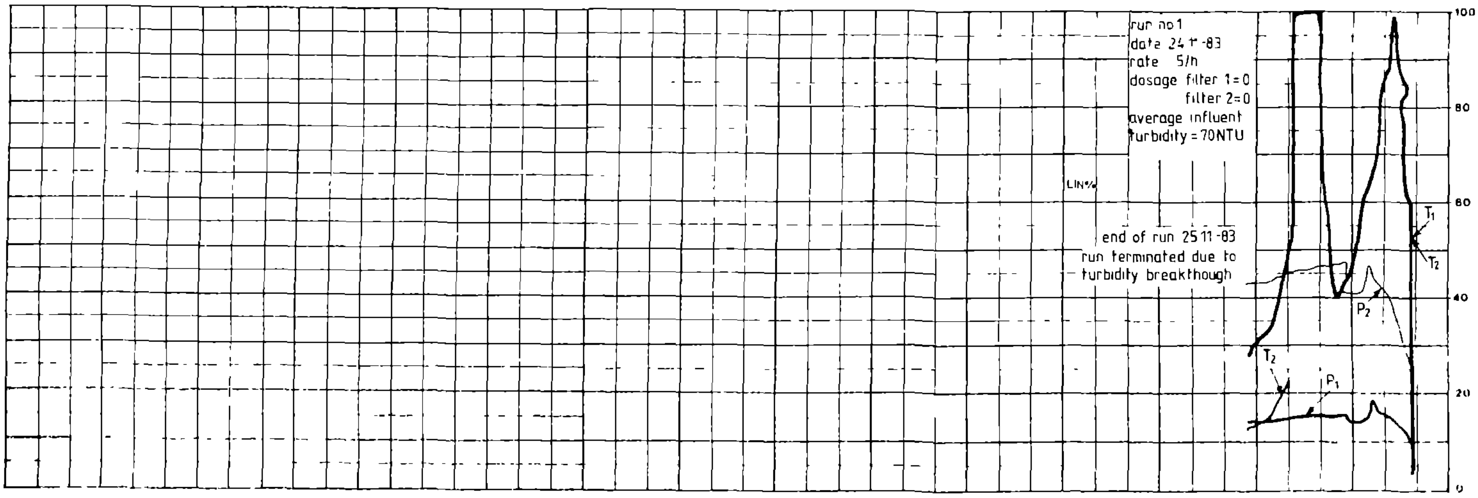
Run No	Date	Time	Rate (m/h)	Dosage (mg/l)		Influent Turbidity (NTU)	Filter 1 Piezometer Readings (cm)											Effluent Turbidity (NTU)		Filter 2 Piezometer Readings (cm)						Effluent Turbidity (NTU)		Remarks
				F1	F2	Hach 2100	1	2	3	4	5	6	7	8	9	10	11	Hach 2100	on-line Gr Lakes	1	2	3	4	5	6	Hach 2100	on-line Gr Lakes	
1	24 11 83	9 00	5	-	-	75	0	0,5	1,5	2,5	4	5	6,5	8	9	10,5	11,5	-	-	0	2	4	8	10	15	-	-	1 1 Run terminated due to break-through of turbidity
		9 30				61	2,5	4	6,5	10	13	16	19	22	25	29	31,5	51	85	2	10	22	35	46	60	53	92	1 2 (F1 stands for filter 1 and F2 for filter 2)
		10 00				62	2,5	4,5	7	10	13	16	19	22	25	28	30,5	34	58	2	11	24	-	49	63	33	62	
		11 00				80	2	3,5	5,5	8,5	11	13,5	16,5	19	21,5	24,5	26,5	33	64	1,8	9	19,75	-	40	52	32	66	
2	24 11 83	14 15	5	-	-	58	-3,5	-3	-3	-2,5	-2	-2,5	-1,5	-1,2	-1	-0,75	0	-	-	-7,5	-6,5	-5	-3,5	-1,5	0	-	-	2 1 Filter run terminated on 25 11 83 due to continued breakthrough of turbidity
		15 00				53	5	7	9,5	12	14,5	17	19,5	22	24,75	27,5	29	34	60	9	16	24,5	33,5	42	51,5	34	80	
		16 00				66	5	7	9	12	14	16,5	19	22	24,5	27,5	28,5	40	65	7	14	22	31,5	40	49	40	70	
	25 11 83	9 30				16	4,5	5,75	7,5	9,5	11,5	13,5	15,5	17,5	20	22,5	23,5	10	40	2,5	7,5	14	20,5	27	34,5	10	40	
3	28 11 83	10 30	5	-	10	30	0	0,5	2	3	4,5	6	7,5	9	10,5	12	13	-	-	0	6	9,5	14	18	23,5	-	-	3 1 Rate of loading suddenly increased uncontrollably thereby reducing the pressures It had no effect on the effluent quality
		11 00				44	1	2,5	5	7	9,5	12	14	15,5	17,5	21	22,5	15	22	1	7,5	16	23,5	31	39	0,5	2	
		11 30				65	1,5	3,5	5,5	8	10,5	13	15,5	18	20,5	22	24,5	34	46	1,5	7,5	17	25,5	33,5	42	0,6	2	
		12 00				75	2	3,5	5,5	8	10,5	13	15,5	18	20,5	23	24,5	35	48	1,5	7,5	17	25,5	33,5	42	0,4	2	
		13 00				180	2	3,5	5,5	8	11	13	16	19,5	20,5	23,5	23,5	62	74	1,5	8,5	17	25,5	33,5	42	0,4	2	
		14 00				120	2	3,75	6	8,5	11	13,5	16	19,75	21,25	24	25,5	59	68	1,5	8,5	17	25,5	33,75	44,5	0,6	1,8	3 2 Run terminated due to break-through of turbidity in filter 2 during the night Filter 1 gave unacceptable effluent quality values up to the end of the run
		15 00				60	2,5	4	6,5	9	12	15	17,5	20	23	26	27,5	55	60	2	4	18	27	35,5	47	0,4	1	
		16 00				63	1,5	3	5	7	9	11	13,5	15,5	17,5	20	21,5	37	48	1,5	7,5	14,5	21,5	29	39	0,45	1	
	29 11 83	9 40																										

Table 6. Cont'd

Run No.	Date	Time	Rate (m/h)	Dosage (mg/l) F1	Dosage (mg/l) F2	Influent Turbidity (NTU) Hach 2100	Filter 1 Piezometer Readings (cm) 1	2	3	4	5	6	7	8	9	10	11	Effluent Turbidity (NTU) Hach 2100	Effluent Turbidity (NTU) on-line Gr Lakes	Filter 2 Piezometer Readings (cm) 1	2	3	4	5	6	Effluent Turbidity (NTU) Hach 2100	Effluent Turbidity (NTU) on-line Gr Lakes	Remarks
4	1 12 83	6 00	5	2,5	2,5	33	1	3	5	6,5	7,5	10	12	14	15,5	17	18	4,7	4	1	4,5	8	12	16,5	22	3,8	4	4 1 Actual breakthrough had occured during the night 4 2 Run terminated on 2 12 83 at 13 00
		9 00		5		70	1,5	3	5	7	9	11	13	15	17	19	20	3	7	1	4,5	8,5	12,5	17,5	23,5	3	4	
		10 00				48	1,5	3	5	7	9,3	11,3	13,3	15,2	17,4	19,75	20,75	3,2	3	1	4,5	8,5	12,75	17,75	23,75	2,7	5	
		11 00				30	1,5	3	5	7	9,3	11,3	13,3	15,5	17,5	20	21,25	0,9	2	1	4,5	8,5	13	18	24,5	1,1	3	
		12 00				75	1,5	3	5	7	9,3	11,3	13,3	15,5	17,5	20,25	21,5	2,5	3	1	4,5	8,5	13	18	24,75	2,6	6	
		13.00				74	1,6	3,2	5,1	7,1	9,3	11,3	13,3	15,5	17,7	20,6	22	2,5	3,5	1	4,6	8,7	13,2	18,2	25	7,3	11	
		14 00				64	1,6	3,3	5,2	7,2	9,3	11,3	13,4	15,6	18	21	22,4	2	3	1	4,7	8,8	13,4	18,5	25,5	5,6	9	
		15 00				58	1,6	3,3	5,2	7,2	9,3	11,3	13,4	15,6	18	21,3	22,7	1,7	2,5	1	4,7	8,8	13,4	18,7	25,8	3,9	7	
		16 00				76	1,7	3,3	5,2	7,2	9,3	11,4	13,5	15,8	18,5	21,8	23,2	2,7	3	1	4,9	9	13,7	19,2	26,7	4,8	8	
		17 00				62	1,7	3,3	5,2	7,2	9,3	11,4	13,6	16,1	18,7	22,1	23,6	2,1	4	1	4,9	9	13,7	19,2	26,7	5,6	10	
		18 00				60	1,8	3,4	5,2	7,2	9,3	11,5	13,7	16,2	19	22,6	24,1	2,1	3	1	4,9	9,1	13,7	19,3	27,1	2,6	6	
	2 12 83	7 00				45	2,5	4,2	6,3	8,7	11,2	13,7	16,5	20	23,2	28	30	5,5	6,5	1,3	5,5	10,5	15,8	22,2	31,5	8	16	
		8 00				75	2,5	4,2	6,4	8,8	11,3	14	10,8	20,2	23,7	28,3	30,5	5,5	8	1,3	5,6	10,6	16	22,5	31,9	15	20	
		9 00				62	2,5	4,2	6,5	8,9	11,4	14,2	17	20,5	24,2	28,7	31	5,6	9	1,3	5,6	10,6	16	22,6	32,2	12	17	
		10 00				56	2,5	4,4	6,6	9	11,5	14,3	17,2	20,6	24,3	29	31,2	5,3	9	1,3	5,7	10,7	16,3	22,9	32,6	12	16	
		11 00				57	2,5	4,5	6,7	9,2	11,7	14,5	17,5	21	24,9	29,5	31,7	5,6	9	1,3	5,8	10,8	16,5	23,2	33	10	16	
		12 00				85	2,5	4,5	6,7	9,2	11,8	14,8	17,7	21,4	25,2	30	32,5	8	12	1,3	5,8	10,8	16,7	23,5	33,5	15	21	
		13 00				71	3	5	7,7	10,5	13,5	16,5	20	24	28,5	33,5	36	18	30	1,5	6,5	12	18	25,5	36,5	23	33	
5	3 12 83	10 00	10	2,5	2,5	90	0	0,5	1,5	2,5	4	5	6,5	8	9	10,5	11,5	5,4	8	0	2	4	8	10	15	4,6	5	5 1 Depth of sand in filter 2 had gone down to 1,3 m due to loss in back-washing whereas depth of filter 1 is still about 1,0 m 5 2 * Dosages increased due to deterioration in effluent quality beyond the acceptable value of 15 5 3 Run in filter run was terminated due to breakdown of of dosing pump whereas breakthrough of turbidity occured in filter 2 during the night NB/effluent quality values of the two filters differ probably due to the difference in depth
		11 00		2,5	5	60	7,75	10,8	14,2	18	21,5	24,5	28	31,5	35	38,5	40	9,5	10	3	10	19	28	35,7	44,5	15	27 *	
		12 00		5		150	6,5	9,5	12,7	16,5	20	23,2	25,8	30	34	37	40	15	17 *	3	10	18,7	27,7	35,5	45	8,5	13	
		13 00				160	6,5	9,5	12,7	16,5	20	23,5	27,2	31	35	40	42,5	9	11	3	10	18,7	27,7	35,5	47	5,7	10	
		14 00				90	6,5	9,5	12,7	16,5	20	23,7	27,5	31,5	36	41,5	44,5	8	10	3	10	18,7	27,7	35,8	48,5	6	11	
		15 00				160	1	9,3	12,5	16,3	20	23,5	27,2	31,4	36	40,5	43	38	38	3	10	18,5	27,7	36	49,5	6,3	12	
		16 00				84														3	10	18,5	27,7	36,5	51	5,8	11	
		17 00				95														3	10	18,5	27,7	37	52,3	7,4	12	
	4 12 83	8 00				45																						

Table 6. Cont'd

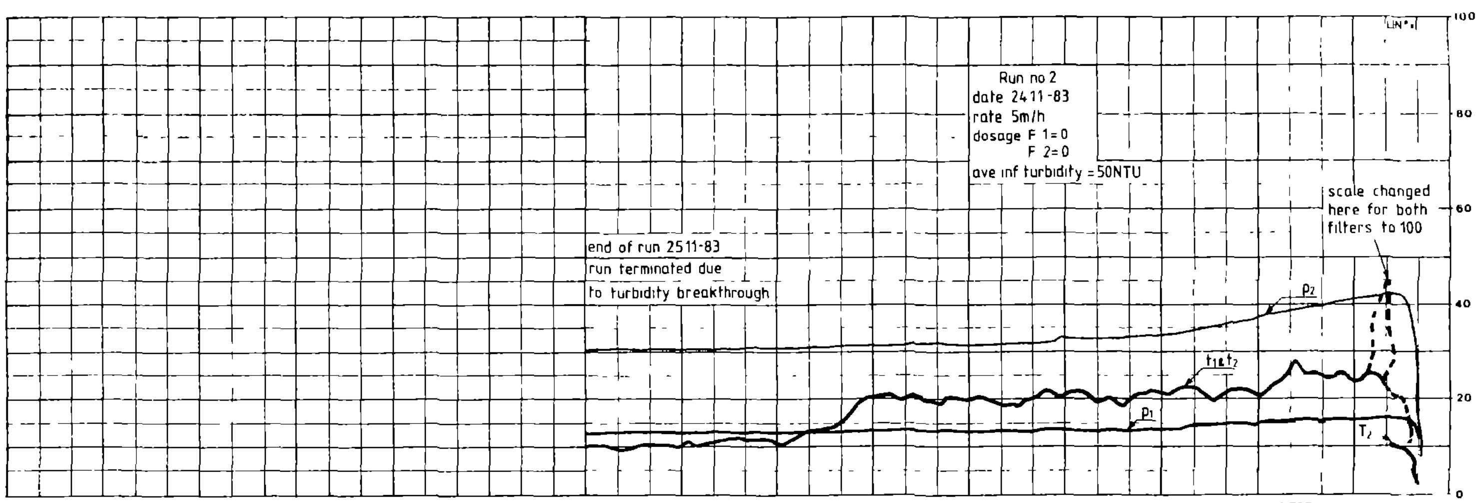
Run No	Date	Time	Rate (m/h)	Dosage (mg/l) F1	Dosage (mg/l) F2	Influent Turbidity (NTU) Hach 2100	Filter 1 Piezometer Readings (cm) 1	2	3	4	5	6	7	8	9	10	11	Effluent Turbidity (NTU) Hach 2100	Effluent Turbidity (NTU) on-line Gr Lakes	Filter 2 Piezometer Readings (cm) 1	2	3	4	5	6	Effluent Turbidity (NTU) Hach 2100	Effluent Turbidity (NTU) on-line Gr Lakes	Remarks
6	5 12 83	10 45	10	5	10	65	0	0,5	1,5	2,5	4	5	6,5	8	9	10,5	11,5	3	3	0	2	4	8	10	15	2	3	6 1 Depth of filter 2 has gone down to 1,2 m
		11 15		15		54	5,8	8,9	12,5	16,2	20,4	24	28,2	32,5	36,8	41,5	43	41*	56*	2,5	10	20,7	31	40,7	50,5	4,5	5	6 2 * Dosage for filter 1 increased to 15
		11 30				44	5,8	8,9	12,5	16,2	20,4	24,2	28,4	32,6	37,2	42,3	44	6	6	2,5	10	20,7	31,2	40,9	51,2	3	4	6 3 ** Dosage for filter 1 raised again to 20 mg/l with no apparent improvement in effluent quality
		14 00				15	5	7,8	10,8	14	17,8	21,3	25,2	30	36	44	47	1	1	2,5	10	20	31,2	41,2	56	1,2	3	
		15 00				92	4,3	6,8	9,5	12,8	16	19,3	23	26,7	38,9	43,5	47	1,5	1	2,5	9,5	19,2	30	40,7	57,5	2	4	
		15 15**		20		89												1,4	1									
		16 00				95	3	5,5	8	10,8	13,7	16,7	20	24,5	30,7	42,5	47,3	1,4	0,8	2,5	9,8	19,5	30,5	42,5	63	1,5	3	
		17 00				69	2,4	4,4	7,5	8,9	11,4	14	17	21	27,2	41	46,9	1,4	0	2,5	9,8	19,5	30,5	42,5	66,5	0,8	3	
		18 00				66	2,4	3	4,9	6,8	8,9	11	13,8	17,4	23,5	39,4	46,5	1	0	2,5	9,8	19,5	30,5	44	70,7	1,4	3	
		18 45				67	1,7	2,6	4,5	6,5	8,5	10,7	13,2	16,8	23,4	41,2	49,5	1,1	0	2,5	9,8	19,7	30,7	44,5	73	1,1	4	
	6 12 83	6 00				42																						
7	6 12 83	7 30	15	10	20	30														0	2	4	8	10	15	6,6	5	7 1 * Run of filter 2 terminated due to fast headloss development
		8 15				39	0	0,5	1,5	2,5	4	5	6,5	8	9	10,5	11,5	6,9	2									7 2 Run of filter 1 terminated after 6,5 hours
		8 30				41														7,7	22	38,5	57	77	97	5,5	1	
		9 00				50	5	6	11,5	15	18,5	22,7	26,5	31	36	45	52	2,5	1	9	22,5	39	58	79	101,5	1,9	1	
		10 00				50	3	5,5	8	11	14	17,4	20,5	24	39	41,8	52,5											
		10 30				55												4,2	1	8,5	21,2	35,7	49,8	76	102,5	5	1	
		11 00				100	2,5	4,5	7	9,5	12	15	18	21	26,5	43	56,5											
		11 30																6	1									
		12 00				130	1,5	3	5	7	9	11,5	14	17	22,5	42	58	2	1									
		13 00				220	0,7	2	3,8	5,5	7,4	9,5	11,8	14,8	20,2	43,5	61	2,7	1									
		14 00				180	0,5	2	3,5	5,4	7,2	9,4	11,9	15	21,7	48,5	62	2,9	1									
		15 00				100	0,5	2	3,5	5,4	7,2	9,4	12	15,4	23,3	52,5	73,2	1,1	1									
		16 00				80	0,5	2	3,5	5,2	7	9,5	12,2	15,7	25,1	55,7	80	3,5										



ADDITIONAL INFORMATION
1 Allowable head loss
= 1,0m for both filters
2 Chemical used is aluminium
sulphate without coagulant
aid
3 P stands for pressure
4 T stands for turbidity
5 1,45 squares = 1hr run
6 Scale

T_1 = 0 - 100% = 0 - 100NTU
T_2 = 0 - 100% = 0 - 100NTU
P_1 = 0 - 75% = 0 - 100cm
P_2 = 0 - 50% = 0 - 100cm

Figure 52. Run 1



ADDITIONAL INFORMATION
1 Allowable head loss
= 1,0m for both filters
2 Chemical used is aluminium
sulphate without coagulant aid
3 P stands for pressure
4 T stands for turbidity
5 1,45 squares - 1hr run
6 Scale

Figure 53. Run 2

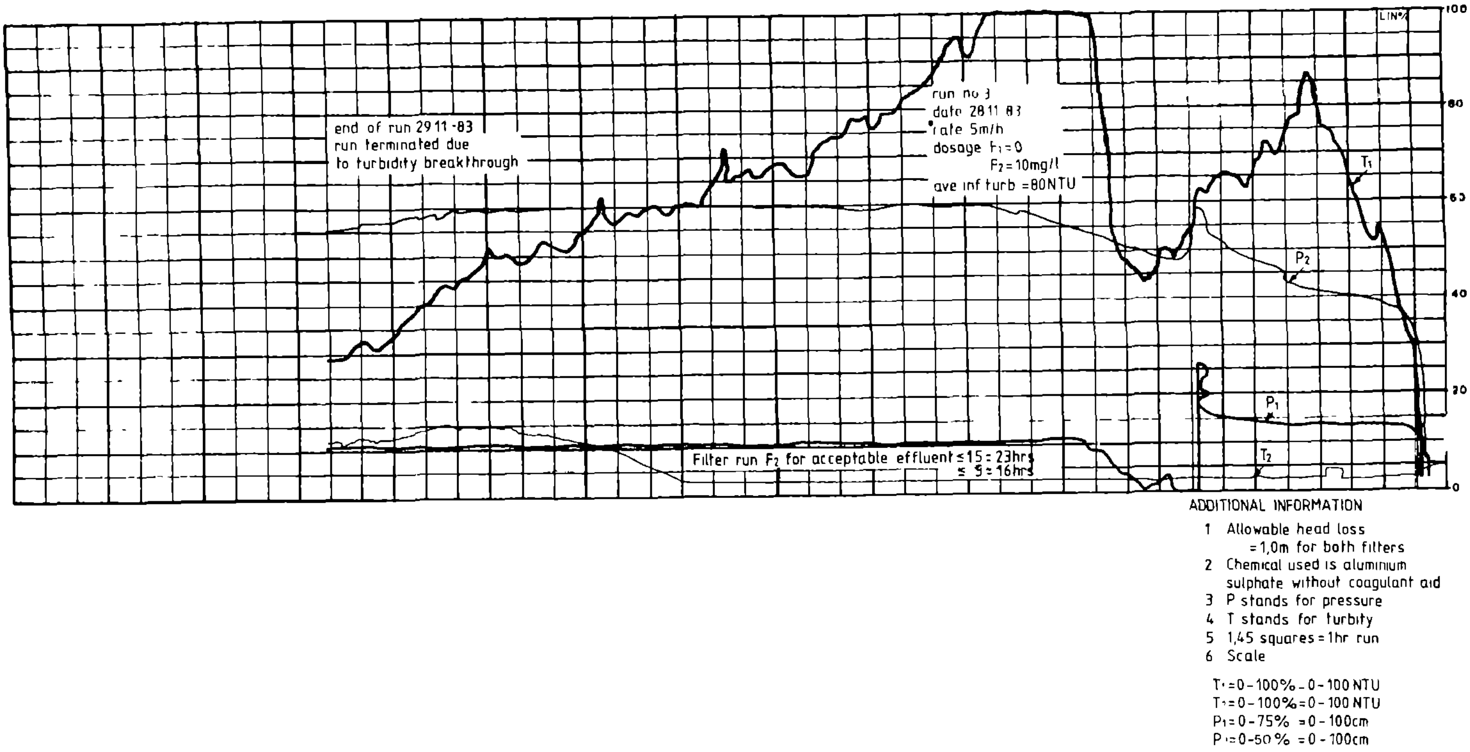


Figure 54. Run 3

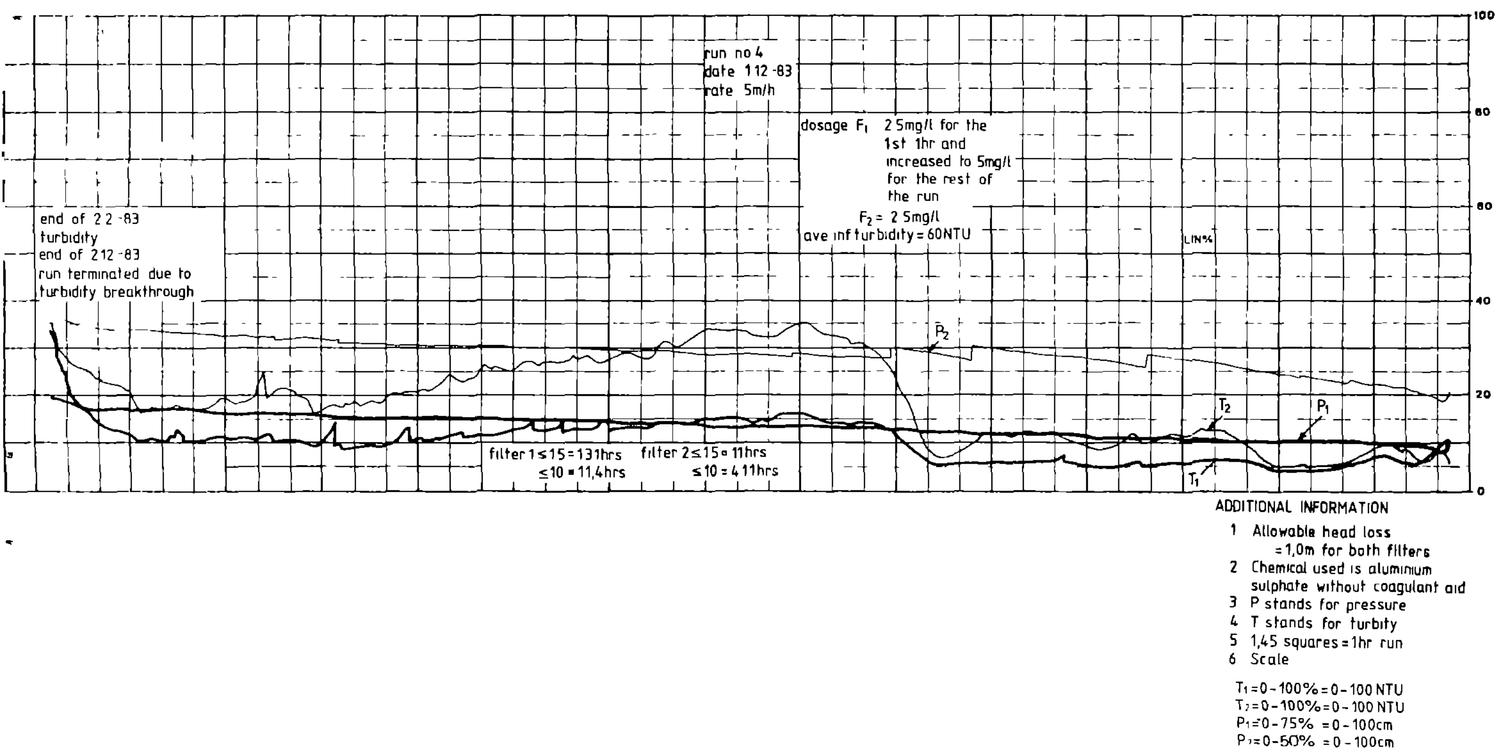


Figure 55. Run 4

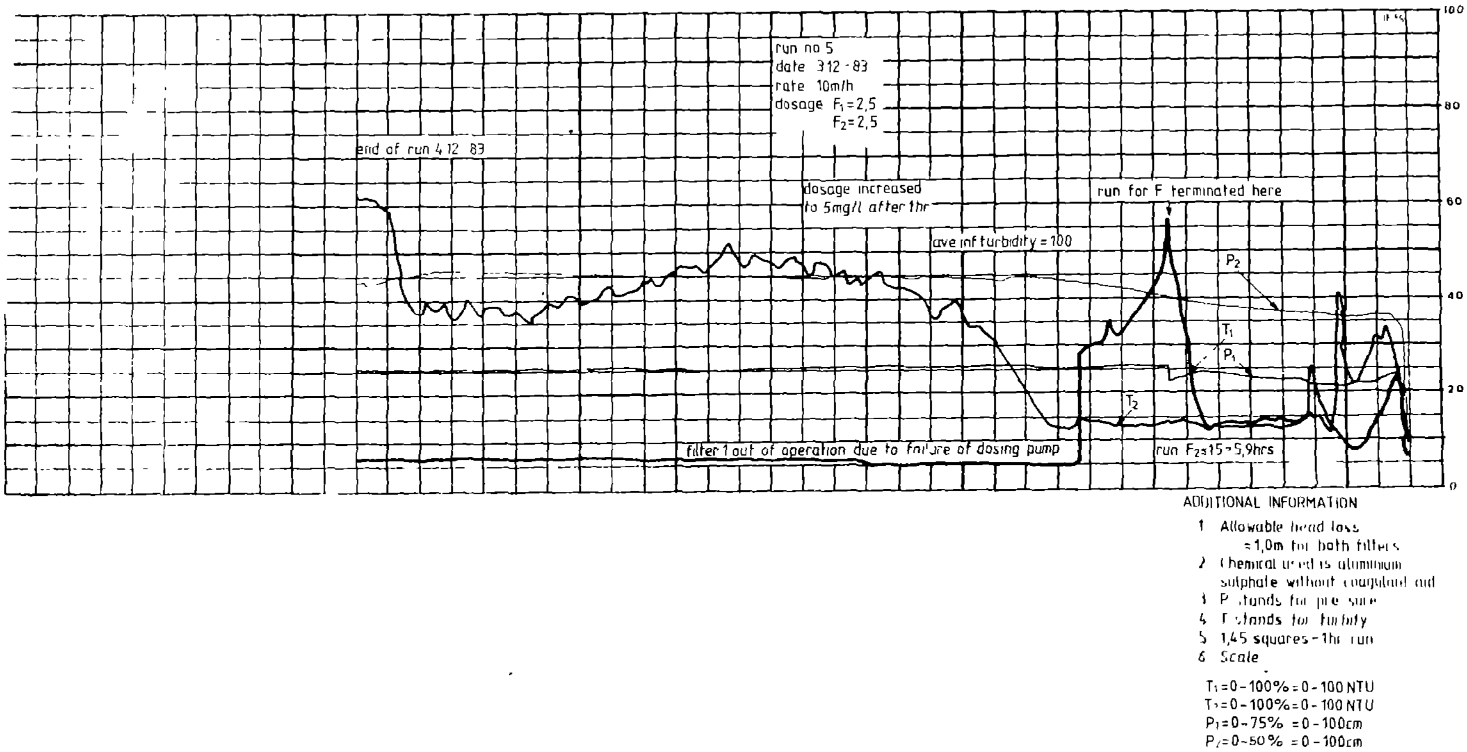


Figure 56. Run 5

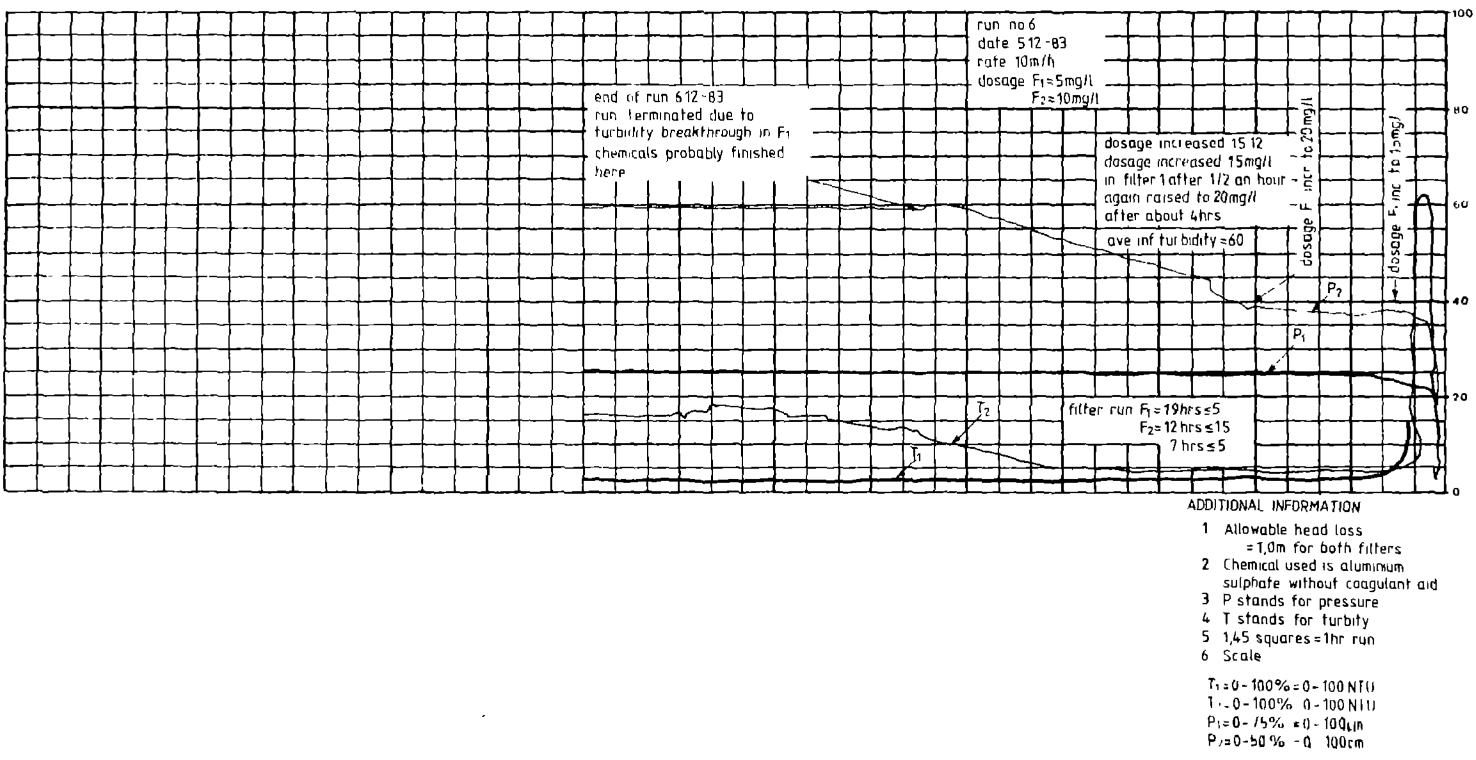
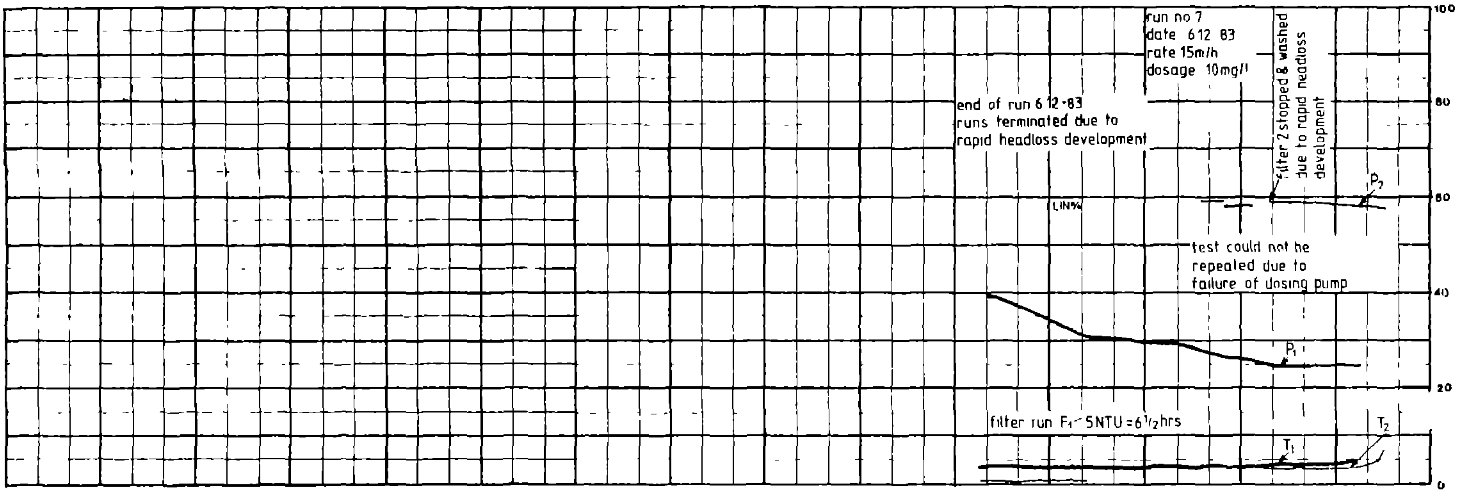


Figure 57. Run 6



ADDITIONAL INFORMATION

1 Allowable head loss
= 1,0m for both filters
2 Chemical used is aluminium sulphate without coagulant aid
3 P stands for pressure
4 T stands for turbity
5 1,45 squares = 1hr run
6 Scale

$T_1 = 0-100\% = 0-100$ NTU
$T_2 = 0-100\% = 0-100$ NTU
$P_1 = 0-75\% = 0-100$cm
$P_2 = 0-50\% = 0-100$cm

Figure 58. Run 7

The results of the direct filtration tests have been summarized in table 7. The summary also highlights the difference in the results obtained from the two turbidimeters. Whereas the relative differences have not been significant for all practical purposes as stated before, the actual readings have special relevance in evaluating the effluent quality. For instance, for run No. 4 the on-line turbidimeter shows no effluent quality equal to or less than 5 NTU. In contrast, the Hach 2100 turbidimeter indicated effluent quality values very close to the criteria for the attended observation periods of 8 and 7 hours in the case of run 4 and run 5 respectively.

Table 7. Summary of the results on the direct filtration tests

Run No.	Rate (m/h)	Influent Turbidity (NTU)	Dosage (mg/l)	Filter run (h)			
				Effluent Turbidity (NTU)*			(NTU)**
				$\leq$15	$\leq$10	$\leq$5 (WHO)	$\leq$5 (WHO)
4		30 - 90	2,5	11	4	-	8
4	5	30 - 90	5	13	11,5	11,5	8
3		30 - 180	10	23	16	16	
5		90	2,5	poor	-	-	
5	10	90 - 160	5	6	-	-	7
6		40 - 100	10	12	7	7	
6	10	40 - 100	15		19	19	Similar
			20				
7	15	40 - 220	10		6,5	6,5	
7		30 - 50	20		2,5	2,5	Similar

* Results according to the on-line Great Lakes turbidimeters as plotted

** Results according to Hack 2100 turbidimeter for the period of observation

Although the results of the test runs without chemical addition indicate removal efficiencies in the order of 50 % (runs 1, 2 and 3 on table 6), it was not possible to reduce the effluent turbidity to an acceptable value according to the criteria established from the influent turbidities tested.

12.31 Rapid mixing and flocculation

As has been pointed out earlier, no mechanical rapid mixing device was incorporated into the process. No separate flocculation unit was provided. Visible flocs were not detected in the supernatant even at higher dosages (10 and 15 mg/l) of alum. Despite this the filters performed well. The possible explanation consistant with theory is that adequate destabilization had occured during the hydraulic mixing stage and flocculation had taken place in the media. Indeed flocs were observed on the media surface just below the water-media interface and up to the end of the clogged zone. Thus it can safely be concluded that for direct filtration with alum alone as coagulant mechanical rapid mixing and separate flocculation units can be dispensed with.

12.32 Effects of turbidity and dosages

From the results on tables 6 and 7 it is apparent that for a particular rate and similar ranges of turbidity, an increase in dosages produced improved effluent quality values and longer filter runs. For ease in comparison the relevant data from table 6 has been reorganized and presented in table 8.

Table 8. Effect of dosage on turbidity removal

Rate (m/h)	Influent Turbidity (NTU)	Dosage (mg/l)	Approx. Effluent Turbidity (NTU) Hach 2100	Reference data on table 6
5	30 - 90	2,5	5	Run No. 4 (date 1.12.83)
	30 - 90	5	2	
	30 - 75	10	0,4	Run No. 3 (date 28.11.83)
10	60 - 90	2,5	9	Run No. 5 (date 3.12.83)
		5	6	
	40 - 100	10	1	Run No. 6 (date 5.12.83)
		15	1	
		20		
15	30 - 50	10	2	Run No. 7 (date 6.12.83)
		20		

Observations made during the experiments indicate that the effluent turbidity is sensitive to influent turbidity fluctuations at relatively lower dosages. This effect was remarkably noticeable even at the rate of 5 m/h with dosages of 2,5 and 5 mg/l. The sensitivity of the effluent quality to fluctuations of the influent turbidity was observed to insignificant for the higher dosages even with higher rates. It is noteworthy also that higher dosages resulted in shorter filter break-in periods for all the rates tested. However, this does not imply that low dosages should not be applied. In this experiment dosages as low as 2,5 mg/l have produced acceptable effluent quality values, for instance, at a filtration rate of 5 m/h for a range of turbidity of 30 - 90 (ref. tables 6 and 7). The significance of this is that during periods when the raw water source is of fairly low turbidity the reduction in dosage rates is possible.

The observations indicated (table 8) also that there is some dependence of the dosage requirements on the filtration rate. Although the available data is not sufficient to draw any conclusions from, it seems plausible that the higher shear intensities require higher dosages to increase the shear strength of the floc.

12.33 Effect of depth on removal

Although removal is taking place throughout the whole depth of the filter bed, visual observation as well as the piezometer readings recorded (table 6) confirm that the clogging advances gradually from the top of the media to the bottom with time.

Since the media is uniform, the curve for removal of turbidity against depth should theoretically be logarithmic. The plottings of the typical curves (figure 60) for the filter indicate this to be roughly so. From the curves it is evident that most of the removal occurs in the first 80 cms. The additional depth providing final polishing. This information alone, however, is not adequate for determination of the economic depth of the media. It is imperative to carry out optimization experiments. The minimum depth which provides the most removal at the time when the allowable head loss is consumed just before turbidity breakthrough occurs should be determined from the relevant optimization studies. It is noteworthy that in this study the 1 m depth media has performed as well as the deeper bed.

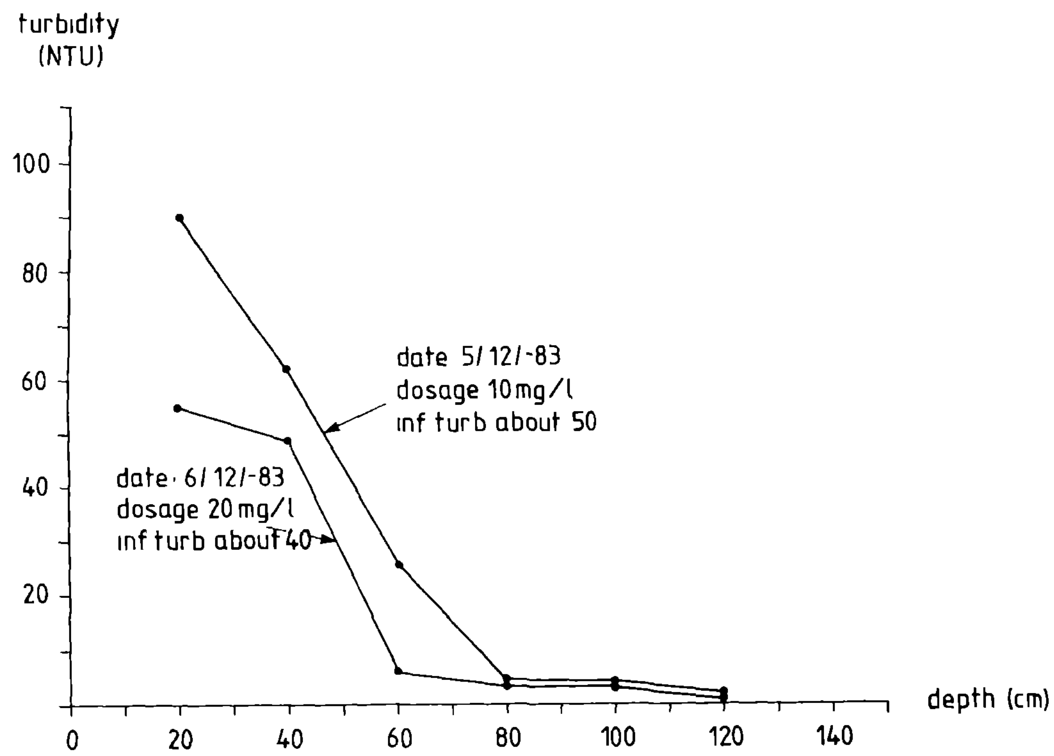


Figure 60. Effect of depth on turbidity removal

12.34 Backwashing of the test filters

The downflow test filters were backwashed mostly with simultaneous air and water without expansion of the bed. Rinsing was carried out at slightly higher rates than the normal backwash rate. Within about half an hour a degree of clarity of less than about 5 NTU was attained invariably. The need for the backwash water was about 10 % of the throughput as measured.

The filters were also at times washed with water alone with a bed expansion of about 20 %. It has been observed that this method is also effective but convection currents were clearly visible in the media during backwash. The effect of this on filter performance has not been investigated.

13. CONCLUSIONS

1. HRF is capable of producing an acceptable quality of water for direct filtration purposes from turbid sources if properly designed. Turbid waters containing suspended solids of up to about 600 mg/l could be efficiently filtered provided that the HRF is designed to operate at filtration velocities which correspond to the Reynold's numbers at the onset of turbulance.

2. The determination of the cross sectional area and length for the design of an HRF plant would depend on local conditions. Therefore, it is imperative that pilot plant tests are made in order to find out the appropriate dimensions. Similar techniques of prediction attempted in this work could be employed to get some idea of the minimum length of the coarse gravel required for the HRF from the relevant pilot plant test data. It should be noted that this minimum length of coarse gravel could be reduced but with the attendant risk of faster clogging of the fine polishing media before the silt storage capacity of the former has been fully utilized.

3. From the tests carried out on the down-flow test filters, it is evident that raw waters of turbidity variations up to 180 NTU could be treated producing an effluent quality turbidity that is acceptable according to WHO standards of less than 5 NTU. Such results were attained with the use of aluminium sulphate alone as coagulant.

   Dosages as low as 2,5 mg/l gave acceptable effluent quality values at the filtration rate of 5 m/h for turbidities in the range of 30 - 90 NTU. Much better effluent quality values and considerably longer runs were obtained for a dosage of 10 mg/l for the same rate. At the rate of 10 m/h a dosage of 10 mg/l gave acceptable effluent quality values whereas a dosage of 20 mg/l produced much better effluent quality and an appreciably longer filter run for similar influent turbidity fluctuations (40 - 100 NTU). Filter runs for the rates of 5 m/h and 10 m/h were terminated due to turbidity breakthrough. On the other hand at the rate of 15 m/h filter runs were terminated due to rapid head loss developments.

4. Although the use of polyelectrolytes as coagulants and/or coagulant aids has not been studied in connection with the direct filtration tests here, from the literature survey it is obvious that their use would yield better effluent quality at considerably lower dosages. However as the number, type and properties of such polymers are numerous and varied care should be taken not to apply them indiscriminately. The problem of continuous availability and cost should also be investigated.

5. The media used in the HRF (18 - 32 mm, 4 - 8 mm) and the down-flow filters (0,8 - 1,2 mm) is uniform (u.c. < 1,3) and coarse. This is of particular significance in both cases due to the prolonged length of filter run it affords as well as its simplicity for construction, operation and maintainance. In the case of the down-flow filters, the problems commonly associated with multi-media filter beds such as media intermixing and loss of material during backwash are avoided. The problem of stratification after backwash is also avoided. The results of the experiments confirm that such monograded coarse media perform very well.

6. Direct filtration with HRF as pretreatment is a technically viable treatment method for turbid surface waters at filtration rates of up to 15 m/h. Due to the reduction in chemical requirements, elimination of flocculation and sedimentation units it is possible that direct filtration with HRF could also be an economically feasible alternative to the conventional treatment process.

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