

Process Analysis and Optimizing of Direct Horizontal-Flow Roughing Filtration

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DISSERTATION

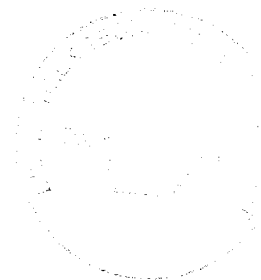
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by

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ABSTRACT

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The demand for safe water is increasing drastically in the urban and urbanizing areas of the developing countries. Many tropical rivers show wide variation in suspended solids and other water quality parameters. Water treatment plants which draw water from such rivers are facing growing problem in delivering desired drinking water quality as well as quantity. For the towns and small cities in developing countries this problem is compounded by their limited financial resources. Development of appropriate water treatment technology is therefore required. Horizontal-flow roughing filtration (HRF), a pretreatment step prior to slow sand filtration, requires low filtration rate (0.5-1.5 m/h) and thus large land area; its application is predominantly limited to rural areas. Process modification of the HRF for application in the urbanized areas is the objective of this study.

Investigations were carried out in lab-scale pilot plants in a controlled environment. Synthetic raw water (with kaolin as suspended matter) was used to simulate highly turbid river water.

In order to improve roughing filter performance and have it applied at a higher filtration rate the process incorporates direct filtration i.e. coagulant is added in a rapid mixing unit prior to HRF. The combined process is called direct horizontal-flow roughing filtration (DHRF). Comparing the performance of HRF and DHRF, it was found that DHRF systematically performed better featuring higher removal efficiencies at higher filtration rates. The first DHRF compartment with coarse grains (20 mm) acted as a multiple-plate settler whereas the second compartment with finer grains (8 mm) exhibited the characteristics of deep bed filtration. It was also found that horizontal filtration mode has higher substantially particle removal efficiency than vertical mode.

A mathematical model to predict the separation of flocculant particles in the gravel bed (grain size < 15 mm) was developed based upon an analogy with parallel plate settlers. Here both the flocculation and sedimentation processes are incorporated. A procedure was also formulated to use in the model the experimental data from column settling tests with actual suspensions for design and operation of DHRF.

The various coagulation and clarification process mechanisms occurring in direct horizontal-flow roughing filtration were investigated. An integrated alum coagulation stability diagram for highly

turbid water (200 NTU) was developed in which the zones of dominant coagulation mechanisms (e.g. adsorption-destabilization, sweep coagulation, combination of sweep and adsorption-destabilization) are demarcated in terms of coagulant dose and pH. These zones in general were shifted towards higher coagulant doses as compared to typical zones for in suspensions of low turbidity water (20 NTU). Coagulation in DHRF is usually achieved in the combined sweep and adsorption-destabilization zone.

The boundaries of flocculation, sedimentation and filtration process in a granular bed have also been demarcated in terms of grain size and filtration rate. The use of coagulants increase the filter coefficient to about 4 fold in 2 compartment DHRF as well as in rapid sand filtration.

The optimum process parameters for DHRF with high turbidity water (200 NTU) was investigated. An optimized DHRF would typically be a 2 compartment filter unit consisting of 4 m first compartment with 20 mm grains and 4 m second compartment with 8 mm grains. Optimum coagulant dose was about 1 mg Al(III)/l and optimum G value in the rapid mixing unit $200-300\text{ s}^{-1}$ with 1 min detention time. DHRF was found to be a versatile pretreatment process in handling wide fluctuations in raw water turbidity (100-400 NTU) with operating conditions like coagulant dose, mixing intensity and time, etc., remaining unchanged. In the presence of humic substances the turbidity and colour removal reduced drastically in HRF and in DHRF. Satisfactory turbidity and colour removal could be achieved only in DHRF by increasing the coagulant dose, decreasing pH, or both. The filter run time in presence of humic substances is however shortened to about half (approximately 2 days).

Surface washing was found to be an appropriate filter cleaning technique. Use of coagulants produced deposits which were more easy to clean. Investigation also suggested that deposits became increasingly difficult to clean with increasing deposit age and thus DHRF should be cleaned at least once a week.

A design guideline for DHRF is proposed which is applicable for predominantly mineral particles in the suspension and for optimized process conditions. For other water types and process conditions column settling tests and jar tests are useful tools for predicting the DHRF performance. A comparative analysis between HRF, DHRF and the conventional process consisting of coagulation, flocculation and sedimentation suggested that DHRF is an appropriate pretreatment technology for towns and small cities in many developing countries.

Key words pretreatment, turbidity, roughing filter, horizontal-flow, vertical flow, developing countries, appropriate technology, optimization, process mechanisms, coagulation, flocculation, sedimentation, filtration, colour, column settling tests, jar tests, filter cleaning, design.

Chapter 1

Development of Water Treatment Technology: Needs and Constraints

ABSTRACT - The demand for safe water is increasing drastically in urban and urbanizing areas of the developing countries. Many tropical rivers show wide variation in suspended solids and other water quality parameters. The towns and small cities which draw water from such rivers are facing growing problems in delivering desired drinking water quality as well as quantity. On the other hand, financial resources of those urbanized areas are limited. Development of appropriate water treatment technology is therefore required. Process modification for improvement of the horizontal-flow roughing filter is the objective of this study.

BACKGROUND

More than 100 countries have advanced to independent nationhood over the past half a century and development is their prime objective. The concept of development was regarded to be economic development after the Industrial Revolution in the Western world. Since the late 1960s this concept has undergone radical changes and it now also incorporates criteria of equity, quality of life and sustainability; the "basic needs" approach (e.g. food, health, housing, education, etc.) emerged (Clarke, 1985). The new model of development was also linked with the concept of national self-reliance. Several other interpretations of development were proposed (e.g. da Costa, 1980; WASH, 1993). Although at present no generally accepted definition of development exists, it is widely understood to be a process resulting at least in improved health and longevity, higher production and living standards, enhanced local problem-solving capability, and increased access to essential goods and services (World Bank, 1992; WASH, 1993).

One of the most important sectoral activities for development has been the provision of safe water supplies to the urban and rural communities. Water supply is a fundamental building block in the development process influencing directly health, economic development, employment, and women involvement in the society. Progress in this sector, preferably together with sanitation, has been most instrumental in reducing infant mortality and morbidity, and for overall improvement in health. The classical example is the drastic impact on health when these services were improved in the industrial countries in the nineteenth and

twentieth centuries. The life expectancy in French cities increased from 32 years in 1850 to 45 years in 1900, with the timing of changes corresponding closely to improvements in water supply and wastewater disposal (Fig. 1) (Preston and Van der Walle, 1978; Briscoe, 1985). Today, water and sanitation services are just as vital: diarrhoea caused death rates are typically 60% lower among children in households with adequate facilities than in those without such facilities. An estimated 80% of all diseases and over one third of deaths in developing countries are caused by the consumption of contaminated water. On average as much as one tenth of each person's productive life in the developing countries is sacrificed to water related diseases with concomitant consequences for education, productivity and well-being.

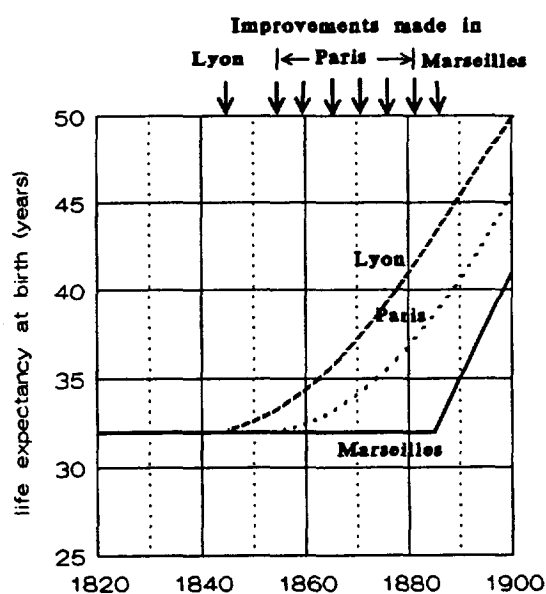


Fig. 1 Life expectancy and improvements in water and sanitation in selected French cities, 1820-1900 (Preston and Van der Walle, 1978; Briscoe, 1985).

The World Bank's (1992b) estimate on the effect of providing access to safe water and adequate sanitation to all who lack it would include, for example:

- 2 million fewer deaths from diarrhoea each year among children under five years of age (about 10 million infants die each year in developing countries from all causes), and
- 200 million fewer episodes of diarrhoeal illnesses annually (out of about 900 million episodes of diarrhoea occurring each year).

The first global policy to increase the provision of drinking water and sanitation was designed

at the 1977 UN Conference in Mar del Plata where the International Drinking Water and Sanitation Decade (1981-90) was launched (UN, 1977). The objective was to achieve safe water and sanitation for all by the year 2000. Development in this sector is taking place against a setting of scarce resources, conflicting priorities, human resource and institutional limitations, and rapid changes in the socio-political environment of the countries. For example, the population growth has so far outstripped the increase in facilities developed over many years of investment. The absolute number of people unserved remained almost the same as when the Decade began. Table 1 shows the number of unserved persons (world-wide) at the end of the International Drinking Water and Sanitation Decade (WHO, 1992). During this period 1.2 billion new users were provided with new facilities in developing countries and the water supply coverage increased from 46% to 68%. In 1990 about 1.3 billion persons, which is more than 20% of world population, still remained unserved, however (unserved: depending on potentially unsafe and/or unreliable water sources found in nature).

Table 1 Population without safe water supply, 1980-1990 (WHO, 1992).

	1980	1990
urban	235	204
rural	1,511	1,089

In addition to the large size of the population requiring safe drinking water there are several related concerns. Some important ones are discussed below.

Rapid urbanization and water demand

The second half of the twentieth century has been a demographic watershed. By mid-century the rate of population growth in developing countries had risen to unprecedented levels as mortality rates declined and life expectancy increased. The world population growth peaked at 2.1% per year in 1965-70 but by 1990 slowed down to 1.7% though remaining high in a number of developing countries. The present world population is about 5.5 billion and is increasing by 93 million a year. There is also a heavy migration in the developing countries of rural population towards urban areas. In 1990 most people lived in rural areas. By 2030 the opposite will be true: the urban population will be twice the size of the rural population. The developing country city population will grow by 160% over this period (World Bank, 1992b). Fig. 2 shows the present and projected rural and urban population in the Asia and Pacific region. It is forecasted that the global economic output share of the developing countries will increase from 48% in 1994 to about 63% in the year 2000 (Wooddall, 1994).

With the expected increase in income level the per capita water consumption of those who are already served would also increase (Briscoe, 1993). Thus, by the year 2030 the urban population will have risen threefold whereas the domestic water demand may rise fivefold (Briscoe, 1992; World Bank, 1992b).

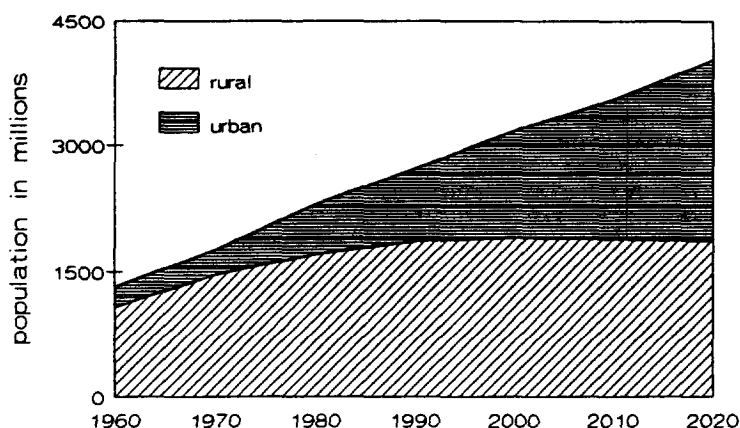


Fig. 2 Rural and urban population in Asia and the Pacific (World Bank, 1992).

Fig. 3 also shows that the global population doubled from less than 3 billion in 1960 to 5 billion in 1990; it may increase to 8 billion by 2010. Moreover, the share of urban population will rise from 15% to 55% between 1960 and 2010. This will have a major impact on the strains upon the water sector, as urban populations call for more water supply and sanitation, food ensured by efficient irrigated agriculture. In addition the technical and institutional requirements will become more complex.

A large portion of the urban population lives in towns and small cities with population less than one hundred thousand. For example, Table 2 provides a distribution of city and town sizes of Bangladesh. Most of these towns and small cities do not enjoy the favourable state funding and economic conditions of in the country's metropolises, and may face the most serious constraints for providing urban infrastructure, as a large share if not most of the urban growth takes place in these areas.

High cost and low income level

The availability of water resources is reaching its limits in many regions. Surface water near towns and cities is becoming increasingly polluted and asks for more elaborate treatment steps. As a result the cost per m^3 of new water production facilities is expected to typically increase two to three times; the ratio between the present and future cost of water treatment in some major cities in developing countries is illustrated in Fig. 4.

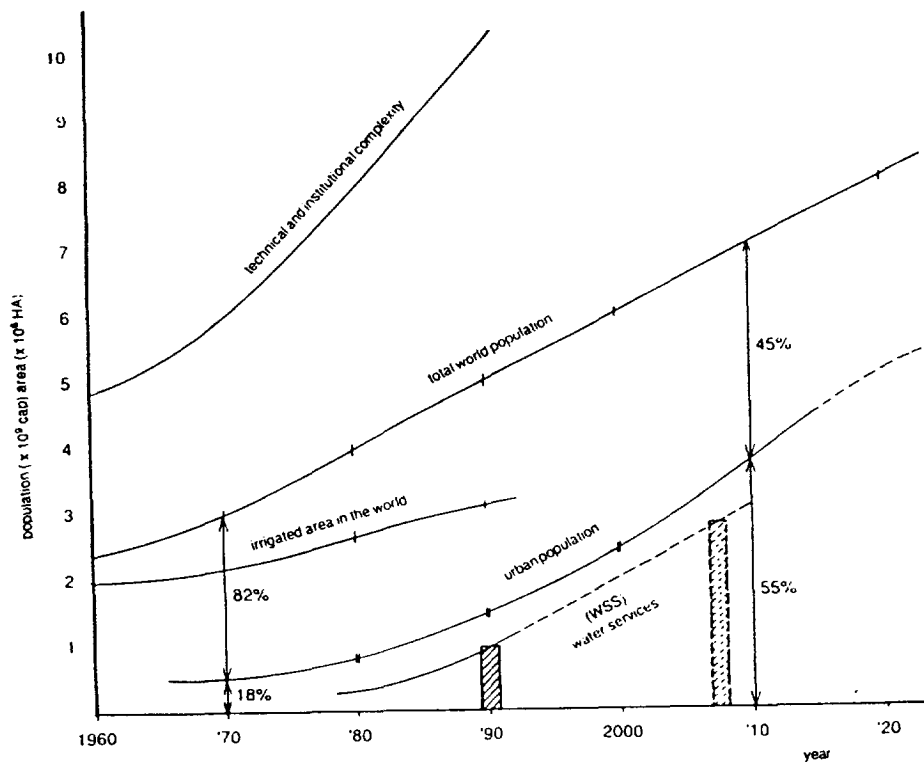


Fig. 3 Growth of global population and the water sector (Alaerts, 1991).

Table 2 Size distribution of Bangladeshi municipalities (Bangladesh Bureau of Statistics; 1993). The figures are according to 1991 census. The total population in 1991 was 111 million and in 1995 is estimated to be 120 million.

category	size (inhabitants) in thousand	number	urban population	percentage of total urban population
metropolitan city	> 500	3	5.6 million	48%
large city	100-500	17	2.6 million	22%
small city	50-100	21	1.5 million	13%
town	10-50	66	2.0 million	17%
totals:		107	11.7 million	100%

The percentage of population without safe water is a direct function of the per capita income (World Bank, 1992b). The capacity to satisfy the high water demand at increasing production cost is mainly restrained by the country's limited financial resources. The number of people living in poverty - living on the equivalent of US\$ 1 a day or less - was 1.1 billion in 1985, around 1992 estimated to be 1.2 billion, and expected to be 1.3 billion by 2000 (Ravallion et al., 1992; Ramphal, 1993; Kerri, 1995). The developing nations are confronted with shortage of funds and increasing competitive demands for scarce financial resources.

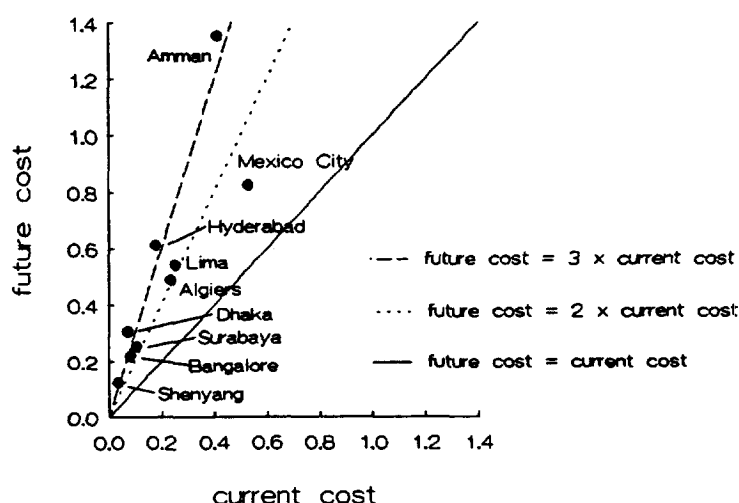


Fig. 4 Current and projected future cost of water treatment (1988 dollars per m³ water). Current cost refers to cost at the time data was gathered and future cost is a projection of cost under a new water development project (le Moigne and Briscoe, 1991).

Operation and maintenance

The problem of providing water supply facilities is compounded by the lack of proper operation and maintenance which is considered to be the biggest issue facing the water and sanitation sector (WHO, 1990; WASH, 1993). A decade ago, on average three years after construction more than half of all public water supply schemes in rural and urban fringe areas of developing countries are no longer functioning properly (Erbel, 1983); there is not much reason to believe that this situation has improved drastically. The World Bank review covering more than 120 projects over 23 years concluded that public water and sewerage utilities have reached acceptable level of performance in only 4 developing countries (Singapore, Korea, Tunisia and Botswana) (World Bank, 1992a).

Water Supply coverage

To achieve full, but basic, coverage in water and sanitation by the year 2000 using

conventional technologies and approaches would require an annual investment five times its current level (\$ 50 billion). However, a two-pronged realistic alternative strategy has been suggested in the New Delhi Statement (UNDP, 1990):

- substantial reduction in cost of services, through increased efficiency and use of low-cost appropriate technology, and
- mobilization of additional funds from existing and new sources.

Thus, if the costs were halved and financial resources at least doubled, universal coverage could be within range by the end of the century. This view was also endorsed by the UN Conference on Environment and Development in Rio de Janeiro - Agenda 21.

If the present water supply policy remains unchanged ("business-as-usual") this would lead to a rise in the number of people unserved as a result of rising unit cost and in population growth. If the investment in water supply is increased by 30% the situation will be improved but the number of unserved population would still rise. A combination of accelerated investment and efficiency reform can make substantial improvement; these three scenarios are illustrated in Fig. 5. The efficiency reform includes the more widespread use of appropriate technology, capacity building, improved maintenance, community participation, etc. Thus, development of appropriate water treatment technology still plays a key role in this endeavour.

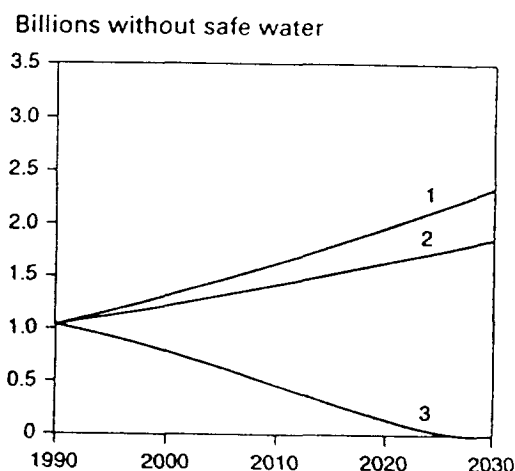


Fig. 5 Three scenarios for safe water supply, 1990-2030: 1. "business as usual", 2. 30% increased investment, and 3. accelerated investment and efficiency reform (World Bank, 1992).

RIVER WATER QUALITY

Rivers are the most important fresh water resource for humans. Social, economic and political development has, in the past, been largely related to the availability and distribution of fresh water in the riverine systems. Rivers are a major source for drinking water supply. Important quality considerations for drinking water are the suspended solids (SS) content, the chemical composition (including natural organic matter) and the bacteriological quality.

SS cause turbidity and are undesirable in drinking water. Also, they interfere with disinfection during water treatment by creating a shielded shelter for disease causing organisms. SS are also considered a major pollutant carrier for many toxic heavy metals, organic pollutants, pathogens and nutrients (Martin and Meybeck, 1979; Meybeck, 1982; WHO/UNEP, 1989).

The major sources of SS in rivers are (WHO/UNEP, 1989): (i) products of continental rocks and soil erosion, (ii) the autochthonous material which is formed within the water body and which usually results from the production of algae and precipitation of minerals, mostly calcite, and (iii) the anthropogenic sources resulting from various human activities. The most prominent amongst these is mechanical erosion of rocks and soil which results from the combined effects of various erosion agents, i.e. running water, wind, temperature differences, chemical weathering, moving ice, and mass movement of materials on slope. Soil erosion is highest in mountainous areas and active volcanic regions. In particular, it is enhanced when climate is characterized by alternating wet and dry seasons as in tropical areas (e.g. monsoon climate in South-East Asia). SS mostly consist of clay minerals such as kaolinite, montmorillonite and chlorite, and also of quartz, feldspars, oxides and organic matter (Drever, 1988).

The chemical composition of a river water depends on several factors, including (Hem, 1985; UNESCO et al., 1992): (i) the proportion of surface run-off and ground water; (ii) the physico-chemical reactions within the river system governed by internal processes, (iii) the mixing of water from tributaries of different quality (in case of heterogeneous river basins), and (iv) inputs of pollutants. The main sources of elements in rivers are (i) chemical weathering, (ii) atmospheric inputs, and (iii) leaching of organic soils. Most chemical weathering reactions derive from the attack of minerals, mostly aluminosilicates, by carbonic acid ($\text{H}_2\text{O} + \text{CO}_2$). This leads to the formation of major cations (Ca^{2+} , Mg^{2+} , Na^+ , K^+) and of dissolved silica (SiO_2) and bicarbonates (HCO_3^-). The least soluble elements (Al, Fe, Mn) remain in the soil which gradually become more enriched in these elements. As a result of this relative enrichment, the soil particles, which are eventually erode during heavy rains, are quite different from parent rocks.

Leaching of organic soils also generates nitrogen, phosphorous and dissolved organic matter in surface water. Human activities such as deforestation, agriculture, etc., can enhance natural processes like erosion and soil leaching, and the input of natural compounds such as inorganic salts and fertilizer, or adds synthetic compounds as pesticides and insecticides (Lamb, 1985; UNESCO et al., 1992). Deforestation alone is reported to be able to raise the SS in rivers 100 times (UNEP, 1991). Urban pollutants, usually treated or untreated sewage effluent may enter the river water.

Variation of water quality

During flood water quality shows marked variation due to different origins of water: surface run-off, sub-surface run off (i.e. water circulation within the soil layer), and ground water discharge. Surface run-off is generally highly turbid and carries a large amount of SS, including particulate organic carbon. Sub-surface run-off leaches dissolved organic carbon and nutrients (N and P) from soils, whereas groundwater provides most of the elements resulting from rock weathering (SiO_2 , Ca^{2+} , Mg^{2+} , Na^+ , K^+). Bicarbonate (HCO_3^-) determines the water buffering capacity and influences pH; it is generally present between pH 6 and 8.2 and derives partly from dissolved carbonate-bearing minerals and partly from CO_2 in the soil.

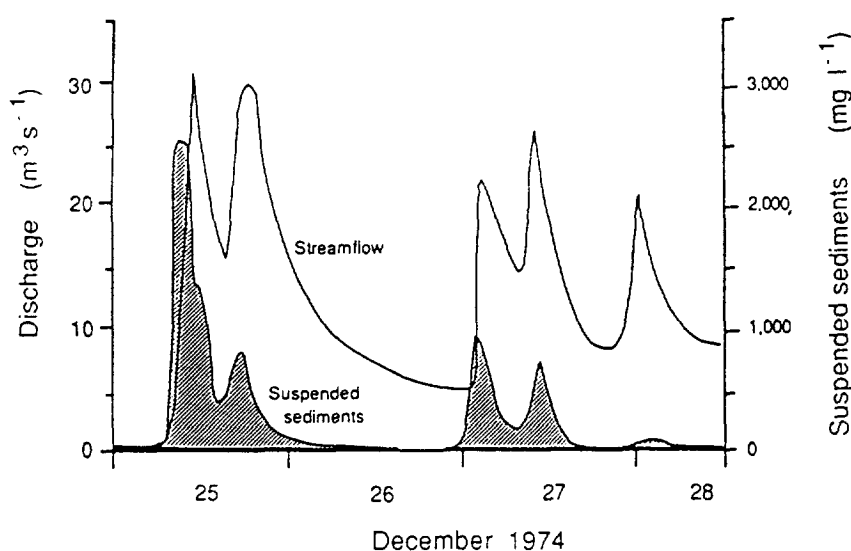


Fig. 6 Temporal relationship of total suspended soils to the discharge of River Exe, Devon, England (Walling, 1977).

The concentration of SS varies dramatically with discharge e.g. during flooding. This is illustrated in Fig. 6 for river Exe in England, where a general correlation occurs with the

peak of SS close to the peak of river discharge. The SS varied from 15-2,500 mg/l, i.e. in excess of two orders of magnitude. Similar relationships were also reported for e.g. Neuse River (Harned, 1982) and Freanch Broad River (Daniel et al., 1982) both in North Carolina, and in River Seine at Paris (Chevreuil et al., 1988).

One of the very few examples of water quality parameter frequency analysis is for the Cauca river valley, Colombia; Fig. 7 shows that the turbidity peaks for the 6 rivers in the valley are generally 2 orders of magnitude higher than their average turbidity. The duration of the peak turbidities vary from 20 days to 90 days in a year. It is also reported that the occurrence of total number of turbidity peaks in a year higher than 1000 NTU vary from 1 to 25. The occurrence of the rate of turbidity increase that is higher than 100 NTU/h vary from 15 to 50.

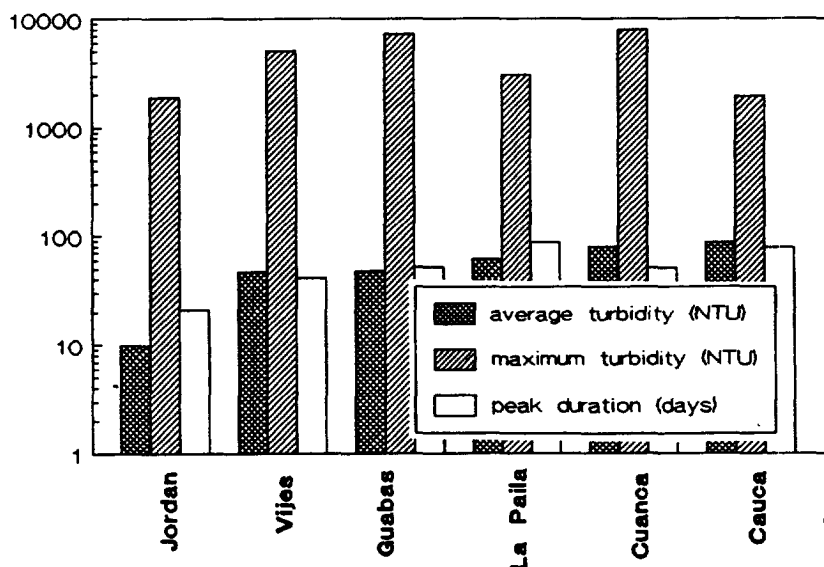


Fig. 7 The average turbidity, the maximum turbidity and the total duration of turbidity peak in the rivers of Cauca valley, Colombia (CINARA, 1995).

The chemical composition of the water also varies with discharge, however, the relationship is dominated by the source of the elements (Wolder and Simmons, 1982; WHO/UNEP, 1989). Substances normally introduced at constant rates (major cations and bicarbonates from the soil) have their input diluted with increasing discharge (increased surface run-off due to rainfall). Thus, during high discharge the buffering capacity of water and pH tends to decrease. The concentration of P and N from sewage outfall also reduces with discharge. Organic matter which is generally related to the flushing of soil during run-off has a limited increase. However, pesticides and insecticides, and other metals associated with SS show

substantial increase. Table 3 shows for selected rivers the water quality demonstrating their wide quality fluctuations (WHO/UNEP, 1992).

Table 3 Water quality fluctuation ranges during 1982-84 of some selected rivers (WHO/UNDP, 1992).

rivers	country	pH range	suspended solids	BOD max	faecal coliform
			(mg/l)	(mg/l)	(no. per 100 ml)
Rhine	Netherlands	6.9 - 8.3	3 - 177	15	<1 - 540,000
Maas	Netherlands	6.7 - 8.3	1 - 310	36	80 - 80,000
Exe	UK	7.0 - 8.9	1 - 62	5	70 - 22,000
Sefid	Iran	7.7 - 8.3	36 - 31,700		170 - 16,00
Shimano	Japan	6.7 - 7.5	9 - 989	4	10 - 5,000
Mapocho	Chili	6.5 - 8.2	4 - 648	2	2 - 720
Balas	Mexico	6.7 - 8.4	10 - 11,900	3	9 - 240,000
Chao Phryma	Thailand	5.5 - 8.0	25 - 280	3	50 - 35,000

WATER TREATMENT TECHNOLOGIES

River water treatment is usually required for potable water supply. In a surface water treatment process sand filters (rapid or slow sand filters) are essential for final removal of impurities, notably SS and the associated impurities, and any particulate material generated during treatment process (such as flocs). The SS concentration that the filters can accept in the influent is limited. However, many rivers on the different continents, especially in tropical regions, show a wide fluctuation in flow as well as in SS, the latter typically 5 to several thousand mg/l. Pretreatment is therefore usually necessary; it helps in producing a desired water quality for subsequent sand filtration and yields considerable savings on overall operating costs (IRC, 1990; Galvis et al., 1993). Fig. 8 shows a schematic representation of a river water treatment scheme with pretreatment.

Table 4 describes common water treatment processes for SS removal. The table distinguishes different pretreatment options and filtration options for final treatment. Prolonged storage and plain sedimentation are relatively simple technologies. In plain sedimentation only coarse matter can be removed; only sand and clay down to 20 μm can be removed within a 2-3

hours detention time. However, depending on the particle size and density distribution the SS may not be low enough even after prolonged storage for direct feeding especially to slow sand filters. Therefore, sedimentation often is applied as a first treatment to remove a substantial part of SS before the other pretreatment steps. Flocculation-sedimentation is an effective particle separation process and generally used in conventional river water treatment scheme. However, it requires coagulant addition and operational control which can be difficult to achieve in rural areas. Flotation are used when raw water contains a high proportion of organic matter of which particle density is close to that of water. Flotation requires a very high degree of operational control. Conventional roughing filters is considered as a promising pretreatment technique for slow sand filtration for its simplicity and as no chemicals are needed. Horizontal-flow roughing filters can handle moderately high turbidity water and have long operating period from weeks to months depending on the raw water turbidity. It is a very simple method appropriate for rural areas in developing countries.

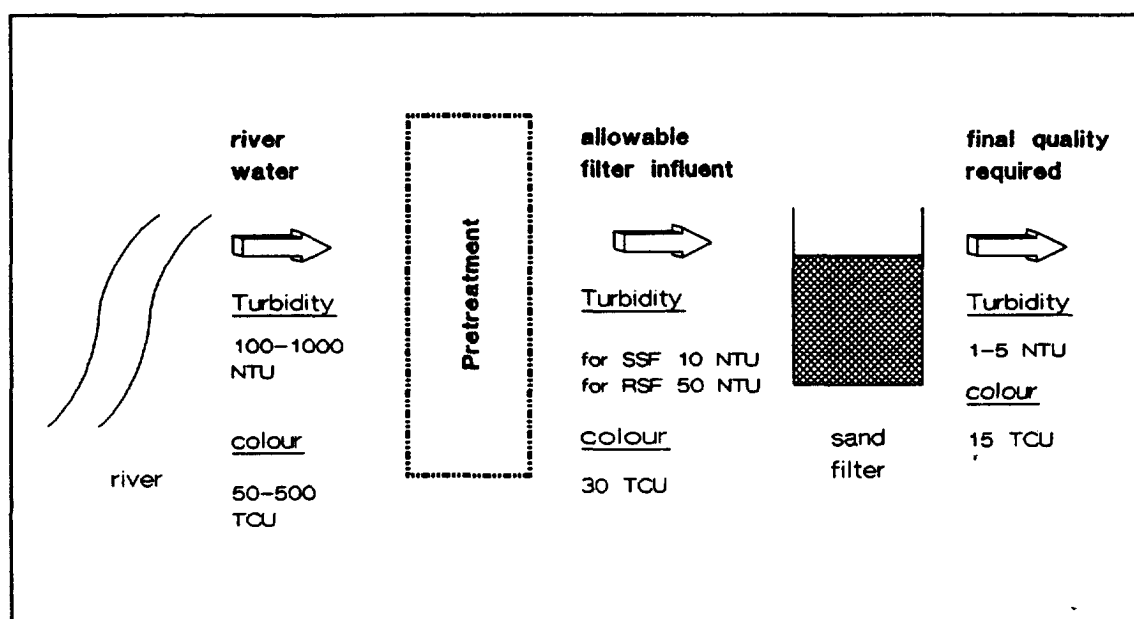


Fig. 8 Schematic representation of a river water treatment scheme with pretreatment, indicating the water quality ranges existing or desired at different locations in the process (SSF: slow sand filtration, RSF: rapid sand filtration, TCU: true colour unit measured in mg Pt-Co/l).

Roughing filtration

Roughing filters are defined as filters with grain sizes larger than 2 mm (Schulz and Okun, 1984). They are claimed to perform "natural" treatment process as no chemicals are generally used and as they do not require sophisticated mechanical equipment. They are thus

Table 4 Common processes for particle (SS) removal.

options	influent conc.on upper limit (mg/l)	removal efficiency (%)	stage of acceptance	supervision level	comments
A PRETREATMENT					
<u>sedimentation/flotation based</u>					
prolonged storage	no restriction	50 - 70	established	low	removes only settleable particles; algal growth risk
plain sedimentation	no restriction	30 - 50	established	low	only mineral particles >20 μ m are removed
flocculation- sedimentation	no restriction	90 - 98	established	medium	coagulants required; sensitive to water quality changes
flotation	20 - 100	90 - 98	established	high	coagulants and dissolved air required; sensitive to water quality changes
<u>roughing filtration</u>					
vertical-flow (up- or down- flow)	20 - 150	80 - 95	emerging	medium	moderate deposit storage capacity; filter cleaning problem
horizontal- flow	200 - 400	80 - 95	emerging	low	high deposit storage capacity; requires large filter volume; filter cleaning problems
B FINAL TREATMENT					
slow sand filtration	10 - 20	90 - 98	established	low	manual cleaning typically after 1 month; large area required
rapid sand filtration	20 - 50	90 - 98	established	medium	backwashing required typically after 24-48 h
direct filtration	20 - 50	90 - 99	established	high	coagulants to be added; backwashing typically after 12-24 h

an adequate pretreatment technology which is well-adapted to slow sand filtration. At the end of last century and the beginning of this century coarse gravel filters were extensively used in Europe (Baker, 1948). Eventually these filters were converted to slow and rapid sand filters. Since the 1960s, the roughing filtration technology has been revived in Europe, however, mainly applied in ground water recharge plants in combination with slow sand filters (Kuntschik, 1976; Nickl, 1983). Over the last decade roughing filtration was studied and developed as a simple and reliable pretreatment process prior to slow sand filtration for water treatment in developing countries (Gregory et al., 1983; Wegelin et al., 1991; Wegelin and Schertenleib, 1993).

Roughing filters are physical filters and intended to retain the majority of the SS. They are applied to treat surface water of high turbidity over prolonged periods. The grains are of successively decreasing sizes arranged in layers or compartments and usually range from 25 mm down to 4 mm. The larger grains allow deep deposit penetration. The bulk of the SS is retained in the first coarser grains which has a high deposit retention capacity at the expense of only low headloss. The subsequent finer grains have more of a polishing function.

Depending on the flow direction roughing filters can be classified as up-, down- and horizontal-flow filters. The filtration rate is low, usually 0.3-1.5 m/h. In vertical-flow roughing filters, the filter bed height is limited to 1-1.5 m due to structural constraints. Hence, the total filter bed length of a three-stage vertical-flow filter will maximally be 4.5 m. Alternatively, horizontal-flow roughing filters can be constructed with unlimited length, which generally varies between 5 and 9 m. Vertical filters can cope with maximum raw water turbidity of 50-150 NTU, whereas horizontal-flow filters can handle 100-400 NTU and short peaks of 500-1000 NTU (Wegelin et al., 1991). Roughing filters are cleaned by repeated fast filter drainage, and when necessary, grains are manually removed, cleaned and replaced.

Horizontal-flow roughing filtration

A horizontal-flow roughing filter (HRF) consists of a box of 3-4 compartments filled with grains of progressively smaller size in each compartment (Fig. 9). The raw water falls over a weir into an inlet chamber where the flow is evenly distributed over the vertical filter cross-section. The water passes in horizontal direction through coarse to fine grain compartments and is collected in an outlet chamber. Thereafter, the pretreated water is discharged over a overflow weir.

The length of the filter box is usually 5-9 m, its height is limited to 1.5 m to allow comfortable manual cleaning when required. The width varies from 2 to 5 m depending on required capacity. The coarsest grain is around 25 mm, and the finest not smaller than 4 mm.

Filtration rate is in the order of 0.5-1.5 m/h. An HRF design guideline specifying the number of compartments, their dimensions and grain sizes, filtration rate and expected removal efficiencies for various raw water turbidity was developed by Wegelin (1986). The effluent weir controls the water level within the filter. With filtration time the filter resistance increases due to clogging of the pores. The water level over the gravel bed thus increases which should be compensated by additional bed height over the outlet weir. Headloss in HRF should not exceed 30 cm; filter grains are filled approx. 30 to 40 cm over the weir level. This keeps the water level always below the gravel bed surface thus preventing direct sunlight from stimulating algal growth and introduction of other contaminations.

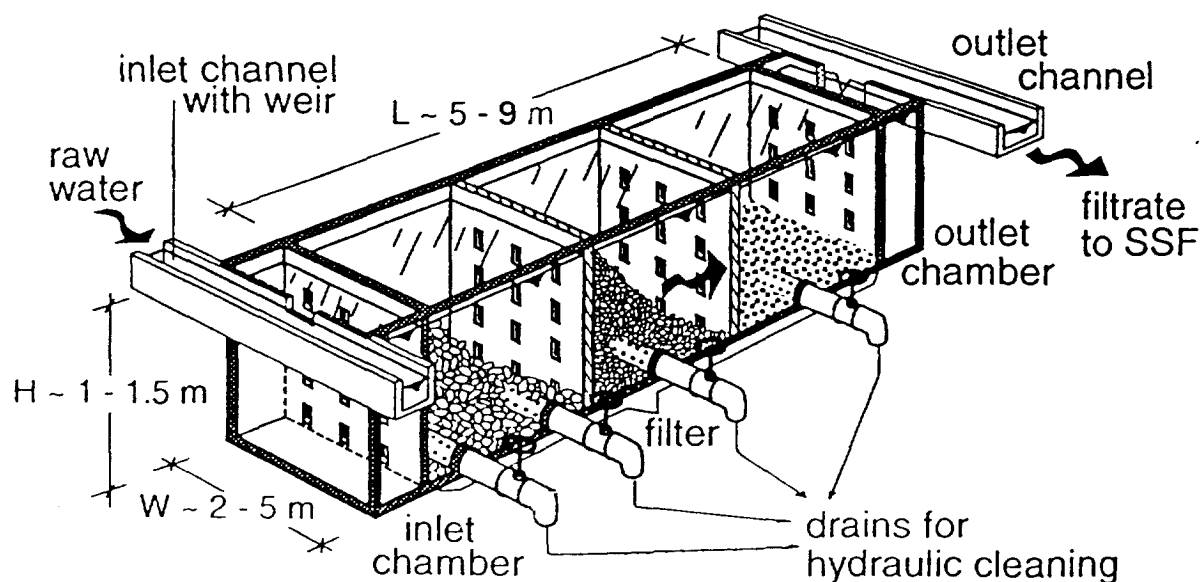


Fig. 9 Layout of a 3 compartment horizontal-flow roughing filter (IWCWD, 1992).

It is argued that sedimentation is the dominant particle removal process in HRF (Kuntschik, 1976; Wegelin, 1987; Boller, 1993). The filter acts as a multi-storied sedimentation basin providing a large surface area on which suspended particles settle. Deposits grow over filter grains in domelike shape. When the slope of the deposit exceeds its slope stability part of it drifts downward in small heaps. This drifting regenerates the removal efficiency of upper filter layers to some extent and prolongs filter operation.

HRF operates at a low filtration rate and requires large installations. Therefore, its application is predominantly limited to rural areas with cheap land in developing countries. HRF is applied prior to slow sand filters in Africa (Burkina Faso, Cameroon, Ethiopia, Ghana, Kenya, Malawi, Swaziland, Sudan, Tanzania and Zimbabwe), Asia (Burma, China, India, Indonesia, Iraq, Malaysia, Nepal, Pakistan and Sri Lanka) and in Australia and Papa

New Guinea. The construction cost is expected to be higher than for conventional flocculation-sedimentation process, however, operation and maintenance cost will be lower as comparatively these are simple and no chemicals are used (Wegelin et al., 1991). However, filter regeneration is not yet fully developed (Collins et al., 1992). After continuous filter operation and periodic hydraulic cleaning the unremoved deposits accumulate and will increase headloss and decrease operating time; manual cleaning therefore can be necessary after some years. Furthermore, as sedimentation is the dominant particle separation process, removal efficiency decreases when the raw water contains a high proportion of colloidal or organic matter.

Innovative application of roughing filters

Roughing filters can be used in the raw water intake to reduce the SS concentration to the subsequent treatment units. In intake filtration the river water is filtered while seeping down through a natural or constructed (roughing filter) river bed. The filtered water is collected by perforated pipes in the river bed (Engels and Poggenburg, 1990; Smet et al., 1990). Application is however limited to temporary water supply because of the gradual clogging of the bed. Due to the difficulties in gravel bed cleaning in a river a shallow (0.6 m) gravel bed, called dynamic roughing filter, can be constructed outside but near the river (Galvis et al., 1992; Galvis et al., 1993). Part of the raw water overflows and part is filtered through the gravel bed. It is expected that the overflowing water partially removes the clogging sediment from the gravel bed surface and hence increases operating time. Its limitation is that the overflow water is wasted and because of its declining rate mode the filtrate flow varies in time causing operational problems downstream. The technology is still in experimental stage and requires further development.

The pebble matrix filter is an offshoot of roughing filtration. It consists of a matrix of large pebbles of about 50 mm diameter. In the lower half of the pebble column the pores are filled with fine sand (Ives and Rajapakse, 1988; Ives, 1990; Rajapakse and Ives, 1990; Ives, 1991). It is intended to protect slow sand filters from highly turbid waters. The downflow filtration rate varies from 0.7-1.2 m/h. Laboratory tests suggest that kaolin clay suspensions of up to 5000 mg/l were reduced to below 25 mg/l. The filter run time was 4-5 days and headloss did not exceed 1.5 m. The filter is cleaned by draining repeatedly and then backwashing. This technology is, however, still to be field-tested.

A compact small-scale treatment system incorporating roughing and slow sand filter was constructed by providing two annular HRF units around one circular slow sand filter within one structure (Mbwette, 1994). A household filter (UNICEF filter) is being promoted by UNICEF East Africa Regional Office (Childers and Claasen, 1987). It consists of two cement tanks, a top raw water storage tank and a bottom filter tank. In the filter tank 25-30 cm

crushed charcoal (grain size 5 mm) is sandwiched between two 20-25 cm sand layers. Raw water flows upward through the filter. The performance of this filter was improved by coagulating the raw water with alum and adding a second downflow sand filter (Singh and Chaudhuri, 1993). The upflow-downflow filter is a two-stage filter made of an upflow roughing filter followed by downflow rapid sand filter and is usually constructed in one structural unit (Merwe et al., 1995; Schulz and Okun, 1984). It is applied in a few small treatment plants in Brazil, Costa Rica and India.

Roughing filtration is used as small iron removal plant coupled with hand pumps in Bangladesh (World Bank, 1991). The ground water is first aerated, then passed through a sedimentation tank and a roughing filter to remove iron flocs. Roughing filters can also be used as a flocculator i.e. gravel bed flocculator where the coagulated water is flocculated due to the velocity gradient developed along the water's tortuous path through the filter pores (Kardile, 1981; Bhole, 1993). Package treatment plants, called adsorption clarifiers utilize the flocculating and particle removal capacity of roughing filters (US EPA, 1990; Goodrich et al., 1992). Coagulated water flows upward through a low density plastic coarse media (roughing filter) where suspended solids are flocculated and partly retained. This is followed by downflow multimedia filtration.

TECHNOLOGY DEVELOPMENT - the way forward

Appropriate technology

A technology not only should be 'low-investment' so that its construction is within the financial possibilities of the developing countries, it should also satisfy operational and socio-economic considerations for effective operation and maintenance. The following considerations are generally accepted for a technology to be suitable in a particular situation (Erbel, 1983; WASH, 1993):

- the technology should be conceptually and physically within the capabilities of the persons responsible for operation and maintenance,
- spare parts and equipment must be available,
- operating cost should be within the financial means, and
- the technology should be attractive and produce a good standard of service.

Appropriate technology for towns and small cities

During the coming years rapid population growth is expected to be particularly pronounced in the towns and small cities of the developing countries, whereas the rural population will tend to stabilize. Large cities similarly will keep expanding. However, large cities are substantially stronger than small towns, because the former have better developed technical and financial capabilities. Providing adequate water supply to towns and small cities is thus becoming a priority. These areas are relatively developed, and goods and services reasonably available. The level of technology can be more sophisticated than in rural areas, however, should be much simpler in construction and operation than in the large metropolitan cities. The selected technology should benefit from the cheap semi-skilled manpower available.

Slow sand filtration is an appropriate and established surface water treatment technology for rural and urban areas alike (Huisman, 1974; Graham, 1988; Collins et al. 1992). Different pretreatment options for slow sand filters have been recommended (van Dijk and Oomen, 1978; IRC, 1979). Except for few cases, rapid sand filters (including direct filtration) are used in larger water treatment plants in large cities; design guidelines including for required pretreatment are available (e.g. Montgomery, 1985; Weisner and Mazounie, 1989; Degremont, 1991; Janssens and Buekens, 1993). However, a gap exists for appropriate technology in the towns and small cities which should be of an intermediate technology.

For the urbanized areas in towns and small cities slow sand filtration and roughing filtration become an expensive technology because of their large land requirement (which is costly in urbanizing areas) and construction volume. Moreover, in case of highly turbid raw water or when excessive organic matter is present the performance of simple pretreatment processes like sedimentation tanks and roughing filters are not adequate. Slow sand filters clog very quickly making it expensive to operate or the produced water quality in terms of organic matter removal is not satisfactory (Collins et al., 1992; Goodrich et al., 1992). Therefore, the development of water treatment technology to obtain higher efficiency and better quality, and also more compact design for cost reduction may be achieved at the expense of the available chemicals and skilled manpower.

OBJECTIVES

Many rivers are characterized by seasonal high turbidity and considerable fluctuation in the water quality. Generally pretreatment is required notably to remove the SS. Development of appropriate pretreatment technology for this purpose for towns and small cities is the focus of this study.

Roughing filtration is an emerging pretreatment technology which is also being innovatively

applied in diverse areas; full potential is yet to be determined. Different types of roughing filters are being developed and promoted. Horizontal-flow roughing filtration has a good removal efficiency and is able to cope with medium to high turbidity water; it can be considered for application in towns and small cities. However, its main limitation for urban application is that it has been mainly developed for small rural communities, and requires large installations. Other limitations like low efficiency when organic and colloidal matter are present in raw water, effective filter cleaning, and capacity to handle fluctuating river water quality require to be addressed. Further development of the horizontal-flow roughing filtration process in the direction of technology sophistication, which can be afforded by the towns and small cities, may be appropriate prior to slow or rapid sand filters.

The main objectives of this study are:

- to modify the horizontal-flow roughing filtration process for urban application in the context of appropriate technology for those areas, and
- to analyze the process mechanisms involved in the modified process in order to allow optimization and further development.

Against the above background the main hypotheses of this study are:

- the performance of horizontal-flow roughing filtration can be improved by combining the roughing filtration principle with direct filtration despite the very high SS loading on the filter consisting of original SS and floc material,
- the combined process would be compact and versatile, and capable of handling wide range of water quality. The coagulant requirement would be similar to direct filtration, and
- the particle removal mechanism in the large grain beds would be similar to sedimentation as in a settling tank, whereas the smaller grain beds perform more similarly to deep bed filtration.

Organization of the thesis

The background and importance of this study is discussed in Chapter 1. Chapter 2 tests the feasibility of the combination of horizontal-flow and direct filtration process. The influence of different process parameters is assessed in a general fashion. A model to describe the

flocculant particle removal in large grains in analogy to multiple-plate settler is developed in Chapter 3. In Chapter 4 the boundaries of different mechanisms occurring in the combined process are determined. The mechanisms of coagulation are demarcated in terms of coagulant dose and pH and the mechanisms related to particle removal in a gravel bed are demarcated in terms of grain size and filtration rate. In Chapter 5 the process parameters of the combined process are optimized; the influence of main raw water variables is assessed in Chapter 6. A potentially appropriate roughing filter cleaning procedure is developed in Chapter 7. Design and operational guidelines are provided in Chapter 8. Conclusions of this study are given in Chapter 8.

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Chapter 2

Direct Horizontal-Flow Roughing Filtration: An Appropriate Pretreatment Technology for Highly Turbid Water¹

ABSTRACT - Horizontal-flow roughing filtration (HRF) requires low filtration rate (0.5-1.5 m/h) and thus large land area; its application is predominantly limited to rural areas. In order to improve HRF performance and be applied at a higher filtration rate the process is combined with direct filtration i.e. coagulant is added in a rapid mixing unit prior to HRF. The combined process is called direct HRF (DHRF). Comparing the performance of HRF and DHRF, it was found that DHRF systematically performed better featuring higher removal efficiencies at higher filtration rates. The first DHRF compartment with coarse grains (20 mm) acted as a multiple-plate settler whereas the second compartment with finer grains (8 mm) exhibited the characteristics of deep bed filtration. It was also found that horizontal filtration mode has higher particle removal efficiency than vertical mode. DHRF could be an appropriate pretreatment technology for towns and small cities in developing countries and has a good potential to substitute the conventional flocculation-sedimentation process and be used before rapid sand filters.

INTRODUCTION

The horizontal-flow roughing filter (HRF) is claimed to have the advantage of being able to tackle relatively high raw water turbidity, whilst at the same time offering long filter run time and a simple technology (Wegelin, 1983; Wegelin et al., 1987). Its applicability is, however, limited e.g., due to its low filtration rate and difficulty to treat with coloured water (Chapter 1). In order to improve the HRF performance a modified configuration which combines it with other water treatment process(es) can be considered. Such a modified process which can overcome the limitations of HRF and on the other hand retain its advantages, notably long run time and ease in operation and maintenance, would be advantageous.

Direct filtration is a process alternative to conventional water treatment (coagulation,

¹Part of this Chapter was published by T. Ahsan, J. P. Buiteman and G. J. Alaerts (1991) *Proc. Filtration Society Conference, Karlsruhe, Germany*.

flocculation-sedimentation and filtration) and is suitable for water sources with low to moderate turbidities. It is defined as a flocculation based treatment system in which filtration is not preceded by sedimentation (AWWA, 1980). Two configurations are possible: after the rapid mixing unit, there may or may not be a separate flocculation unit prior to filtration. In the first option larger flocs are to be filtered out of the water (German: "Flocken-filtration"), whereas in the second microflocs are in development inside the filter (German: "Flockungs-filtration") and subsequently removed.

The upper limits of raw water turbidity and colour are the two variables which limit the application of direct filtration. Limits that are numerically of the same order of magnitude have been proposed. Culp (1977) suggested turbidity should be < 25 NTU and colour < 25 units (expressed as mg Pt/l). The AWWA Filtration Committee's report on direct filtration (1980) concluded that turbidity > 15 NTU and colour > 30 -40 units could create filtration problems. Edzwald (1987) suggested turbidity should be < 20 -30 NTU and colour < 30 -40 units (as Pt-Co/l). Others investigations (e.g. McCormick and King, 1982; Wagner and Hudson, 1982; O'Melia, 1985) also suggested that turbidity should be below 25 NTU. An evaluation of 100 direct filtration plants in the USA also indicated that in practice the raw water turbidity is generally < 25 NTU (Letterman and Logsdon, 1976). Similarly it is practice that a lower coagulant dose is applied in direct filtration as an excessive dose will limit filtration run time. For example, aluminium sulphate (alum) dose should be < 15 mg alum /l for satisfactory filter run (Wagner and Hudson, 1982).

When comparing direct filtration with conventional treatment, direct filtration is more sensitive to changes in raw water quality and coagulation conditions partly because of the lower detention time between coagulation and filtration. Therefore, higher standards of quality control and operator vigilance are required. The filter run time is shorter than that of common filters, usually < 10 -20 h (AWWA, 1980). On the other hand, overall capital costs in certain cases can be 30% lower (Culp, 1977; Wagner and Hudson, 1982; Janssens et al., 1986). In case of relatively low raw water turbidities, the finished water from direct filtration plants is generally similar to that of conventional plants (Collins et al., 1987) and of very good quality i.e. turbidity 0.1 to less than 1 NTU (Letterman and Logsdon, 1976; McCormick and King, 1982; Craig, 1985). The colour removal is about 90% at optimum coagulant dose and pH (Fettig, 1988).

Direct filters are applied to low turbidity water (< 25 NTU), however roughing filters can generally cope with higher turbidity (50-200 NTU) and conventional process has practically no limitation. To obtain a higher deposit retention capacity and thus longer run time, coarse to fine media filtration had been suggested in direct filtration (Shea et al., 1971; Letterman et al., 1979). Multi media filters (e.g. Hutchison and Foley, 1974; Wagner and Hudson, 1982; Craig, 1985), graded up-flow filters (Odira et al., 1987) or two-stage filters (Janssen

and Bernhardt, 1992) could partially address the problem with their relatively higher deposit retention capacities at lower headloss in their larger grain compartment. For rapid sand filters some authors advocated to use higher filtration rates, larger grains and greater depths from economic considerations (AWWA, 1969; Helmut and Savage, 1974; Ives, 1974).

Direct filtration seems to be capable of handling some of the limitations of HRF in particular its low filtration rate, and difficulties in treating coloured water. It has limitations too (e.g. accepting low influent turbidity only, short filter run time), but, based on the underlying process mechanisms, the limitation of both technology could conceivably be addressed by adding coagulant to HRF i.e. converting HRF to direct horizontal-flow roughing filtration (DHRF). A counter-indication for the possible success of DHRF would be the increase in suspended mass, already high due to extra floc formation.

In the combined DHRF process the raw water is passed after rapid mixing onto the horizontal filter unit consisting of 2 to 4 compartments with coarse to fine grains. Fig. 1 shows a schematic representation of the DHRF process. The advantages and disadvantages of HRF and direct filtration, and the expected effect of their combined process DHRF is summarized in Table 1.

In this Chapter a first development of HRF to DHRF is described. The merits of both the processes are compared to test the hypothesis that DHRF combines the advantages of HRF and DF process.

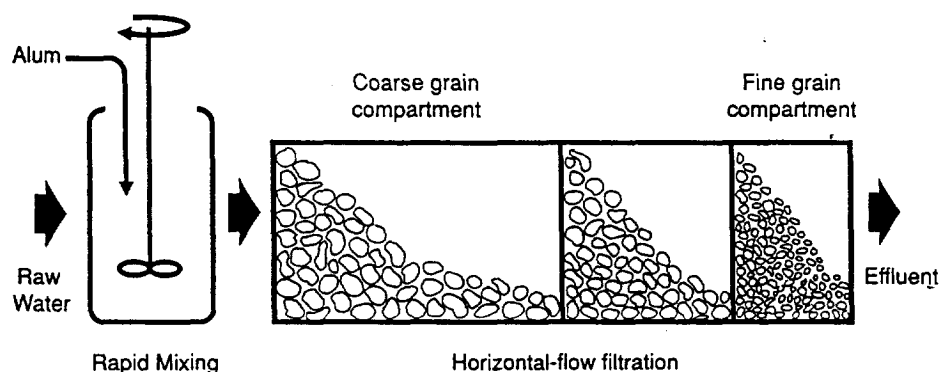


Fig. 1 Schematic diagram of direct horizontal-flow roughing filtration (DHRF) with 3 filter compartments.

MATERIALS AND METHODS

Experimental Set-up

Two parallel lab-scale pilot plants were operated. The schematic diagram of one set-up is

shown in Fig. 2. It consisted of a raw water preparation system, a coagulant dosing system and a roughing filter unit.

Table 1 Characteristics of horizontal-flow roughing filter (HRF) and direct filtration (DF), and their combination direct horizontal-flow roughing filter (DHRF).

HRF		Direct Filtration (DF)	
advantage	disadvantage	advantage	disadvantage
treats highly turbid water (50-200 NTU)	low filtration rate (0.5-1.5 m/h)	high removal efficiency (95-99%)	low water turbidity (5-25 NTU)
good removal efficiency (80-90%)	difficulties with stable colloids, colour, etc.	removes colour and colloidal matters	higher operating costs (chemical and backwashing)
long run time (weeks to months)	cumbersome filter cleaning	high filtration rate (10-20 m/h)	short run time (10-20 h)
simple technology			skilled operator required

DHRF	
expectation	investigation and development required
treats highly turbid water (> 200 NTU)	process mechanisms
good removal efficiency (90-99%)	optimum design
long filter run time (days to weeks)	removal efficiency of turbidity, colour, etc.
medium filtration rate (1-10 m/h)	filter cleaning, operational flexibility

Delft tap water ($\text{pH} \approx 8.1$, $\text{Ca}^{2+} \approx 50 \text{ mg/l}$, $\text{SO}_4^{2-} \approx 80$, $\text{HCO}_3^- \approx 130 \text{ mg/l}$, $\text{Al} \approx 10 \text{ } \mu\text{g/l}$, total hardness $\approx 8.9^\circ\text{DH}$) was fed continuously to the constant head water tank. From there water was passed through a flow meter (Brooks, tube size BM 75-1) into the kaolin mixing tank where it was mixed with kaolin slurry (kaolin fine powder from Reidel-de Haën, Germany) continuously added by a peristaltic pump (Watson Marlow, type 101 U/R). The required kaolin slurry was prepared every 2 days by mixing 50 g kaolin powder per litre tap water in a 100 litre polyethylene bucket. The stock slurry was kept under continuous and

vigorous mixing with two electronically controlled constant speed mechanical stirrers (Tamson, type RW20 DZM) in order to maintain a homogeneous suspension. The same type of stirrer was used in the kaolin mixing tank. The turbid suspension was passed to a plain sedimentation tank (1 h detention time) to eliminate coarse particles. Particle size analysis showed that after sedimentation particle sizes were in the range of 1-15 μm with mean 2 μm (by volume).

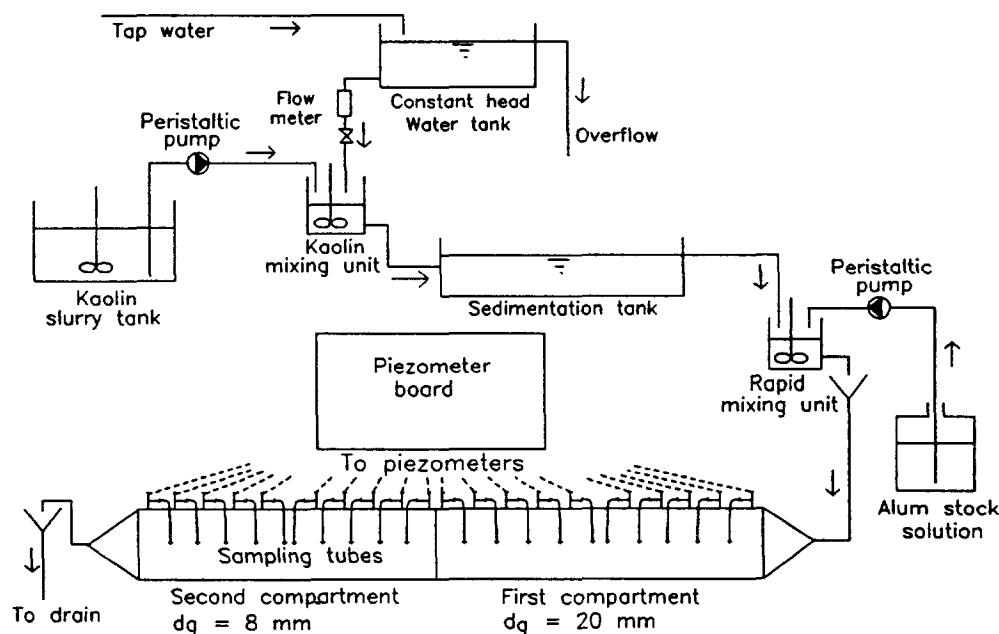


Fig. 2 Schematic diagram of the experimental set-up (DHRF pilot plant with two compartments).

Alum stock solution of 0.25% aluminium sulphate $\text{Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$ (3.75 mM $\text{Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$) analytical grade (Merck) was prepared daily. The strength of the solution was within the recommended limit of $< 0.5\%$ (Kawamura, 1976; Griffith, 1972). The required amount of alum solution was pumped to the rapid mixing unit and mixed by the same stirrer type (blade size 75 mm x 25 mm) with the kaolin suspension. The desired velocity gradient G in the rapid mixing unit could be obtained by adjusting the speed of the stirrer. The coagulated suspension was then passed on to the roughing filter unit. When coagulation was not applied (i.e. filter was operated in HRF mode) the rapid mixing unit was bypassed.

The container of the filter bed was made of perspex tube, each 20 cm in diameter and 1 or 2 m long except when otherwise mentioned. Each tube unit was filled with grains of specified

size and could be connected in series to obtain desired lengths of bed compartments. They could be placed in horizontal or in vertical alignment. Filter grains were packed and compacted carefully so that no setting of the grains occurred during filter operation. The required flow rate ($\pm 2\%$) could be adjusted by regulating the valve after the flow meter (Fig. 3) and remained constant throughout a filter run. The water fell freely in a narrow vertical tube (2 cm diameter) at the filter inlet, the raising head in the tube compensated the increase in headloss (~ 20 cm) in the filter; the change of suspension residence time during a filter run due to raising head is negligible. Funnels and perforated plates were fitted at the inlet and outlet of the filter to provide uniform flow distribution over the filter tube cross-section. Air vent pipes were placed at each 2 m to release any air entrapped inside the filter tubes. Perforated air pipes were placed along the bottom of the filter. After each run, high water flow and simultaneous vigorous air scour were provided to clean the filter. The wall effects were presumed to be minimal (see Chapter 3).

Water samples were collected from the filter bed at predetermined intervals at the inlet and outlet, and also from sampling ports at each 20 cm of bed length. To obtain representative samples the velocity in the entrance of the sample tubes during sampling was adjusted to be comparable to the in-pore velocity in the filter bed (Ives, 1966; Fox, 1966). Sample tubes were drilled out of 10 mm perspex rod with entrance diameter of 5 mm over 10 mm length and then narrowed to 2 mm throughout the rest of the rod. The sampling tubes were placed vertically (3 cm inside the grains), and the higher flow velocity in the narrow sections prevented any deposition within it. The sampling tubes were gently flushed before sampling and only 3 to 4 at a time so that no substantial reduction in filtration rate occurred. Piezometric levels at the corresponding sampling sections were measured through another set of ports connected to the manometer board.

Turbidity (Dr. Lange Trübungsphotometer LTP 4) was the main indicator of the removal performance and was expressed in NTU (Nephelometric Turbidity Unit). Fig. 3 shows the relationship of turbidity with suspended solids. For uncoagulated raw water the numerical values of turbidity and suspended are equal; similar relationship was found in most natural waters (Wegelin, 1986; IWACO, 1991). In case of coagulated suspension and samples drawn along the filter length a linear relationship up to about 170 NTU is observed; the ratio of turbidity to suspended solids was, however, higher. The effluent breakthrough was ascertained on the basis of the observation of a sharp increase in effluent turbidity. Particle size analysis (HIAC Model PC-320) was done by sensor type CBM-60 and CBM-150 which could measure the sizes down to 1 and $2.5\ \mu\text{m}$, respectively. In order to assess the proportion of kaolin particles less than $1\ \mu\text{m}$ particle size measurement was also performed with Sedigraph (model 5100, Micromeritics) which determines particle sizes on the basis of settling rates.

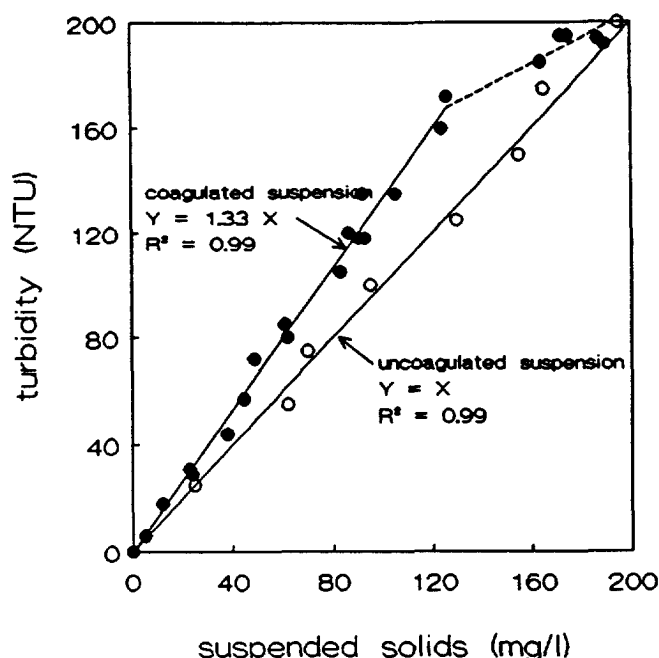


Fig. 3 Relationship of turbidity and suspended solids for (i) raw water with uncoagulated kaolin suspension, and (ii) samples drawn from different DHRF filter lengths and when raw water (200 NTU) is coagulated with 1 mg Al(III)/l.

Electrophoretic mobility (EM) of particles was measured by using a zeta potential meter (Tom Lindstrom AB) under 90 volt potential difference and by averaging 5 measurements in each direction by switching the signs of the electrode. The aluminum ion concentration was measured using graphite furnace atomic absorption spectrophotometry (Perkin-Elmer Model 1100 B). Samples were filtered through 0.45 μm membrane before analysis to remove suspended solids.

RESULTS AND DISCUSSION

Kaolin suspension

Fig 4a show the particle volume concentration of river waters and of a kaolin suspension: the Ruaha and Ruvu are Tanzanian rivers, the Sihl a Swiss river (Boller, 1993). The Rhine sample was taken from Basle, Switzerland (Filella et al., 1993), and the Danube samples from Vienna, Austria (Kralik and Augustin-Gyurits, 1991). Different lower and upper limits for the particle size determination were applied: Boller used a lower limit of 1 μm , Filella a lower limit 0.2 and upper limit of 5 μm , and Kralik and Augustin-Gyurits used a lower limit of 0.3 μm . The lower limit applied with the case of kaolin suspension (kaolin fine

powder from Reidel-de Haën, Germany) under this study was $1\ \mu\text{m}$. Although each river suspension has its own characteristics and concomitant behaviour in treatment, the graph suggests that kaolin suspension overall is well representative of many natural rivers on the count of particle size distribution.

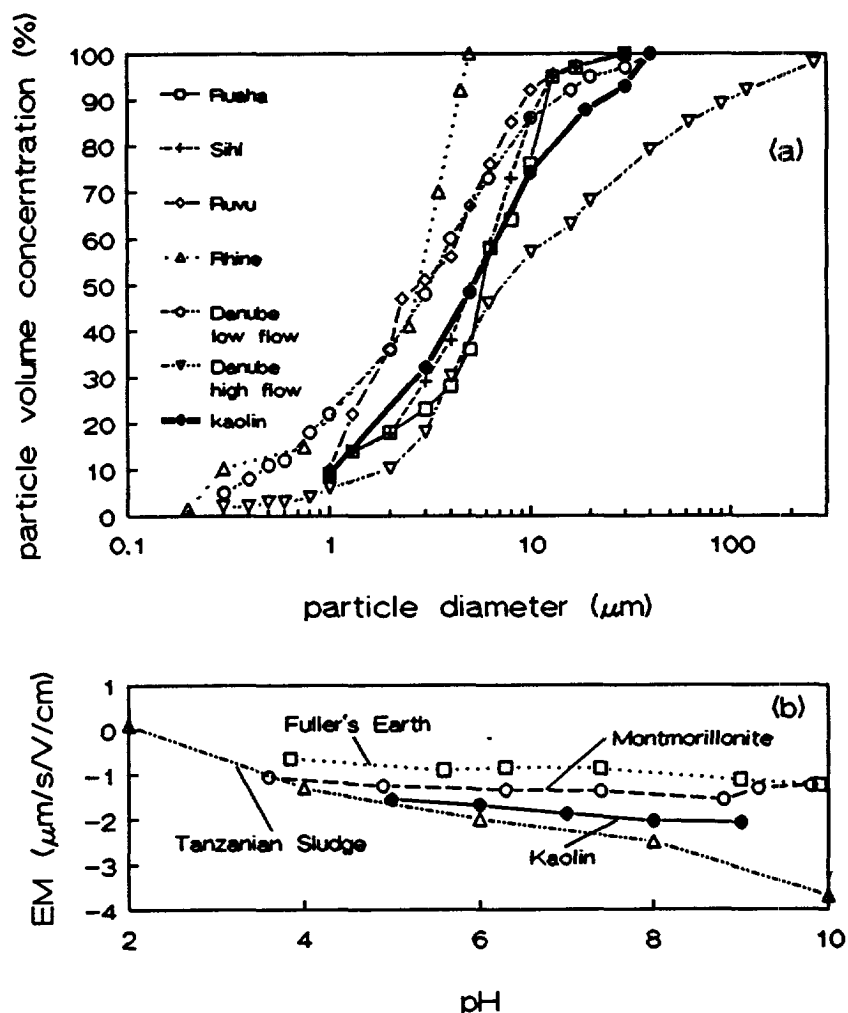


Fig. 4 Comparing kaolin used in this study with naturally occurring particles in terms of (a) Particle size distribution (Kralik and Augustin-Gyurits, 1991; Boller, 1993; Filella et al., 1993) and (b) electrophoretic mobility (Black and Hannah, 1961; Boller, 1993).

Fig. 4b shows the electrophoretic mobility (EM) against pH of Tanzanian sludge taken from a drinking water reservoir (Boller, 1993), two clay types (Black and Hannah, 1961), and of the kaolin suspension of this study. Kaolin appears also representative of the particle surface charges. When evaluating the performance of a process, it is desirable to use reproducible and well described model suspension so that the temporal variability inherent to natural

sources can be avoided. As kaolin suspensions are considered representative of the colloid-chemical systems of many rivers, they are well established as a model suspension in research (McCooke and West, 1978). Packham (1962) found that particle removal from kaolin suspension after coagulation with aluminium sulphate at different pH is in good agreement with the coagulation of many British rivers. However, the model suspension do not reflect the variation in physio-chemical characteristics and the effect of colour and organic matters.

HRF and DHRF performance

To assess the respective performance of HRF and DHRF a 2 m long filter with two compartments (first compartment length $L = 1$ m and grain size $d_g = 20$ mm, and second compartment $L = 1$ m and $d_g = 4$ or 8 mm) was used. The influent turbidity was 160 and 510 NTU and filtration rate 3 and 5 m/h. In DHRF mode a constant dose of 1 mg Al(III)/l was added in the rapid mixing unit with velocity gradient $G = 200 \text{ s}^{-1}$ and mixing time $t = 4$ min. The filters in both modes were operated till terminal headloss of 40 cm. The process conditions and the results of the filter runs are summarized in Table 2. It can be observed that DHRF systematically yielded higher removal efficiencies and longer filter run time. For these experimental conditions improved performance on removal efficiency was 20-50% higher. The difference in the efficiencies were more pronounced for higher filtration rates and for larger grain sizes in the second compartment. The filter run time is about double for all DHRF runs as compared to HRF runs under similar conditions.

Table 2 also shows that the rate of headloss development in DHRF is consistently lower, about half of HRF, even though the addition of coagulants creates additional voluminous hydroxides in the suspension. The lower headloss is possibly due to:

- (i) The enhanced removal of particles in the first compartment with coarse grains. In DHRF mode about half the total suspended solids (including alum hydroxides) was retained in the first compartment whereas in HRF mode only about one fourth is retained. Higher headloss in the first compartment of DHRF is thus observed. Hence, in HRF about twice the amount of suspended solids is to be removed in the next finer grain compartment. The finer grain compartment is more sensitive to headloss for the same amount particles deposited than the first compartment with coarse grains.
- (ii) Use of coagulants in DHRF may have created deposits which are easier to be dislocated and to drift under gravity to the bottom of the filter (see below). This leaves cleaner upper pores in the upper part of the bed and reduces net hydraulic resistance to flow.
- (iii) In case of DHRF the use of coagulants and in-pore flocculation may have produced flocs which are larger and have surface characteristics less prone to headloss. It is reported that

in direct filtration the headloss development is less when a flocculation unit precedes filtration and produces larger flocs compared to when there is no flocculation unit (Edzwald et al., 1987). Headloss development is also reported to be less when the influent particle sizes are larger (Tobiason, 1993; Veerpaneni, 1993).

Table 2 Results of HRF and DHRF performance. The total filter length was 2m consisting of two equally sized compartments. Grain size in the 1st compartment = 20 mm and in 2nd compartment = 4 or 8 mm. Filter run operated till terminal total headloss of 40 cm. Coagulant dose in DHRF = 1 mg Al(III)/l.

process conditions				Filtration results			
mode	influent turbidity	grain size in 2nd compt.	filtration rate	1st compartment		total filter	
				removal efficiency	headloss	removal efficiency	run time
	(NTU)	(mm)	(m/h)	(%)	(cm)	(%)	(h)
HRF	160	4	3	33	3	76	26
DHRF	160	4	3	54	7	97	51
HRF	160	8	3	24	2	51	15
DHRF	160	8	3	58	5	93	32
HRF	160	4	5	22	3	52	13
DHRF	160	4	5	44	16	96	26
HRF	510	4	5	22	7	37	15
DHRF	510	4	5	45	19	66	27
HRF	510	8	5	16	3	33	15
DHRF	510	8	5	56	5	80	21

In a second series of experiments HRF and DHRF removal performance was assessed with 8 m long filter, composed of two equally sized compartments. The grain size of the first compartment was 20 mm, and of the second compartment 8 mm. The initial turbidity was 200 NTU and filtration rate was 5 m/h. In DHRF mode a constant dose of 1 mg Al(III)/l was added in the rapid mixing unit with $G = 200 \text{ s}^{-1}$ and $t = 1 \text{ min}$.

Fig. 5 shows the effluent turbidity as function of filtration time. A third HRF run with $v = 1 \text{ m/h}$ (typical HRF filtration rate) is also compared in the same figure. The much lower

effluent turbidity that can be reached by DHRF was again demonstrated. The average effluent turbidity of DHRF was about 2 NTU and breakthrough period 95 h. During the operation the average effluent turbidity for HRF with $v = 5$ m/h was 56 NTU and with $v = 1$ m/h was 22 NTU. The breakthrough did not occur in HRF during the investigated time.

Particle Deposition in DHRF

The transparent perspex filter tubes allowed visual observation of particle and floc deposition. The horizontal movement of flocs with water flow was combined with a gravitational downward drift. The solids settled on the top of the grains in the shape of heaps of sometimes several mm height. As the deposits exceeded their slope stability, small lumps of settled matter drifted downward and resettled at the bottom of the filter. As a result there was a gradual building up of deposits upward from the filter bottom. This mode of particle deposition was more pronounced in the first compartment with larger grains. The deposits of DHRF also appeared to drift more easily downward than those of HRF. Similar drifting of deposits was also reported in case of the HRF till smallest grain size of 4 mm (Wegelin, 1987). Fig. 6 schematizes the mechanism described. This drifting process is advantageous as the flow-through capacity and hence the removal capacity of the upper layers is restored to a certain extent.

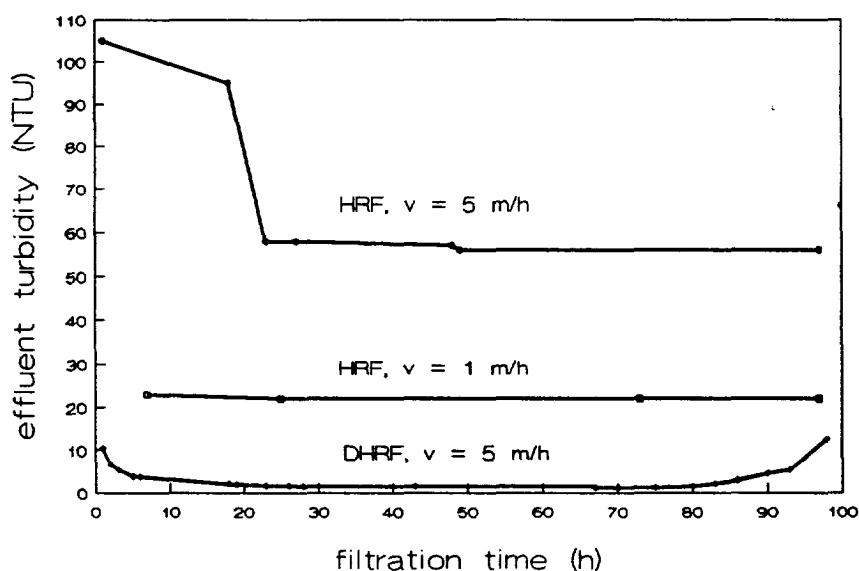


Fig. 5 Comparing the effluent turbidities of HRF and DHRF. Total filter length = 8 m. 1st compartment length = 4 m, grain size = 20 mm; 2nd compartment length = 4 m, grain size = 8 mm; initial turbidity = 200 NTU, coagulant dose in DHRF = 1 mg Al(III)/l.

Figs. 7a-d illustrate the variation of residual turbidity along the filter length for different time intervals and for flow rate 1-7 m/h. The filter consisted of a 2 m long first compartment (20 mm grains) and 2 m second compartment (8 mm grains). The initial turbidity was 200 NTU and coagulant dose 1 mg Al(III)/l. The residual turbidity profiles along the length of the coarse compartment showed a general pattern: first it decreased a slower rate and then at a higher rate and again with gradually slower rate. This shape of the profile resembles the typical Kynch's curve for settling pattern of flocculant particles i.e. inverted S shaped (Degremont, 1991), and also similar to the observed general pattern of turbidity removal of flocculant particles in a rectangular sedimentation tank (Akers, 1975; Masschelein, 1992). As the filtration time increases the filter pores especially in the first compartment are gradually filled up with deposits and the particle removal capacity decreases. At the end of the run the coarse grain compartment loses its removal capacity and all turbidity is passed onto the second compartment.

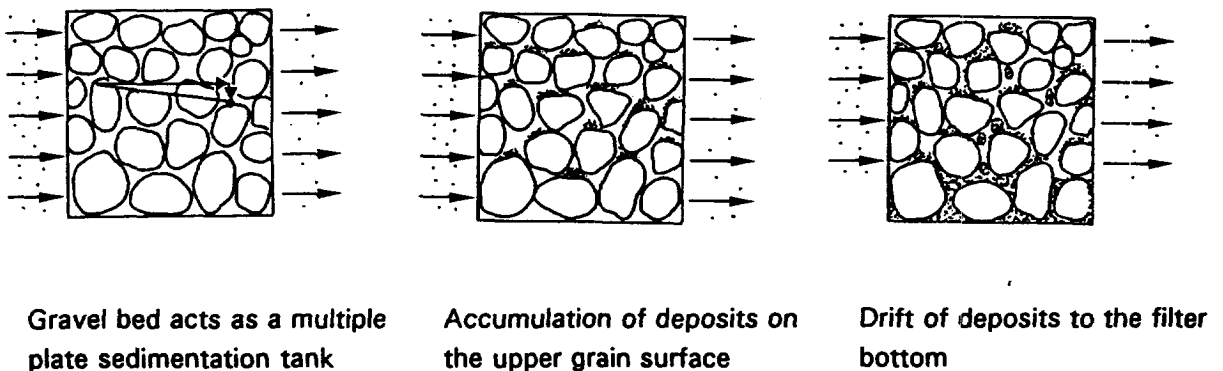


Fig. 6 Schematic diagram showing settled deposit drifting in DHRF (Wegelin, 1986).

The shape of the profiles of the second compartment resemble the typical exponential residual turbidity profile in a rapid sand filter (e.g. Ives, 1960; 1975) stressing that filtration is the dominant process. The present study also showed that when breakthrough occurred in the first compartment the effluent turbidity in the second compartment deteriorated rapidly. It was also observed that the second compartment can accommodate the bulk of incoming turbidity for a limited period when breakthrough occurs in the first compartment.

Filtration rate

Average effluent turbidity increases and filter run time decreases with increasing filtration

rate (Fig. 8). These experiments were conducted on the comparatively short filter length (4 m) in order to allow for rapid assessment of the relative performance; in an optimum design the DHRF is expected to be 7-8 m long and as such, the effluent turbidity will be lower and run time longer. The maximum effluent turbidity allowed in set of experiments was 20 NTU; relatively higher limit because of the shorter filter length. The results indicate that higher filtration rate up to 7 m/h may be feasible in DHRF.

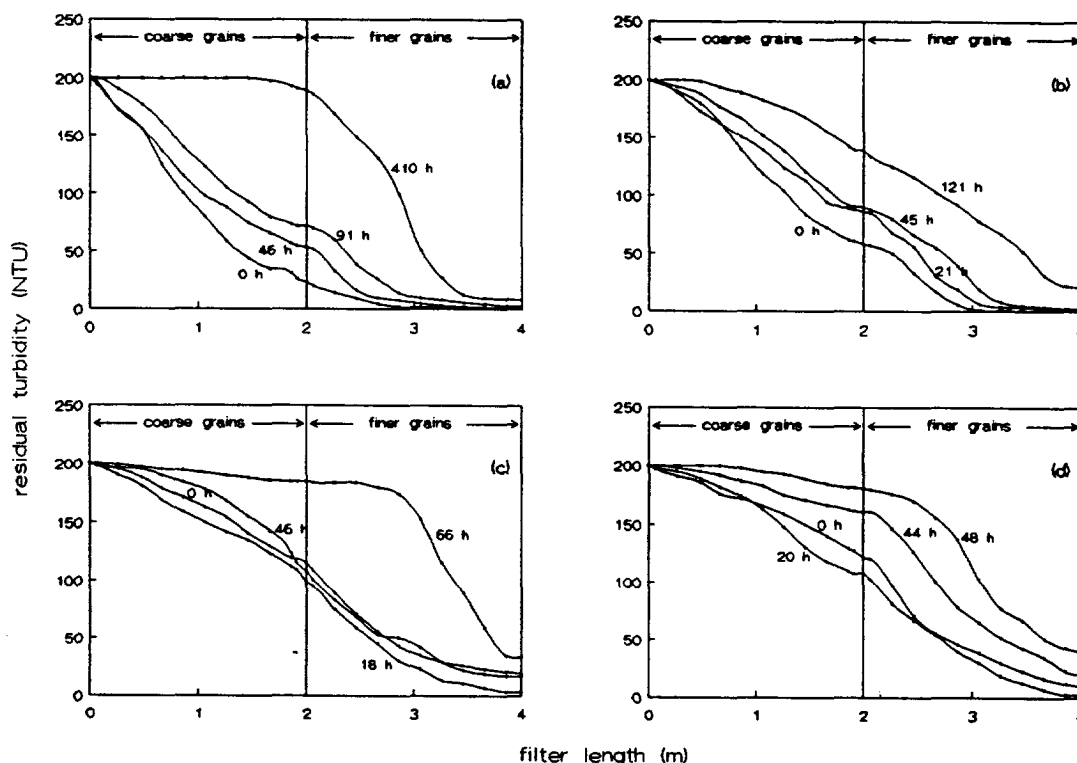


Fig. 7 Residual turbidity profiles for different DHRF filtration rates along the filter length. Total filter length = 4 m; initial turbidity = 200 NTU; coagulant dose = 1 mg Al(III)/l. (a) Filtration rate $v = 1$ m/h, (b) $v = 3$ m/h, (c) $v = 5$ m/h and (d) $v = 7$ m/h.

Comparing vertical and horizontal flow filters

The height of a vertical filter bed is generally limited to 1.0-2.0 m as higher height increases the construction cost of the walls and foundation. If higher bed height is required then a number of filter units in series can be constructed. Each such unit has to be provided with individual piping, supernatant water height, filter controls, etc. The length of a horizontal filter, on the other hand, is not limited by the structural requirements. Unrestricted filter length with a number of compartments can be built economically in a single structural unit. Filter cleaning by backwashing may not be feasible in horizontal or vertical roughing filters

as it would require exceedingly high velocities for bed expansion. Roughing filters are normally cleaned by downward hydraulic flushing (see Chapter 7). A vertical roughing filter should then be preferred in up-flow mode as most of the deposits will be retained in the lower bed section near the inlet and their removal will be easier by downward flushing.

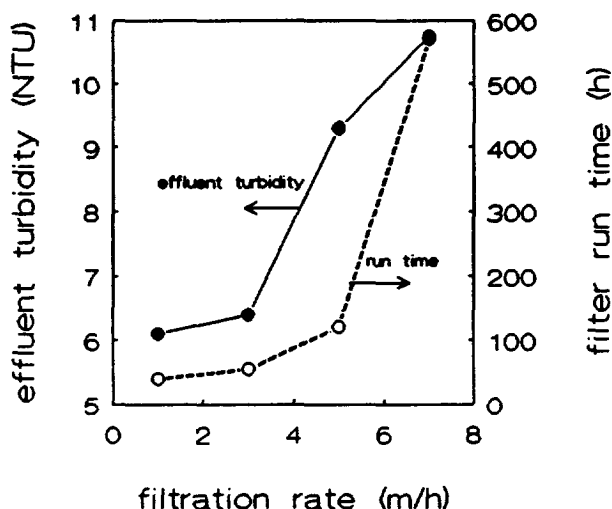


Fig. 8 Variation of effluent turbidity and filter run time with filtration rate (process conditions: see Fig. 7).

To compare roughing filter performance in vertical up-flow and in horizontal flow direction 3.5 m long filters consisting of a 1.75 m first compartment (grain size = 20 mm) and a 1.75 m long second compartment (grain size = 8 mm) were operated. Both filters were operated in conventional and in direct filtration mode. When no coagulation was used filter runs were conducted at filtration rate of 3 m/h. Initial turbidity (200 NTU) and other process conditions were similar. The results for the conventional filter mode are summarized in Table 3. Better filter performance in terms of effluent turbidity (by 35%) and run time (by 35%) is observed with the horizontal flow.

In direct filtration mode the filter performance in the two different flow directions was tested for filtration rates of 1.5-7.0 m/h. Fig. 9 compares their residual turbidities and their run times. The horizontal flow also appears to have higher removal efficiencies (by 5-30%), but however, only marginally longer run time. At low filtration rate of 1.5 m/h the difference in their performance is small (5%) but it becomes appreciable (10-30%) at progressively higher filtration rates. The higher particle removal efficiencies in horizontal flow filters agrees with the theoretical predictions. In a three-dimensional computer simulation of particle capture within filter pores Burganos et al. (1991, 1993) showed that particle removal efficiency is higher in horizontal filters than in vertical ones. They also showed that the difference in removal efficiency increases for higher filtration rates.

Table 3 Horizontal and vertical roughing filter performance in conventional filtration mode. In both the cases total filter length = 3.50 m. 1st compartment length = 1.75 m, grain size = 20 mm; 2nd compartment length = 1.75 m, grain size = 8 mm; initial turbidity = 200 NTU; filtration rate = 3 m/h.

flow direction	effluent turbidity (NTU)	filter run time (h)
horizontal	90	70
vertical (up-flow)	125	50

Fig. 10a and 10b show the headloss profiles for different run time for the vertical and horizontal filter, respectively. The filtration rate was 3 m/h. In the horizontal filter a gradual headloss build up over both the filter compartments is observed indicating an even distribution of deposits. In contrast, the vertical filters have higher headloss in the initial sections. About 60% of the total headloss occur of the first compartment in its first 5 cm and 60% of the total headloss in the second compartment in its first 25 cm. This indicates the high deposit accumulation in the initial sections. The total filter headloss in the vertical filter was consistently higher (80-100%) than in the horizontal filter. For example, in case of 3 m/h filtration rate and at same specific deposit of 15 g/l, the total headloss in the vertical filter is 83 cm compared to 48 cm in the horizontal filter, again with major flow resistance occurring in the initial sections.

The lower removal efficiency and higher headloss in vertical flow roughing filters could also be attributed to the location of deposit accumulation within the filter. In an horizontal filter (DHRF) the settled deposits drift downward and are distributed in the bottom layers. There the deposits remain undisturbed, except for the upper deposit surface, by the water flow. The upper part of the filter remains relatively clean with lower net hydraulic resistance, and the bulk of the flow will pass through at this layer. The particle removal still occurs along the total filter length, however, at a reduced efficiency due to higher interstitial velocity (i.e. similar to higher filtration rate).

In a vertical filter the settled deposits drift downward in both the compartments. As a result the deposits accumulate within the first few cm from the inlet. This was visually clearly observed in the first compartment with larger grains. Unlike the horizontal filter the total flow has to pass through the clogged section. This condition is similar to when the ultimate specific deposit (removal efficiency = 0) is reached in a vertical rapid sand filter. The flow

is concentrated in the preferential tubular passages (wormhole) (Ives, 1987; Baumann and Ives, 1991). An increasing fraction of total flow experiences the higher interstitial flow velocities in the wormhole passages. The high interstitial velocity increases the headloss and no effective particle removal occurs (Ives and Clough, 1985; Ginn et al., 1992). With increasing filtration time the height of the clogged section increases reducing the effective filter length and particle removal capacity. A similar progressive clogging front was observed in direct filtration (Adin and Rebhun, 1974). Preferential flow through the tubular passages in the deposits was visually observed near the filter walls. At higher filtration rates this phenomenon was accelerated and further reduced particle removal. The described different modes of deposit accumulation in horizontal and vertical roughing filters is schematized in Fig. 11.

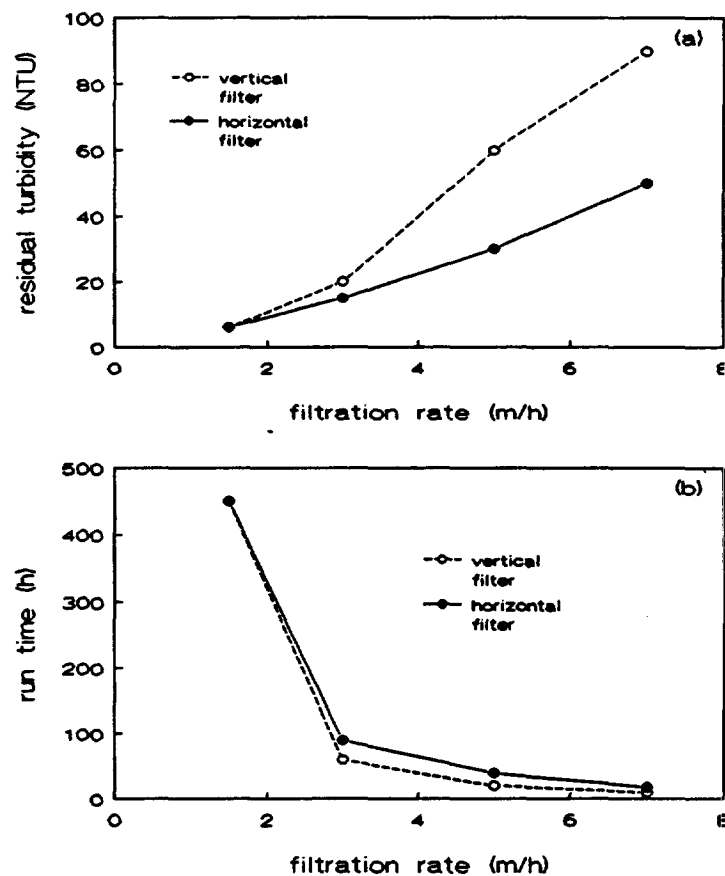


Fig. 9 (a) Effluent turbidity and (b) filter run time in horizontal and in vertical filters. Total filter length = 3.50 m. 1st compartment length = 1.75 m, grain size = 20 mm; 2nd compartment length = 1.75 m, grain size = 8 mm; initial turbidity = 200 NTU; coagulant dose = 1 mg Al(III)/l.

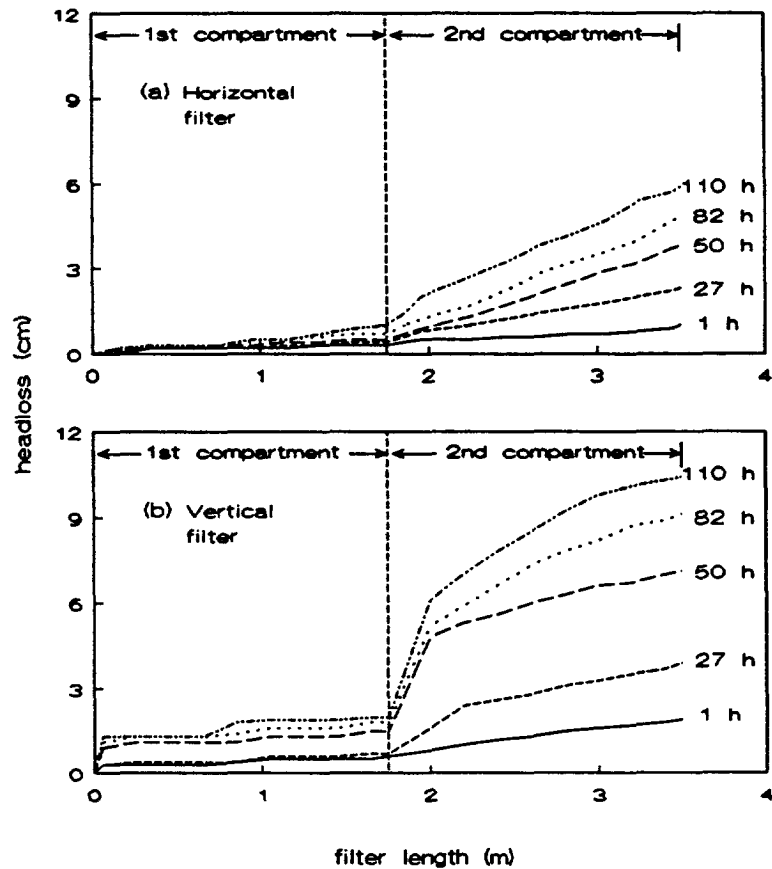


Fig. 10 Headloss profile along the filter length at different run time. Total filter length = 3.50 m. 1st compartment length = 1.75 m, grain size = 20 mm; 2nd compartment length = 1.75 m, grain size = 8 mm; initial turbidity = 200 NTU; coagulant dose = 1 mg Al(III)/l; filtration rate = 3 m/h. (a) Vertical up-flow filter, and (b) horizontal filter.

Parameter evolution along DHRF

Evolution of residual turbidity and headloss along a typical DHRF run is studied. Fig. 12a demonstrates that the residual turbidity profiles along the filter length remained stable (increase < 5 %) during most of the working period (4 days).

The filtration coefficient is mainly function of specific surface area (of grains and deposits) and of interstitial velocity causing shear stress on the deposits (Ives, 1975b). The high deposit storage capacity of the larger grains in the roughing filters at the expense of low headloss (causing low shear stress) defers the filter breakthrough and thus ensures stable effluent turbidity for a long operating time. The shape of the residual turbidity profiles again suggests that the dominant particle removal mechanism in the first compartment with coarse grains

is sedimentation and in the second compartment with relatively finer grains is deep bed filtration.

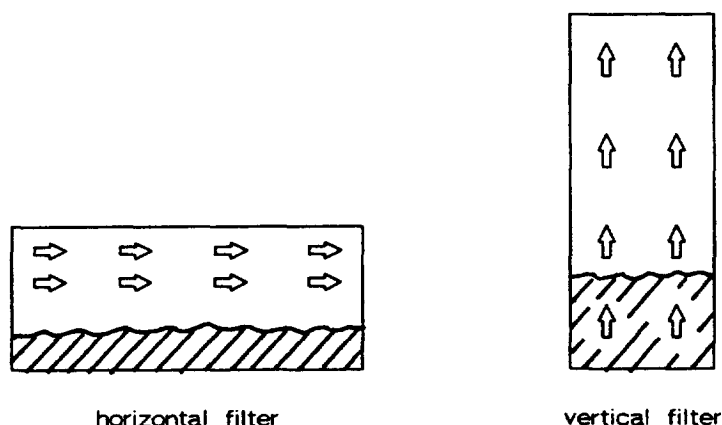


Fig. 11 Schematic representation of the mode of deposit accumulation in (a) horizontal flow and in (b) vertical up-flow roughing filters.

The headloss over the first DHRF compartment at the end of 95 h run time is only 3.5 cm (Fig. 12b) and specific deposit about 21 g/l. The total headloss in the second compartment is 13.7 cm with less specific deposit of about 7 g/l due to the higher sensitivity of deposit development in smaller grain. With increasing filtration time the rate of headloss development is higher in filter sections near to the inlet of the second compartment. This is due to the higher deposition rate there. The total DHRF filter headloss (17.2 cm) is lower than to the allowable total headloss of 20-30 cm in HRF (Wegelin, 1986). The insert in Fig. 12 show the total headloss development with run time. The linear relationship indicates the absence of any cake formation in the inlet sections and the deep penetration of deposits (Ives, 1975b).

The particle size distributions (by particle volume) of the raw water suspension and after 1 day filtration at various stages in the treatment line are presented in Fig. 13a and 13b. Coagulation in the rapid mixing unit increased the median particle diameter of raw water from about 5 to 12 μm (Fig. 13a). The median particle diameter gradually increased from 12 at the inlet to 16 μm at the filter effluent (Fig. 13b). Along the filter length the peak of the distribution curve is also gradually flattened and standard deviation of the particle size increased. The increasing particle size along the filter length suggests that flocculation (in-pore) is still continued even at the 8 m filter length. The increase in particle size is more pronounced in the second DHRF compartment than in the first, possibly because of the higher velocity gradient in the second compartment. The larger particle sizes will be easier to remove in the subsequent sand filters.

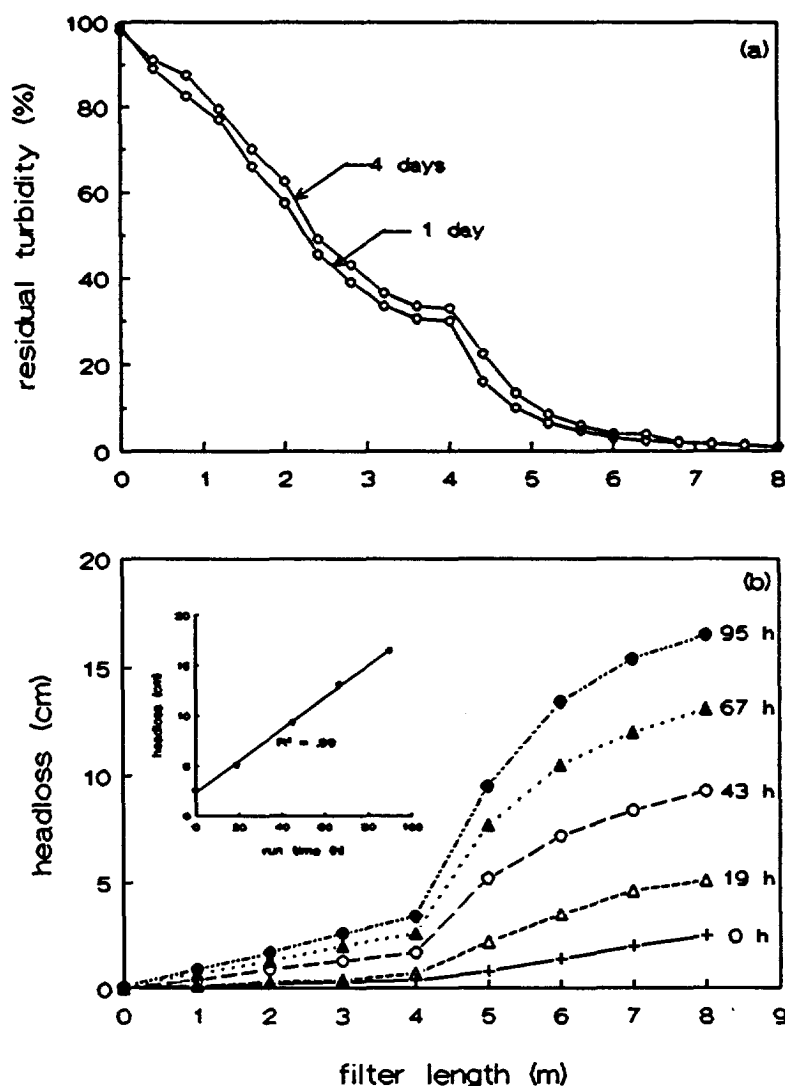


Fig. 12 (a) Residual turbidity and (b) headloss profiles at different times along the filter lengths. Initial turbidity = 200 NTU, coagulant dose = 1 mg Al(III)/l, filtration rate = 5 m/h, total filter length = 8 m consisting of first compartment (4 m) with 20 mm and the second compartment (4 m) with 8 mm grain sizes.

In-pore flocculation and its effect on filtration efficiency are a subject of controversy: some authors emphasized its importance (Shea et al., 1971; Cleasby, 1976; Treanor, 1976; Culp, 1977) while others argued for its limited role (Fitzpatrick and Spielman, 1973; Ghosh et al., 1975; Vreeken et al., 1978; Yeh and Ghosh, 1981). Ives and Al Dibouni (1979) abandoned experimental attempts to measure in-pore flocculation rates because of excessive particle retention in the filter. They concluded that no experimental techniques could be devised to separate the effect of filtration and flocculation. More recent experimental and computer

modelling work by Graham (1986, 1988) again stressed the importance of in-pore flocculation.

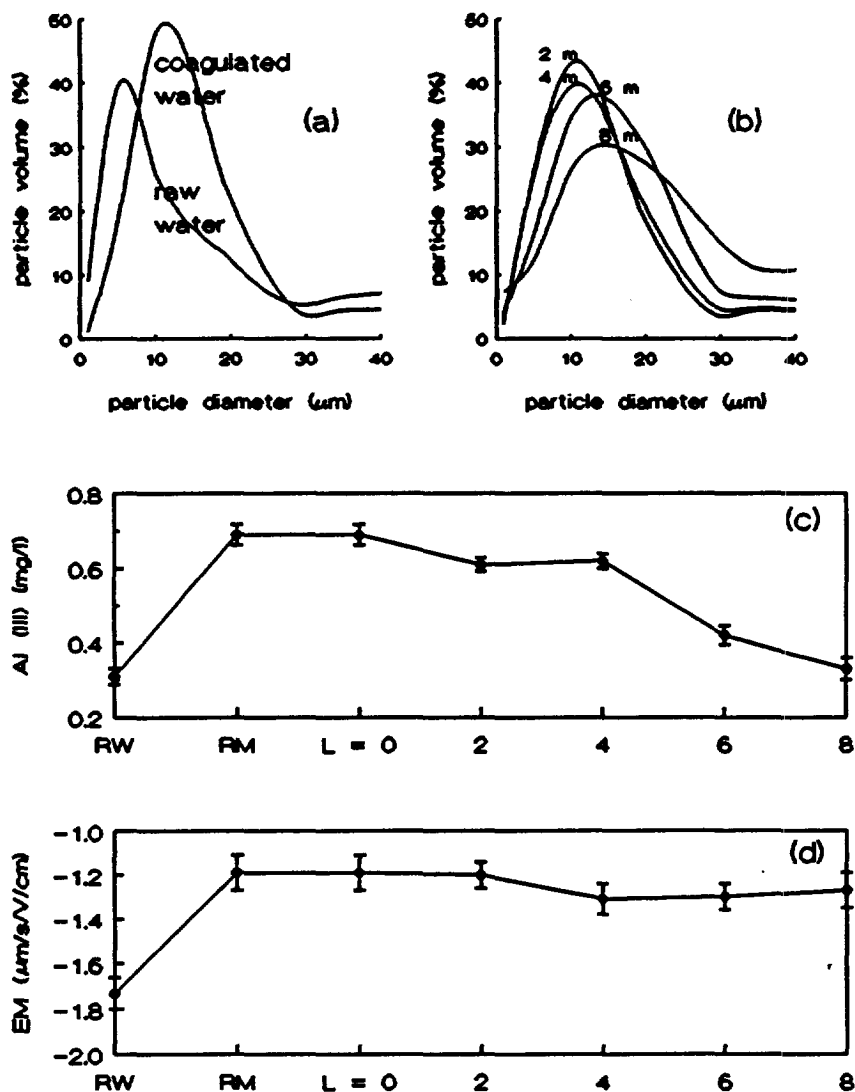


Fig. 13 Evolution of parameters along the process of DHRF. (a) particle size distribution of raw and coagulated water, (b) particle size distribution along the filter length, (c) Al concentration, and (d) EM.

In the DHRF filter two competing processes occur simultaneously that may effect particle size distribution. Particles, preferentially the larger and heavier ones, are removed along the filter length by sedimentation or filtration resulting in a higher proportion of smaller particles in the liquid. On the other hand, the remaining particles are subject to further flocculation and are being agglomerated into bigger aggregates. The observed increase in median particle

size may demonstrate the significant contribution of continuing in-pore flocculation. On the other hand it was argued that the particle detachment in rapid sand filters, are due to increased interstitial velocity, can also contribute to the increased particle sizes (Mintz, 1966). However, detachment is probably low during the initial filtration time when the headloss and shear stress on deposits are low and the observed increase in particle size is mainly contributed by flocculation.

The change of residual Al concentration is shown in Fig. 13c. The residual Al concentration includes (i) any dissolved Al species, and (ii) the Al species adsorbed onto kaoline particles and (iii) Al content in kaolin particle itself and which had passed through the $0.45\ \mu\text{m}$ filter paper. Sedigraph analysis of raw water revealed that about 9% of particle mass has particle diameter less than $0.5\ \mu\text{m}$. The Al concentration of raw water was $0.29\ \text{mg Al(III)/l}$. Addition of $1\ \text{mg Al(III)/l}$ in the rapid mixing unit increased the residual Al to $0.69\ \text{mg Al(III)/l}$, the balance being due to the Al precipitates (micro-flocs) and Al adsorbed onto kaolin particle retained in the filter paper. The dissolved Al is not expected to remain in solution in the filter as the adsorption of monomer and polymer species onto mineral particles and the formation of hydroxide precipitates are effectively complete within 1-7 s (Amirtharajah, 1987).

Along the filter residual Al gradually reduced to $0.33\ \text{mg/l}$. This is, however, above the WHO guideline of $0.2\ \text{mg/l}$. It is expected that most of the remaining aluminium (adsorbed onto and contend in the kaoline particles) and microflocs $< 0.45\ \mu\text{m}$ can be removed in the subsequent slow or rapid sand filter which are more effective in removing small particles than roughing filters. The gradual removal of Al especially in the second compartment with relatively smaller grains also suggests the progressive removal of smaller particles together with the adsorbed Al species. Thus sub-micron particles are also removed in roughing filtration possibly by diffusion.

The particle removal depends strongly on the zeta potential (which is proportional to the EM) of the suspended particles (Black and Hannah, 1961; Hall, 1965; Johnson and Amirtharajah, 1983; Dentel and Gossett, 1988; Chang and Vigneswaran, 1990). Addition of alum in the rapid mixing unit caused the EM of the raw water to decline from -1.7 to $-1.3\ \mu\text{m/s/V/cm}$ (Fig. 6e). The EM, however, further remained almost unchanged along the filter length indicating that the particles were already maximally destabilized in the rapid mixing unit. EM of the particles in DHRF effluent was lower than that of particles from pretreatment processes where no coagulants are used. The EM of effluent particles in e.g. HRF would be approximately equal to that of raw water EM ($-1.7\ \mu\text{m/s/V/cm}$).

The use of coagulant in the DHRF resulted in lower effluent turbidity (2-3 NTU) as well as higher particle size (median size $\sim 16\ \mu\text{m}$) and lower (25 %) surface charge. Boller (1993)

found that the particle with diameter $> 10 \mu\text{m}$ are effectively removed in HRF whereas, about 10% and 15% of initial concentration of particle diameters of 3.6 and $1.1 \mu\text{m}$, respectively, remained in the effluent. These characteristics of DHRF effluent would favour the particle separation in the following slow or rapid sand filters as compared to the effluent of other pretreatment where no coagulant is used (such as HRF).

Feasibility of DHRF

HRF is mainly developed for small communities in rural areas. Filtration rate is low (0.5-1.5 m/h) large surface area (about same as slow sand filter) and construction volume is required. It also has difficulties with very high raw water turbidity ($> 200 \text{ NTU}$) or when colour or colloidal matters are present. Higher filtration rate (3-7 m/h) can be applied in DHRF, yet yielding much higher removal efficiencies. The required surface area and construction volume in DHRF would therefore be about 4-6 times lower than and thus can be feasible for urban application. The results also suggest that DHRF can treat higher raw water turbidity ($> 200 \text{ NTU}$). The operating cost will be a little higher in HRF as about 1 mg Al(III)/l coagulant dose is required; of course it is much lower than the flocculation-sedimentation process which requires about double chemicals and stricter process control. The DHRF can be constructed with locally available materials and manpower in developing countries. Operation and maintenance are also expected to be comparatively simple.

The results so far suggest that coagulation in HRF has been an effective process modification in terms of better particle removal efficiency and higher filtration rates. Optimum coagulant dose and mixing conditions, filter configuration and flexibility in process control are required to be investigated. DHRF performance with coloured and other different water type are also to be determined. Another important criterion for the successful DHRF operation will be its deposit removal. Preliminary evidence shows that DHRF sludge can be more easily mobilized during hydraulic flushing than not coagulated sludge.

The feasibility of DHRF prior to slow sand filters would be restricted to rural or urban areas where coagulants are regularly available. The existing HRFs can easily be modified to DHRFs by constructing a rapid mixing unit before the HRF. This will increase the discharge of HRFs by applying higher filtration rate, and/or improve the quality of effluent to slow sand filters. The performance of DHRF also reveals that it may have a good potential to substitute the convention flocculation-sedimentation unit and applied before rapid sand filters. This expands the scope of DHRF application in urbanized areas and possibly also in industrialized countries.

CONCLUSIONS

The HRF process has been modified to DHRF by combining it with direct filtration. Some of the limitations of HRF like lower filtration rate could be overcome. DHRF yielded good effluent quality (~ 2 NTU) and can be applied at a higher filtration rate (3-7 m/h) than HRF (0.5-1 m/h). Further investigation on optimizing DHRF design and its capability to deal with coloured and other different water quality, filter cleaning, etc. are required to be investigated.

The first compartment of DHRF with coarse grains acted as a gross particle collector in which sedimentation appears to be the main process. The second compartment with finer grains can be considered to act similar to deep-bed filtration.

Roughing filtration in horizontal flow direction systematically produces higher removal efficiencies, both in conventional filtration and in direct filtration mode, than vertical up-flow. The difference in their efficiency is small (5%) in low filtration rate of 1.5 m/h but appreciable (10-30%) at 3-7 m/h. The headloss in horizontal direction is about half than vertical direction.

The large grain sizes in DHRF promoted even deposit distribution and stable residual turbidity profiles over the filter length during the working period, and linear headloss development. Residual Al concentration gradually decreased over the filter length. Significant contribution of in-pore flocculation, especially in the second compartment of DHRF was observed.

DHRF effluent had favourable characteristics such as low turbidity, larger particle sizes, lower surface charge than effluent from pretreatment options where no coagulants are used. Thus the still remaining particles in DHRF effluent can be removed more easily and effectively removed in the following slow sand filters.

Because of the relatively low construction and operating cost, and comparatively simplicity in operation and maintenance, DHRF could be an appropriate pretreatment technology for urbanized areas too. Findings also indicate that it has a good potential to substitute the flocculation-sedimentation process and be used before rapid sand filters.

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Chapter 3

Modelling Separation of Flocculant Particles in a Horizontal-flow Gravel Bed ¹

ABSTRACT - Vertical or horizontal-flow gravel beds can be used in water treatment as roughing filters. In order to improve the performance of horizontal-flow roughing filtration (HRF) this process has been modified earlier by applying a constant coagulant dose of 1 mg Al(III)/l prior to the filtration (direct horizontal-flow roughing filtration). After rapid mixing, the destabilized suspension is passed on to the horizontal-flow gravel bed which is divided into 2 or 3 compartments. The first compartment, here referred to as the gravel bed, contains the largest grain size (15-25 mm diameter) and acts as a combined flocculator and settler. A mathematical model for the gravel bed was developed based upon an analogy with a parallel plate settler in which both the sedimentation and flocculation processes are incorporated. An experimental methodology was derived for applying the model to different suspension types using the results of column settling tests. The model was validated experimentally. The procedure developed here would allow the optimization of design and operation of the gravel bed for different raw water characteristics and for different process conditions.

Key words - particle separation, gravel bed, roughing filtration, modelling, sedimentation, flocculation, column settling test

INTRODUCTION

Horizontal-flow roughing filtration (HRF) in gravel beds has been proposed as a viable pretreatment process prior to slow sand filters in developing countries. They would be typically 8-12 m long and divided into 3 or 4 compartments with the first containing coarse gravel (15-25 mm diameter) followed by compartments with increasingly finer (8-5 mm diameter) grains. The design guideline for HRF (Wegelin *et al.*, 1987) based on laboratory tests suggests that the filtrate quality is 3-5 NTU for influent turbidity up to 200-400 NTU and filtration rate 0.5-1 m/h. However, field experience shows that effluent turbidity can remain as high as 50 NTU for such influent quality (Basit and Brown, 1986). Other limitations of HRF are that it operates at low filtration rate and consequently needs large

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filter volume, and that particle removal efficiency is low when influent water contains organic substances or colloidal particles of high stability (Ahsan *et al.*, 1991).

Considerable improvement in treating high-turbidity water has been achieved earlier by applying a small constant amount of coagulant, typically 1 mg Al(III)/l prior to the gravel bed (direct horizontal-flow roughing filtration-DHRF) (Ahsan *et al.*, 1991). An optimized DHRF would be 8 m long, 1.5 m high, 1.5-2.5 m wide, and divided into 2 compartments. The first compartment is 4-5 m long and filled with the largest gravel size. Lab scale pilot plant studies showed that water with high turbidity (200-400 NTU) and filtered at 5 m/h yields a constant effluent quality of 1-3 NTU with filter run time 3-5 days and approximately 15 cm total headloss development. DHRF has shorter filter runs than other roughing filter types of (weeks to months) because of the latter's lower initial turbidity (~20-100 NTU), filtration rate (< 1 m/h) and removal efficiency (70-90%). DHRF filter loading at breakthrough is 15-20 g/l. The turbidity removal profiles along the length of the gravel bed indicate that the first gravel bed compartment functions simultaneously as clarifier and flocculator. The subsequent compartment(s) with finer grains have the characteristics of deep bed filters.

Hereafter, the first compartment with large grains is referred to as the gravel bed. In order to optimize its configuration the objectives of this study were: (i) to develop a mathematical model for the gravel bed relating to the two underlying process mechanisms *i.e.* flocculation and sedimentation, and (ii) to outline a simple simulation experiment that allows use of the model for different suspension characteristics.

THEORETICAL FRAMEWORK

Camp and Stein (1943) related the settling velocity of discrete particles to be removed in a sedimentation tank to the surface overflow rate V_0 . Particles with settling velocities lower than V_0 are removed in proportion to their settling velocities. In the so-called high-rate settlers the plan area BL of a tank is increased by installing a number of inclined parallel plates or a bundle of tubes in it; the overflow rate can still be used as a basis for design (Yao, 1973). If a tank is divided by a number of parallel horizontal plates N_p (including the bottom) (see Notations-Appendix II):

$$V_0 = \frac{Q}{N_p \cdot BL} \quad (1)$$

In the case of flocculant suspension, the two principal causes of flocculation during sedimentation are (i) differential settling and (ii) velocity gradients (Camp and Stein, 1943).

The latter is more pronounced in a high-rate settler.

Because of the difficulty to determine individual particle settling velocities it is common practice to employ batch-type laboratory column settling tests as a basis of determining the settling velocity distribution (*e.g.* Metcalf and Eddy, 1991). For a column depth H_{CS} and detention time T_0 , Camp and Stein (1943) showed that the overflow rates in the column settler and in the flow-through tank ($N_p = 1$) can be equated

$$V_0 = \frac{H_{CS}}{T_0} = \frac{Q}{BL} \quad (2)$$

For discrete particles the settling velocities remain constant over the depth. The settling behaviour of flocculant suspension is complex due to particle size growth with depth and subsequent increase of particles' settling velocity. For a flocculant suspension under study ($C_0 = 200$ NTU, coagulant dose = 1 mg Al(III)/l) Fig. 1 shows the remaining concentration against column depth, for varying detention time t . At $t = 0$ the column has a uniform concentration C_0 (100%). If the column cross section area is A then the total amount of suspended solids till column depth H_{CS} is $A \cdot H_{CS} \cdot C_0$. At $t = t$ the remaining suspended solids amount down to H_{CS} is $A \int_0^{H_{CS}} C_t dH$ where C_t is the remaining concentration at time t and depth H . The ratio of remaining to initial concentration is $A \int_0^{H_{CS}} C_t dH / A \cdot H_{CS} \cdot C_0$ (area AED divided by area ABCD in Fig. 1) and the removal ratio RR can be expressed as

$$RR = \frac{\text{area } ABCE}{\text{area } ABCD} \quad (3)$$

Column settling predictions show good performance similarity for discrete particles in settling tanks but a safety factor (1.25-1.75) is commonly used to compensate for non-ideal flow in the tank (*e.g.* Schroeder, 1979; Metcalf and Eddy, 1991). The performance of the high-rate settler for removing flocculating particles can be correlated with that of a column settler by a similar safety factor in the overflow rate; the safety factor SF is a function of the suspension characteristics and settling system configuration, notably column depth (Sow and Thanh, 1983):

$$V_{0,HRS} = \frac{V_{0,CS}}{SF} \quad (4)$$

where $V_{0,HRS}$ and $V_{0,CS}$ are the overflow rate of the high-rate settler and the column settler,

respectively. If SF corresponds to column depth H_{CS} , then introducing V_0 from eqn. 2 in eqn. 4

$$V_{0,HRS} = \frac{H_{CS}}{T_0 \cdot SF} \quad (5)$$

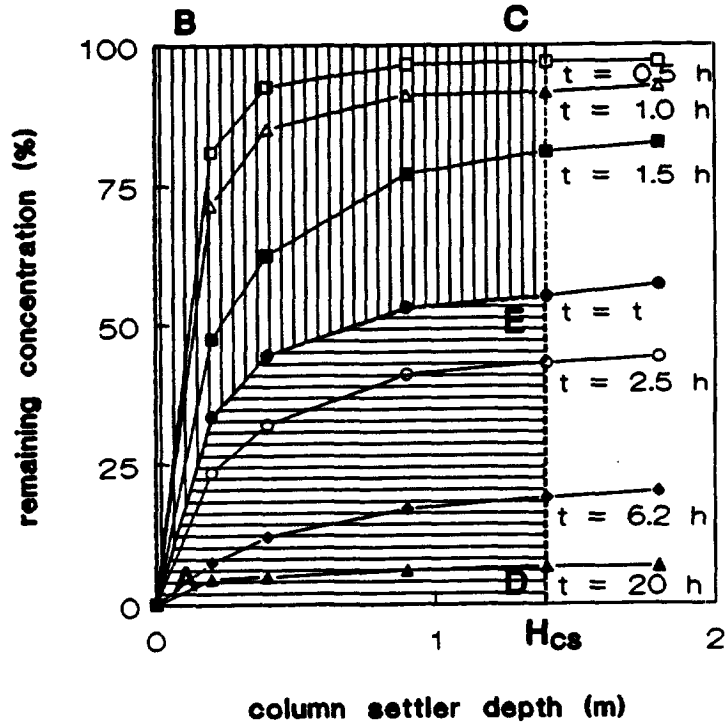


Fig. 1 Remaining suspended solids concentration vs. column depth with variable detention time t in the column settling test. $C_0 = 200$ NTU, coagulant dose = 1 mg Al(III)/l and rapid mixing time = 1 min. H_{cs} is the sampling depth in the column settler.

For a maximum floc size limited to a p -fold agglomeration of primary particles, the basic flocculation rate equation for a total initial number of flocs n_i is given as (Harris *et al.*, 1966).

$$\frac{dn_i}{dt} = -\frac{\gamma_p \alpha_p a^3}{\pi} G \phi n_i \quad (6)$$

where α_p is the collision efficiency, a the ratio of collision radius of a particle/floc to its physical radius, ϕ the floc volume fraction, γ_p the particle size distribution function, and G the velocity gradient.

In a granular bed the power dissipated by fluid flow is evinced as pressure drop ΔP . Ives (1975) defined G in rapid sand filters as

$$G = \left[\frac{v \Delta P}{\epsilon \mu \Delta L} \right]^{\frac{1}{2}}, \quad (7)$$

where G was derived using the Carman-Kozeny equation for linear flow resistance valid for the laminar flow regime ($Re < 5$). For a typical gravel bed with grain size 15-25 mm, porosity 0.4-0.45, flow rate 3-10 m/h, and temperature 5-25 °C, $Re = 10-50$. Wright (1968), studying the flow pattern through similar granular media, found that flow is in the steady inertial regime ($5 < Re < 90$) which means that there is no fluctuation of micro-velocity within the flow. However, the relationship between headloss and flow velocity is not linear, as in the case of the laminar flow. The pressure drop when $5 < Re < 100$ is approximated by (Huisman, 1984)

$$\frac{\Delta P}{\Delta L} = 130 \frac{\nu^{0.8} (1 - \epsilon)^{1.8}}{g} \frac{v^{1.2}}{d_g^{1.8}}. \quad (8)$$

G in the gravel bed is then

$$G = \left[\frac{0.114 (1 - \epsilon)^{1.34}}{\nu^{0.45} \epsilon^2 d_g^{1.34}} \right] v^{1.48}. \quad (9)$$

For typical gravel bed conditions as described above, the calculated G value is rather low at $G = 2-4 \text{ s}^{-1}$.

MODEL FOR THE GRAVEL BED

Several researchers (*e.g.* Swanson and Williamson, 1980; Wegelin *et al.*, 1987) suggested that sedimentation is the main particle separation process in a HRF gravel bed. This could also be visually observed from the trajectory of suspended matter. Coagulated suspensions consist of relatively large size flocs and, therefore, this hypothesis probably holds even more true for such suspensions. V_0 describes the particle/floc removal by sedimentation, and hence it will determine the removal ratios in the gravel bed. In the modelling approach presented here, the gravel bed configuration was schematized to derive expressions for the overflow rate and flocculation rate as a function of bed length. The column settling tests were used to simulate the processes in the gravel bed. The expressions for the overflow rate and the horizontal flocculation rate were incorporated in the formalized relationship that uses the results from the column settling test. A modified procedure to predict the gravel bed

performance with the help of the column settling test results is presented. The following assumptions were made:

- The gravel bed acts as a flocculator and settler simultaneously.
- One dimensional laminar flow occurs in the longitudinal (horizontal) direction. The total flow is distributed uniformly over the vertical cross-section of the gravel bed.
- Sedimentation is the particle removal mechanism. Particles are removed from suspension when they touch the top of a gravel grain.
- Flocculation in the gravel bed happens due to the differential settling of the flocculant particles, and due to the velocity gradients created by the flow in the horizontal direction.
- Particles are considered to agglomerate up to a certain maximum floc size. A constant average particle size distribution along the gravel bed is considered for simplicity in computations.
- G value is constant during the run time as the headloss increase is very low (about 1 mm/m bed length at the end of run time).

The gravel bed is conceptually approximated by a multiple-plate settler, hereafter referred to as the model settler. A gravel bed with length L , width B and height H (Fig. 2(a)) can be thought of as being filled with gravel of average diameter d_g in parallel layers, which can be further schematized as solid plates of thickness d_g (Fig. 2(b)). In order to compose the model settler configuration simulating the gravel bed's physical structure, overall dimensions are considered identical. Let N_p be the unknown number of equivalent plates and d_v the distance between them. Discharge Q and interstitial velocity are the same in the gravel bed and the model. The overflow rates (eqn. 1) in the two cases are equated by equating the net total horizontal surface area in the gravel bed S_0 and that of the model.

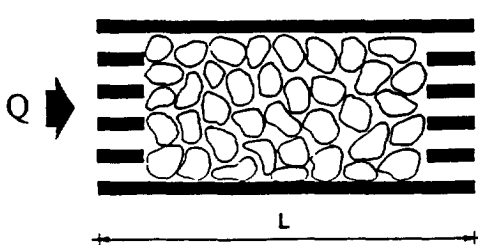


Fig. 2a Gravel bed of the first compartment of DHRF.

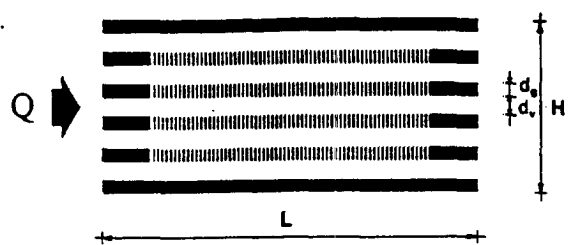


Fig. 2b Equivalent multiple plate model of the first compartment of DHRF.

The total length L of the model settler consists of a number of vertical sections (of equal thickness Δx) perpendicular to the longitudinal axis, and l_x is the distance up to the x -th section measured from the inlet of the model. Because S_0 is difficult to quantify with precision due to the irregularities in the gravel packing, a dimensionless correction factor k_0 is introduced. The net surface area available for sedimentation within a distance l_x is expressed as

$$S_{0,x} = k_0 B l_x \cdot N_p. \quad (10)$$

The in-pore velocities, and hence the detention times for the same lengths in the two systems are equated to obtain comparable removal efficiency at the same length. The in-pore (or between-plate) velocity is the superficial velocity v , divided by the porosity. For obtaining the same in-pore velocities for any given discharge Q , the porosity of the gravel bed ε is equated with the porosity of the model settler

$$\varepsilon = \frac{d_v \cdot k_0 B l_x N_p}{H k_0 B l_x}. \quad (11)$$

The number of plates N_p is (Fig. 2(b))

$$N_p = \frac{H}{d_v + d_g}. \quad (12)$$

From eqn. 11 and 12 we find

$$N_p = \frac{1 - \varepsilon}{d_g} \cdot H. \quad (13)$$

Now the overflow rate up to distance l_x can be expressed as

$$V_{0,x} = \frac{Q}{S_{0,x}} = \frac{Q}{k_0 N_p B l_x}, \quad (14)$$

and after introducing the value of N_p (eqn. 13) and $Q = vBH$ in eqn. 14, as

$$V_{0,x} = \frac{1}{k_0 (1 - \varepsilon_0) l_x} v d_g. \quad (15)$$

Flocculation due to horizontal flow

When a destabilized suspension is flowing through the gravel bed, particles are flocculated. At the same time particles are removed from the suspension and deposited onto the gravel. The particle concentration changes with the travel path length, also influencing the flocculation kinetics with the path length. Adjustment of the flocculation expression is thus required to allow for the changing flocculation rate. For a particular suspension its characteristics γ_p , α_p and a can be approximated as constants. Eqn. 6 can be expressed for the x -th section as

$$\frac{dn_{i,x}}{dt} = -k_1 \cdot \phi_x n_{i,x} G \quad (16)$$

where $n_{i,x}$ is the total particle number concentration, ϕ_x the floc volume fraction and k_1 a suspension characteristics constant. The ratio of the remaining floc volume fraction to the initial floc volume fraction in the suspension at the x -th section is equal to 1 minus the removal ratio (RR) up to the previous section *i.e.*

$$\frac{\phi_x}{\phi_0} = (1 - RR_{x-1}). \quad (17)$$

As an constant average particle size distribution is assumed, the floc volume fraction at any section will be proportional to the total number concentration in that section, therefore,

$$n_{i,x} = k_2 \phi_x, \quad (18)$$

where k_2 is a proportionality constant. Then from eqn. 16, 17 and 18

$$\frac{dn_{i,x}}{dt} = - \left[k_1 \right] \left[k_2 \cdot \phi_0^2 (1 - RR_{x-1})^2 \right] G. \quad (19)$$

Applying the expression of G (eqn. 9) in eqn. 19 and the transformation $dt = \epsilon dl/v$ we obtain

$$\frac{dn_{i,x}}{dl} = - \left[k_1 \right] \left[k_2 \cdot \phi_0^2 (1 - RR_{x-1})^2 \right] \left[k_3 \cdot v^{0.48} \right] \quad (20)$$

where k_3 is a constant defined as the first set of factors between parentheses in eqn. 9 divided

by ϵ ; it describes the granular media characteristic related to flow resistance. Eqn. 20 is now a general expression of flocculation kinetics due to horizontal flow of suspension in the gravel bed at the x -th bed section.

The flocculation rate is likely to change with bed length l_x due to reduction in ϕ (represented by the factor $(1-RR_{x-1})$). However, here the average flocculation rate \bar{dn}/dt can be used as the effect of $(1-RR_{x-1})$ on the model predictions is not substantial (see further). \bar{dn}/dt is defined as the summation of the flocculation rates in all the sections up to the x -th section divided by the number of sections:

$$\frac{\bar{dn}}{dt} = \sum_{k=1}^x \frac{1}{x} \frac{dn_{t,k}}{dt} = - \left[k_1 \right] \left[k_2 \cdot \phi_0^2 \right] \left[k_3 \cdot v^{0.48} \right] \sum_{k=1}^x \frac{1}{x} (1-RR_{k-1})^2. \quad (21)$$

Application of column settler data

Applying eqn. 5 for the model settler, the detention time up to a distance l_x can be expressed as

$$T_0 = \frac{H_{CS}}{V_{0,x} \cdot SF}. \quad (22)$$

Series of column settling tests with flocculant suspension showed that below a certain depth the percentage suspended solids remaining does not decrease further with increasing depth, irrespective of detention time (Fig. 1). This was found to occur at the sampling port at 0.9 m depth and downward. This implies that the removal ratio RR (according to eqn. 3) remains essentially unchanged with further increase in depth. Similar observations on the constant removal ratio beyond a certain depth (depending on suspension characteristics) are also reported in cases of water (Camp and Stein, 1943) and waste water treatment (White and Allos, 1976). This phenomenon occurs as with the column depth, the floc size and also its settling velocity increase. A maximum stable floc size exists beyond which the shear force becomes so high that it can erode part of the flocs. Smaller particles/flocs will continue to grow till the maximum floc size. As a result, after a critical depth the particle size distribution remains in a dynamic equilibrium. The settling velocities remain constant and the net effect is similar to that of discrete particle settling. Hereafter, the zone below this critical depth is referred to as pseudo-discrete settling zone. For a particular depth H_{CS}^* in this zone eqn. 22 can be re-written as

$$T_0 = \frac{H_{CS}^*}{V_{0,x} \cdot SF} = \frac{K_H}{V_{0,x}} \quad (23)$$

where K_H is a constant ($= H_{CS}^*/SF$).

In the gravel bed flocculation also occurs and hence the resultant G value and the net flocculation rate will be higher in the gravel bed than in the settling column. In other words, after a detention time T_0 in the gravel bed, an agglomeration efficiency is achieved that can be reached in the column settler only at a longer detention time T_{CS} . It is reasonable to presume that T_{CS} would be higher than T_0 by an amount that is function of the additional (average) flocculation rate (eqn. 21). Thus, the simulating detention time in the column settler (T_{CS}) which represents the equivalent time required to achieve the same agglomeration efficiency in the gravel bed at time T_0 is

$$T_{CS} = f \left[\frac{\bar{dn}}{dl} \right] \cdot T_0. \quad (24)$$

Using eqn. 21 the function of the average flocculation rate can be expressed in a general form as

$$f \left[\frac{\bar{dn}}{dt} \right] = f \left[\sum_{k=1}^x \frac{1}{X} \frac{dn_{t,k}}{dl} \right] = K_F \cdot \phi_0^p \cdot v^q \cdot \sum_{k=0}^x \frac{1}{X} (1 - RR_{k-1})^r, \quad (25)$$

where K_F , p , q and r are constants. Using this general expression in eqn. 24 we obtain

$$T_{CS} = \left[K_F \cdot \phi_0^p \cdot v^q \cdot \sum_{k=0}^x \frac{1}{X} (1 - RR_{k-1})^r \right] \cdot T_0. \quad (26)$$

Using eqn. 23 for detention time T_0 and eqn. 15 for the overflow rate $V_{0,x}$, the simulating detention time T_{CS} (eqn. 26) becomes

$$T_{CS} = \left[K_F \cdot \phi_0^p \cdot v^q \cdot \sum_{k=1}^x \frac{1}{X} (1 - RR_{k-1})^r \right] \left[K_H \cdot \frac{k_0(1 - \varepsilon)l_x}{v d_s} \right] \quad (27)$$

Combining the constants K_F , K_H and k_0 into K

$$T_{cs} = K \cdot \left[\phi_0^p \cdot v^q \cdot \sum_{k=1}^x \frac{1}{x} (1 - R_{k-1})^r \right] \left[\frac{(1 - \varepsilon) l_x}{v d_s} \right] \quad (28)$$

Eqn. 28 allows to calculate the simulating detention time T_{cs} in the column settler. The constants K , p , q and r and the corresponding settler depth (H^*_{cs}) are to be determined experimentally. Once reference T_{cs} and H^*_{cs} are known the removal ratio of the gravel bed can be determined from settling graphs (such as in Fig. 1) and eqn. 3.

MATERIALS AND METHODS

Experimental set-up

Experiments with a continuously operated lab-scale gravel bed were carried out under controlled conditions typical for DHRF. The model raw water was a kaolin suspension in Delft tap water (kaolin fine powder from Reidel-de Haën, Germany). The suspension was passed through a plain sedimentation tank (1 h detention time) to eliminate coarse particles. This presettled suspension was of 200 or 400 NTU ($\pm 2\%$) and particle size analysis (HIAC PC-320; sensor CBM-60) showed mean particle size by volume was $11.6 \mu\text{m}$ (standard deviation of the distribution $10.5 \mu\text{m}$). This suspension was delivered to the rapid mixing unit where a predetermined constant dose of $\text{Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$ analytical grade (Merck) was injected by a volumetric dosing pump (stock solution $3.75 \text{ mM Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$ prepared daily).

The container of the gravel bed was made of perspex tubes, each 20 cm in diameter and 2 m long except when otherwise mentioned. Tube units could be connected in series. They were filled with gravel with typical size of 15-25 mm (average 20 mm and porosity 0.42). The wall effect was presumed to be low despite a comparatively small tube diameter to grain size ratio of 10:1, because the headloss in the gravel bed with large grains is so low (1 mm/m bed length) that it is comparable to the headloss along the straighter pathways along the wall. Also, earlier comparison between the filter columns under study and a filter column of larger diameter with column to grain diameter ratio 20:1 showed turbidity profiles without statistically significant deviation. Care was taken during gravel packing to minimize any short-circuiting opportunities near the walls. No settling of gravel packing occurred during the filter run. Visual observation of particle velocity along the walls or top provided semi-quantitative evidence of the absence of short-circuiting along the walls. Water samples were collected iso-kinetically from the gravel bed at predetermined time intervals at the inlet, outlet and also from sampling ports at each 20 cm of bed length. Turbidity (Dr. Lange Trübungsphotometer LTP 4) was used as the main indicator of the removal performance and was expressed as NTU (Nephelometric Turbidity Unit). Filter breakthrough was defined to

occur when the effluent turbidity sharply increased.

Initial suspension turbidity of 200 and 400 NTU was selected to represent moderate and high turbid water. Filtration rates of 1, 3, 5 and 7 m/h were used which were within the feasible range of DHRF application. Experiments were also conducted with shorter bed lengths (2 m) to run a number of parallel experiments in a short time. Also, long bed lengths (8 m) were used to determine the gravel bed performance beyond the usual 4 m length of DHRF compartments. For a typical 4 m bed length the filter ripening period was short (3 h) and run time 4-5 days. The same suspension was also used in the column settling analysis. The column was made out of perspex and was 2 m high and 20 cm in diameter. Each time average values of duplicate column settling tests were taken; deviation was $< 3\%$.

Calibration of the model

Separate gravel bed runs were made with bed lengths of 2 and 4 m to calibrate the model. Effluent turbidity and also the turbidities in the different bed sections remained essentially constant (deviation from time-average over working stage run time $< 10\%$) for a long time (about 4 days). The time-average turbidity value was taken to calibrate the model. Settling graphs (*e.g.* Fig. 1) were drawn by conducting column settling tests. For a bed length l_x the corresponding T_{CS} value (eqn. 28) was calculated. The calculated T_{CS} values and a preselected reference column settler depth (H^*_{CS}) were applied in the settling graph to determine the removal ratios (according to eqn. 3) up to l_x . A spreadsheet computer program was developed for the consecutive computations. The computer program also calculated the model constants (K , p , q , and r in eqn. 28) by a trial and error procedure. The model parameters were selected that best fitted the experimental results.

Several column depths in the column settling test were tested to determine their suitability to simulate the gravel bed process. Predictions with depth > 0.9 m (pseudo-discrete settling zone) yielded identical results (see below). A reference column depth of $H^*_{CS} = 1.4$ m was selected as an optimum, combining minimal depth and a safety margin from the starting point of the pseudo-discrete settling zone. This H^*_{CS} was used in all further calculations.

Calculations showed that p and r equal zero. Therefore, the factors containing the term ϕ_0 and $(1 - RR_{x-1})$ in eqn. 28 can be neglected. This probably occurs because the differences in flocculation and settling kinetics are directly simulated in the column settling analysis and incorporated in the removal ratio calculation procedure. The value of q was determined to be 0.5 and of K 3.46. Therefore, eqn. 28 can be simplified to

$$T_{CS} = 3.46 v^{0.5} \left[\frac{(1 - \varepsilon) l_x}{v d_s} \right] \quad (29)$$

RESULTS AND DISCUSSIONS

Validation of the model

The model was expected to predict the average residual turbidity profile during its working stage (except the relatively short ripening period). In order to validate the model, experiments were carried out with 8 m bed lengths. This allows to test whether the model could be extrapolated beyond the 2-4 m lengths that were used for the more comprehensive calibration procedure. The results of the model predictions based on the column settling tests and the calculated model constants are shown together with the experimental results for the filtration rates of 3 and 5 m/h in Fig 3. The error bars represent \pm one standard deviation during an experimental run. C_0 was 200 NTU, G 200 s^{-1} and T_{RM} 1 min. The turbidity removal profiles along the gravel bed length agreed well ($R = 0.89$) to very well ($R = 0.98$) with the model predictions and also showed that the model is applicable for varying lengths. Beyond 5 m bed length the remaining turbidity was moderately lower (10-20%) than the model prediction. The particle size at the gravel bed inlet was 13.3 μm (standard deviation of the distribution 8.4 μm) and after 4 m length 12.8 μm (standard deviation 8.9 μm).

The applicability of the model for a wider filtration rate (1-20 m/h) was examined with a rectangular gravel bed ($L = 2$ m, $B = 0.2$ m and $H = 0.4$ m) while retaining all other process conditions. The increasing effluent turbidity with increasing filtration rates is shown in Fig 4. It can be observed that the model appeared well applicable for a wide range of practical filtration rates. According to the model, differences in cross-section or depth would not influence results. This was confirmed in these experiments where a rectangular cross-section with a depth twice that of the circular tube's diameter was used with all other conditions identical. The effect of different initial suspension turbidity was also studied in a 4 m tubular long bed. In Fig. 5 model predictions and the experimental results are compared for two suspensions with characteristics different from the one previously described: (i) $C_0 = 200$ NTU, coagulant dose = 1.25 mg Al(III)/l, and (ii) $C_0 = 400$ NTU and coagulant dose = 1 mg Al(III)/l. As can be seen, the model (where column settling tests were carried out with appropriate suspension) also fits well in the cases of higher effluent turbidity as well as for higher filtration rates.

Process mechanisms

Higher filtration rates induce higher G values within the gravel bed which are expected to

cause higher flocculation rates. The effect of this higher rate is not directly reflected in the experimental batch process of the column settler, but it is compensated in the model through the parameter v^q ($q = 0.5$) (eqn. 29). The validity of the model for a wide range of filtration rates (1 to 20 m/h), and hence flocculation rates, supports the earlier hypothesis that the gravel bed acts as a combined flocculator and settler.

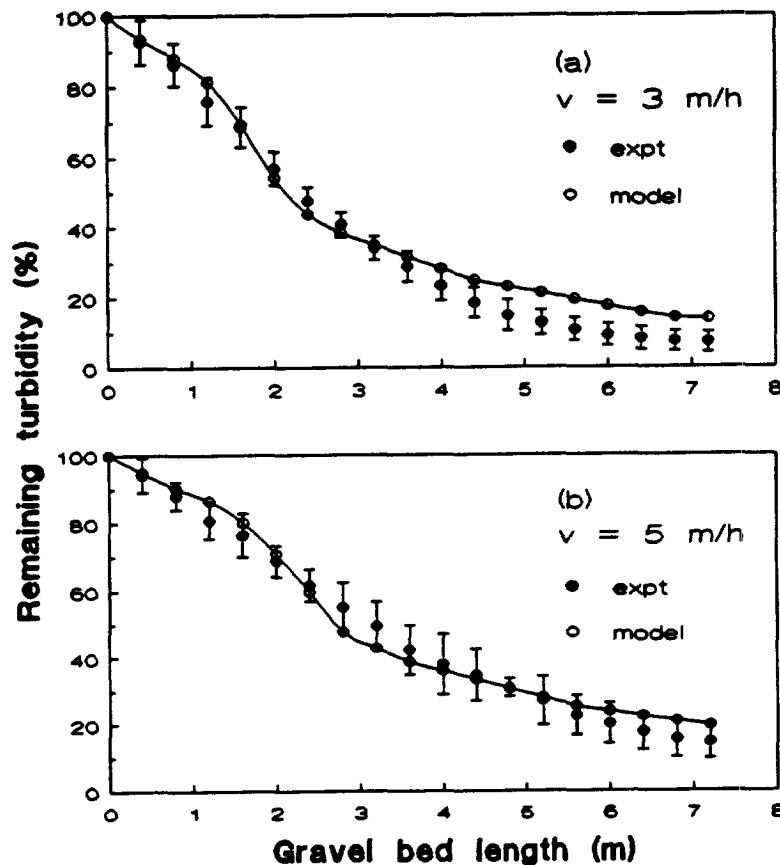


Fig. 3 Results of the model prediction and of the experiments with flocculated kaolin suspension as a function of the gravel bed length ($C_0 = 200$ NTU, coagulant dose = 1 mg Al(III)/l, $d_0 = 20$ mm) and for filtration rates (a) $v = 3$ m/h, (b) $v = 5$ m/h.

The theoretical average flocculation rate is proportional to $v^{0.48}$ (eqn. 21). The calculated value of q ($=0.50$) was similar. With increasing v , the contribution of the flocculation factor ($v^{0.5}$) increases whereas that of the overflow rate factor (v^{-1}) decreases (eqn. 28 and 29). The net effect is that T_{CS} and hence the removal ratio in the gravel bed, are function of $v^{0.5}$. Wegelin *et al.* (1987) and Lebair (1992) reported that the removal ratio in HRF it is proportional to $v^{-0.88}$ and $v^{-0.52}$, respectively. The removal efficiency in rapid sand filters is also inversely proportional to filtration rate (Ives, 1975). The prediction of effluent quality at a particular bed length is based on the overflow rate pertaining to that length. The validity

of the model over the whole bed length suggests that the overflow rate, corrected for flocculation, can be used as a main design parameter for the gravel bed. Similar recommendations were made for plain and high-rate settlers (e.g. Camp and Stein, 1943; Yao, 1973). This critical role of overflow rate supports the hypothesis that sedimentation is the dominant particle removal mechanism.

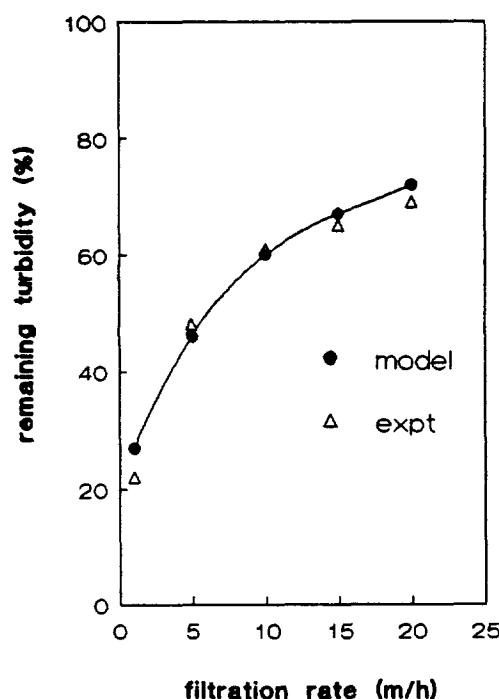


Fig. 4 Remaining turbidity against filtration (flow) rate. $L = 2$ m, $d_g = 20$ mm, $C_0 = 200$ NTU, coagulant dose = 1 mg Al(III)/l.

Influence of the column depth

The influence of H_{cs}^* on the model prediction is shown in Fig. 6. The values of the model constants were kept unchanged, but samples from depths (H_{cs}^*) 0.9 and 1.8 m (both within the pseudo-discrete settling zone and 0.5 and 0.4 m, respectively, distant from the reference sampling depth of 1.4 m) were also used in the computational procedure. Results derived from the three sampling depths varied only by 3%. This confirms the earlier assumption that selection of any depth in the pseudo-discrete settling zone would give essentially the same results. Therefore, the experimental procedure can be simplified and be reliable.

Application of the model

The effect of the different flow rate (Fig. 4), and of the different influent suspension

characteristics with varying turbidity and coagulant dose (Fig. 5), on the turbidity profiles was simulated well by the model through the use of the column settling tests. In the computations the same model constants ($K = 3.46$ and $q = 0.5$) and reference column depth ($H_{cs}^* = 1.4$ m) were used. It can be extrapolated that moderate changes in media characteristics (*e.g.* grain size and shape) and water chemistry would not reduce the validity of the approach, however, the model constants could require to be re-calibrated.

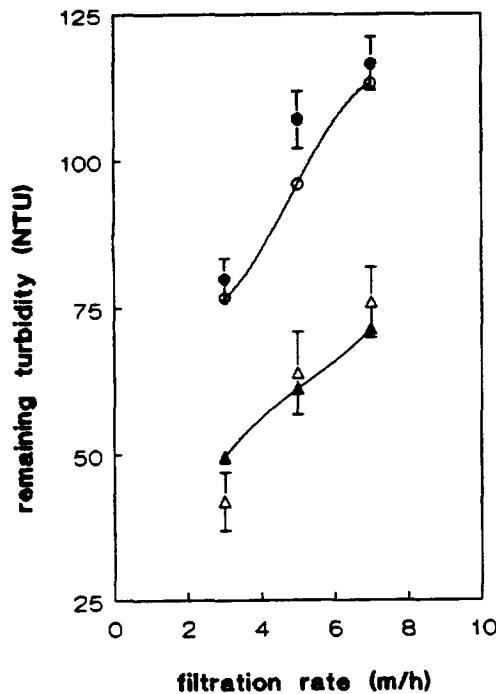


Fig. 5 Effluent concentration (NTU) against filtration (flow) rate for 2 suspension characteristics. Suspension (i) $C_0 = 200$ NTU, coagulant dose = 1.25 mg $Al(III)/l$; symbols \blacktriangle and \triangle represent model prediction and experimental data, resp.; and, suspension (ii) $C_0 = 400$ NTU, coagulant dose = 1 mg $Al(III)/l$; symbols \circ and \bullet represent model prediction and experimental data, resp. $L = 4$ m, $d_0 = 20$ mm.

The model is not intended for deep bed filtration where plain sedimentation is not the main removal process. Thus, filter grains should be larger than a critical minimum size. The model is expected to be valid in the roughing filtration range (grain size > 4 mm) and in the working stage of the filtration cycle. The mean particle size at the gravel bed inlet was $13.3 \mu m$ (standard deviation of the distribution $8.4 \mu m$) and after 4 m it changed to $12.8 \mu m$ (standard deviation $8.9 \mu m$). The net effect of the two simultaneous processes *i.e.* sedimentation (preferential removal of larger particles) and flocculation (progressive formation of larger aggregates) resulted in these similar particle size distributions along the bed. Therefore, the assumption of an average constant particle size distribution was justified.

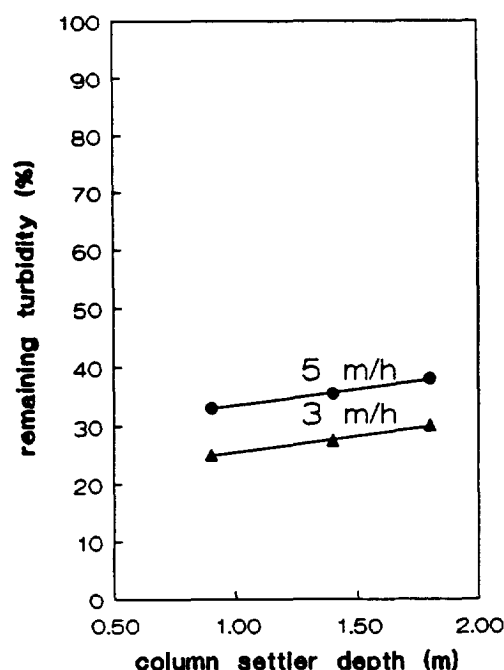


Fig. 6 Sensitivity of the effluent concentration predicted by the model to the reference depth of column settler H_{cs}^* with model constants $K = 3.46$ and $q = 0.5$. $L = 4$ m, $d_g = 20$ mm, $C_0 = 200$ NTU, coagulant dose = 1 mg Al(III)/l.

CONCLUSIONS

A model to predict the particle separation in the gravel bed (average gravel size 20 mm) has been developed as an analogy to a multiple-plate settler and where both the flocculation and the sedimentation processes are incorporated. A procedure is also formulated to use the experimental data from column settling analysis in the model.

The modelling procedure allows to optimize the coagulation process condition in the rapid mixing unit. A particular raw water can be coagulated under different coagulation conditions (with respect to coagulant dose, G and T_{RM}); column settling test results obtained with the same suspension can be inserted into the model to predict the turbidity removal performance as a function of the gravel bed length. As the performance of the gravel bed strongly influences the total performance of DHRF (Ahsan *et al.*, 1991), optimizing the process conditions (coagulation condition and filtration rate) in the gravel bed will result in optimization of the overall DHRF performance.

The model appears applicable for varying gravel bed lengths and depths. The model constants ($K = 3.46$, $q = 0.5$) and the reference sampling depth of the column settler ($H_{cs}^* = 1.4$ m)

hold for the range of flocculated suspension characteristics (turbidity 200-400 NTU, coagulant dose 1-1.15 mg Al(III)/l) and filtration rate (1-20 m/h) for the gravel bed under study. The model constants may, however, require to be re-calibrated accordingly for different gravel bed characteristics. The model approach would be valid for many natural waters with a wider range of suspension characteristics and water chemistry as the different settling characteristics will be incorporated in the column settling tests.

The model supports the earlier hypothesis that for flocculating suspensions the gravel bed acts as a combined flocculator and settler. The model parameter (v^*) indicates that the higher flocculation rate, as a result of higher flow velocity, contributes positively to the removal ratio. The model also supports the assertion that sedimentation is the dominant particle separation process.

APPENDIX I. - REFERENCES

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APPENDIX II.-NOTATION

The following symbols are used in this paper:

- a = ratio of collision radius of a particle/floc to its physical radius
- A = cross sectional area of column settler, m^2
- B = width, m
- C_0 = initial particle concentration
- C_t = particle concentration at time t
- d_g = average diameter of filter grain, m
- d_v = depth between two consecutive plates of model settler, m
- DHRF = Direct Horizontal-flow Roughing Filtration
- f = function
- g = gravitational acceleration (9.81), m/s^2
- G = velocity gradient ($= \sqrt{(P/\mu V)}$), s^{-1}
- H = height or depth (in settling column), m
- $H_1...H_5$ = depths of column settler measured from surface to the sampling ports, m

- H_{CS} = particular depth of column settler, m
 H_{CS}^* = reference H_{CS} in the discrete settling zone, m
 HRF = Horizontal-flow Roughing Filtration
 K, K_H, K_F = constants
 K_A, K_B = floc aggregation and breakup constant
 k = integer values
 k_0, k_1, k_2, k_3 = constants
 L = length, m
 ΔL = length of an elementary gravel bed, m
 l = length of gravel bed measured from the inlet
 l_x = l up to x -th vertical section (Δx)
 N_p = number of plates in multiple plate settler
 NTU = Nephelometric Turbidity Unit
 n = particle number concentration
 n_t = total n
 $n_{t,x}$ = n_t at x -th section
 P = total power dissipated, Nm/s
 ΔP = pressure drop across a gravel bed, N/m²
 p, q, r = exponents in eqn. 32
 Q = discharge ($= vBH$), m³/h
 RR = removal ratio
 S_0 = total available surface area on the top of grains, m²
 $S_{0,x}$ = S_0 up to l_x , m²
 SF = safety factor
 T_0 = detention time in column settler, h
 T_{CS} = simulating detention time in column settler, h
 T_{RM} = rapid mixing time, min.
 V = volume of a reactor, m³
 V_0 = overflow rate, m/h
 $V_{0,x}$ = V_0 up to x -th section, m/h
 $V_{0,CS}, V_{0,HRS}$ = V_0 in column settler and in high-rate settler, respectively, m/h
 v = superficial velocity, m/h
 Δx = thickness of equally divided vertical sections of the gravel bed length

Greek letters

- α_p = fraction of collisions which results in aggregation
 γ_p = particle size distribution function
 ε = gravel bed porosity
 ϕ = floc volume fraction

$\phi_0, \phi_x = \phi$ initially and at x -th section

$\mu =$ dynamic viscosity, Ns/m^2

$\nu =$ kinematic viscosity, m^2/s

Chapter 4

Process Mechanisms¹

ABSTRACT - The various coagulation and clarification process mechanisms occurring in direct horizontal-flow roughing filtration is investigated. An integrated alum coagulation stability diagram for high turbidity water is developed in which the zones of dominant coagulation mechanisms are demarcated in terms of coagulant dose and pH. The boundaries of flocculation, sedimentation and filtration process in a granular bed have also been demarcated in terms of grain size and filtration rate. The use of coagulants increase the filter coefficient to about 4 folds in roughing filtration as well as in rapid sand filtration.

INTRODUCTION

The processes in Direct Horizontal-flow Roughing Filtration (DHRF) can broadly be divided into (i) coagulation in the rapid mixing unit, and (ii) clarification in the subsequent granular media. Here clarification refers to either one of or combination of the flocculation, sedimentation and filtration process.

Coagulation in the rapid mixing unit of DHRF is carried out for better particle separation in the subsequent granular media. To ascertain the optimum coagulation condition in terms of dosage and pH in conventional water treatment plants, a comprehensive design diagram for relatively low turbidity water (15-25 NTU) was developed indicating zones of different coagulation mechanisms (Amirtharajah and Mills, 1982). DHRF, on the other hand, is intended to be used for highly turbid waters (200-400 NTU). Experimental studies of coagulation, flocculation and sedimentation of highly turbid waters are stated to be limited (Guibai and Gregory, 1991). Also, the particle separation processes in a conventional flocculation-sedimentation process are different from those in a granular medium. Therefore, the design diagram developed for conventional water treatment plants may not be directly applicable to DHRF.

Coagulated water from the rapid mixing unit is passed onto the granular media of DHRF for clarification. In the water and wastewater treatment granular media can be used as a filter as well as a flocculator. The three important parameter groups which determine the general performance of a granular medium are: (i) the suspension characteristics, (ii) the granular medium characteristics, and (iii) the flow rate.

A filter can be slow sand, rapid sand or roughing filter mainly depending on its grain size.

¹Part of this Chapter submitted as a paper by T. Ahsan and G. J. Alaerts in *Wat. Res.*

Gravel bed flocculators, which are intended to enhance particle agglomeration for its subsequent better removal, use higher flow rates and a wide range of larger grains. Different processes such as flocculation, sedimentation and filtration can be simultaneously operative in a granular medium and their resultant determines its performance. In slow and rapid sand filters, filtration is the dominant process whereas, in a gravel bed flocculator, flocculation is dominant. However, in roughing filters more than one process can be dominant depending on the grain size and flow rate.

It has been reported by many researchers (e.g. Wegelin et al., 1986; Boller, 1993) that sedimentation is the dominant particle removal process in roughing filters. For coagulated waters it was observed that for 20 mm diameter grains the dominant mechanism is sedimentation whereas, 8 mm grain media exhibit the characteristics of deep bed filtration (Chapter 2). The limits of the dosage-pH domains in which the different mechanisms (flocculation, sedimentation and filtration) are applicable and their translation in terms of grain sizes and flow rates, are not well defined in literature. It is important to understand the different removal mechanisms in order to optimize the overall process.

This Chapter investigates the boundaries of the applicability of mechanisms associated in the two different areas relevant to DHRF: coagulation of highly turbid water, and the processes related to clarification in the granular media. The specific objectives are:

- to investigate the boundaries of different coagulation mechanisms (in terms of coagulant dose and pH) for highly turbid water and to determine the range of optimum coagulation parameters for the influent to the roughing filters, and
- to investigate the boundaries of dominant mechanisms related to the flocculation, sedimentation and filtration in a granular medium in terms of grain sizes and flow rates.

COAGULATION IN THE RAPID MIXING UNIT

Coagulation may be defined as destabilization of particles and is achieved by adding chemicals (coagulants) in a rapid mixing unit (Weber, 1972). Flocculation is considered as a separate process taking place after coagulation. It involves in aggregation of the particles to form larger settleable or filterable flocs. Coagulation can be followed by (i) flocculation and sedimentation in conventional plants, (ii) direct filtration, or (iii) roughing filtration for pretreatment in DHRF.

Aluminium Sulphate is today the most widely used coagulant in Europe, the United States (AWWA, 1981) and developing countries. Selection of the optimal coagulation process for formation of aggregates having appropriate physical and chemical properties are essential for effective particle removal in water treatment plants (Boyd and Ghosh, 1974; O'Melia, 1985; Wiesner et al., 1987).

Four mechanisms of coagulation are recognized:

- (i) double layer compression (Black and Chen, 1965);
- (ii) adsorption-destabilization i.e. adsorption to produce overall particle charge

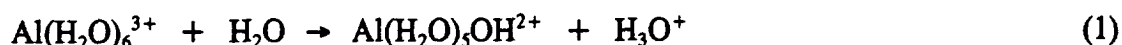
neutralization (O'Melia and Stumm, 1967; Hahn and Stumm, 1968), or change surface charge in patches that will electrostatically attract counter charged patches on other particles (Gregory, **);

- (iii) sweep coagulation i.e. enmeshment in a coagulant hydroxide precipitate (Packham, 1965); and
- (iv) adsorption to permit interparticle bridging (only applicable to organic polymeric coagulants) (Stumm and O'Melia, 1968).

Under practice conditions, coagulation in water and wastewater is most often accomplished by adsorption on the colloid of oppositely charged soluble and insoluble hydrolysed coagulant species and subsequent destabilization (adsorption-destabilization), enmeshment of the colloid within hydroxide precipitate, or both (O'Melia, 1972; Alaerts and Van Haute, 1982; Amritharajah and Mills, 1982; Johnson and Amritharajah, 1983). Adsorption-destabilization occurs when metals such as Fe(III) or Al(III) are hydrated in water and produce charged aquocomplexes. Because of adsorption of those onto oppositely charged surfaces, the colloids become destabilized i.e. their repulsive forces are reduced. This mechanism implies the possibility of restabilization (over-charging) and also of the existence of a stoichiometric relationship between colloid surface area and coagulant dose (O'Melia and Stumm, 1968; O'Melia, 1972; Alaerts and Van Haute, 1981).

The theory of enmeshment of colloids in a precipitate was advanced to explain the non-stoichiometric behaviour of coagulants and colloids (AWWA, 1971). If sufficient alum is added to a colloidal dispersion in water, oversaturation occurs (e.g. when $[Al] > 1.3 \text{ mg/l}$ at pH 7.5) in such extent that large quantities of aluminum hydroxide precipitate rapidly, enmeshing colloidal particles in the growing flocs and physically removing the particles from the solution (O'Melia, 1972).

When aluminium salts are dissolved in water at $\text{pH} \leq 4$, the metal ion Al^{3+} hydrates, coordinating six water molecules and forming an aquametal ion, $\text{Al}(\text{H}_2\text{O})_6^{3+}$. As the pH is raised or as the concentrated coagulant is added to buffered water, hydrolysis (liberation of one or more hydrogen ions from the H_2O molecules) decreases the net charge on the hydrated metal ion resulting in the formation of several hydrolysis species. OH^- ligands successively replace the six coordinated H_2O molecules. For example, the first hydrolysis reaction can be written as:



For simplicity H_3O^+ can be expressed as H^+ and H_2O ligands attached to the Al ions can be omitted. The hydrolysis reactions are shown in Annexure I. Based on this the stability diagram showing concentrations of the different aluminium species as a function of pH can be drawn (Fig. 1). The nature of the original particles and presence of other reactive solutes will also determine the extent of destabilization (Alaerts and van Haute, 1981).

The coagulation diagram

The integrated alum coagulation diagram in Fig. 1 was developed by Amirtharajah and Mills (1982) by superimposing extensive data from literature on coagulation mechanisms (e.g. Packham, 1962; Hanna and Rubin, 1970; Matijevic, 1973; Rubin and Kovac, 1974) on the

stability diagram of alum. Thus, it was possible to synthesize the pH-alum dosage conditions for effective destabilization with different coagulation mechanisms in a single diagram. It should be noted that the stability lines and the boundaries of different mechanisms only provide an underlying framework to understand the coagulation reactions mechanisms. The optimum coagulation range was also validated by Amirtharajah and Mills (1982) using experimental data from Tekippe and Ham (1970) and Trussel (1978) for different types of low turbidity waters. A coagulation diagram for iron coagulation was also prepared (Johnson and Amirtharajah, 1983). In a series of pilot-plant experiments (Amirtharajah and Suprenant, 1984) it was verified and confirmed that the adsorption-destabilization zone in the diagram is the most effective (good removal, relatively low headloss) for direct filtration.

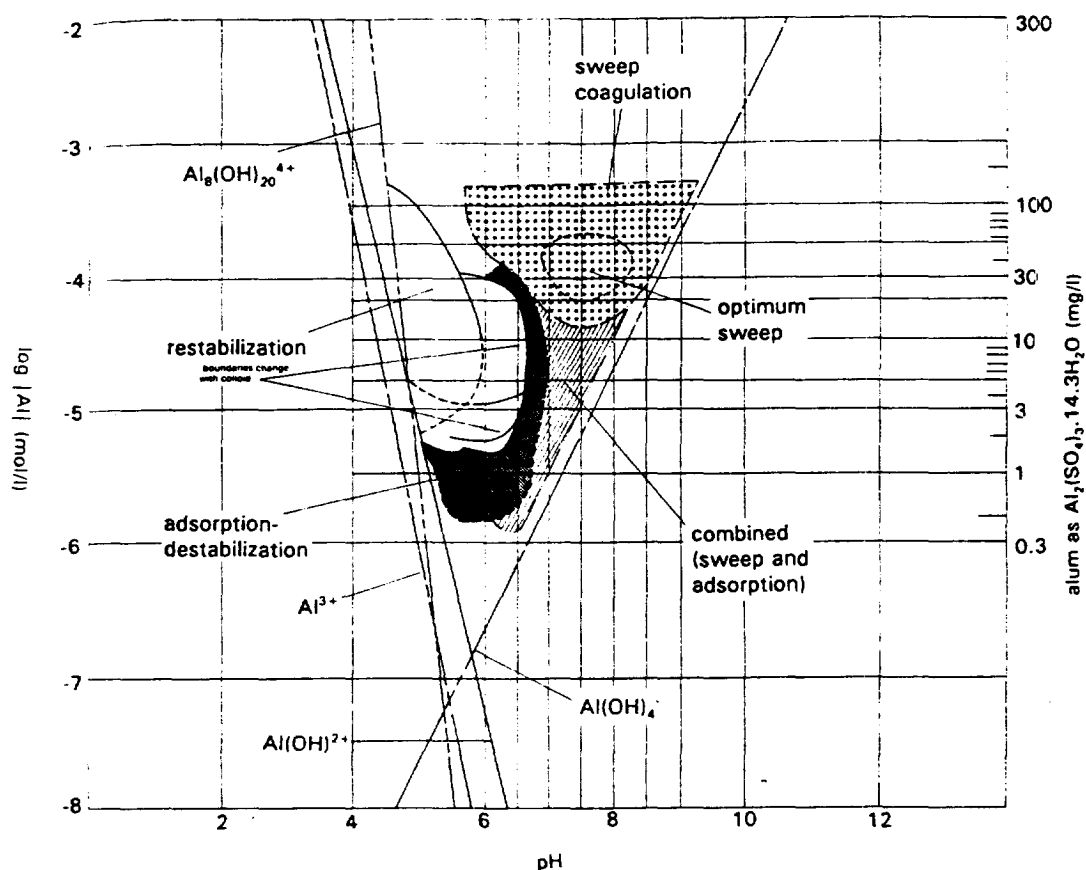


Fig. 1 Integrated coagulation diagram for alum coagulation with low turbidity water (15-25 NTU) (Amirtharajah and Mills, 1982).

In most cases, jar tests can be used to determine the optimum coagulation condition range required for direct filtration (Yao et al., 1971; Habibian and O'Melia, 1975). With the help of a series of jar tests Edward and Amritharajah (1985) showed that optimum domains of colour removal can be predicted by the alum coagulation diagram as well. A similar approach of construction of coagulation diagrams for particular coloured waters was also used by other researchers in regular coagulation and in direct filtration (e.g. Dempsey, 1984; Hundt and O'Melia, 1988; Graham et al., 1992; Rebhun and Lurie, 1993). Therefore, the coagulation

diagram is likely to be a useful graphical method for roughing filters treating coagulated waters as well (as in this case in DHRF).

The alum coagulation diagram in Fig. 1 was prepared for relatively low turbid waters (15-25 NTU). However, the turbidity or colloid type and concentration themselves affect the coagulation diagram. For example, the lower boundaries of the restabilization zone (Fig. 1) can be expected to change with colloid concentration as a stoichiometric relationship exists between colloid surface area and coagulant dose (Stumm and O'Melia, 1967, 1968). Coagulation before roughing filters on the other hand is intended for pretreatment of highly turbid waters (100-400 NTU and occasionally higher). Although it is expected that the general framework of the alum coagulation diagram would remain similar, the boundaries of different mechanisms are likely to change.

PROCESSES IN A GRANULAR BED

Filtration

Filtration is defined as the process of removing suspended solids from a fluid by passing it through a porous medium (AWWA, 1971; Faust and Aly, 1983; James Montogomery Inc., 1985). The filtration process can also be classified according to the location where the deposits are retained: surface filtration and deep bed filtration. In surface or cake filtration the bed consists of small size grains and most of the deposits are retained on and in the top layer. For example, in slow sand filters (grain size < 0.5 mm) most of the deposits are retained in the first 5 to 10 cm of the bed depth. As a consequence the majority of the headloss occurs at the top surface of the bed. In deep bed filters, e.g. rapid sand filters (grain size ~ 1 mm) or roughing filters (grain size > 2 mm) particles are retained along the passage of the suspension through the bed. As the grain size is bigger, the deposits penetrate deeper into the bed. The maximum headloss occurs at certain depth from the top surface, depending on the grain size and filtration condition.

Fig. 2 shows the operational ranges of typical different granular media units (slow sand filters, rapid sand filters, roughing filters, DHRF and gravel bed flocculators) in terms of required flow rate and grain size. The influent turbidity before these units varies; about 10-20 NTU before slow sand filters, up to 50 in rapid sand filters. It can be few hundred NTU before roughing filters, DHRFs or gravel-bed flocculators. As seen in Fig. 2 the range of the parameters is relatively narrow for slow and rapid sand filters. In roughing filters (with uncoagulated water) a wide range of grain sizes can be used but the flow rate is limited to low values within a narrow band (0.5 to 1.5 m/h). Preliminary design for DHRF shows that a higher flow rates and a similar wider range of grain sizes (5-20 mm) can be selected (Chapter 5 and 8). The gravel bed flocculator (described latter) also covers a wide range of flow rate and grain size.

The filter efficiency or filter coefficient (λ) has been defined by Iwasaki (1937) by the following kinetic filtration equation of the first order in to particle concentration:

$$\frac{\partial C}{\partial L} = -\lambda C. \quad (2)$$

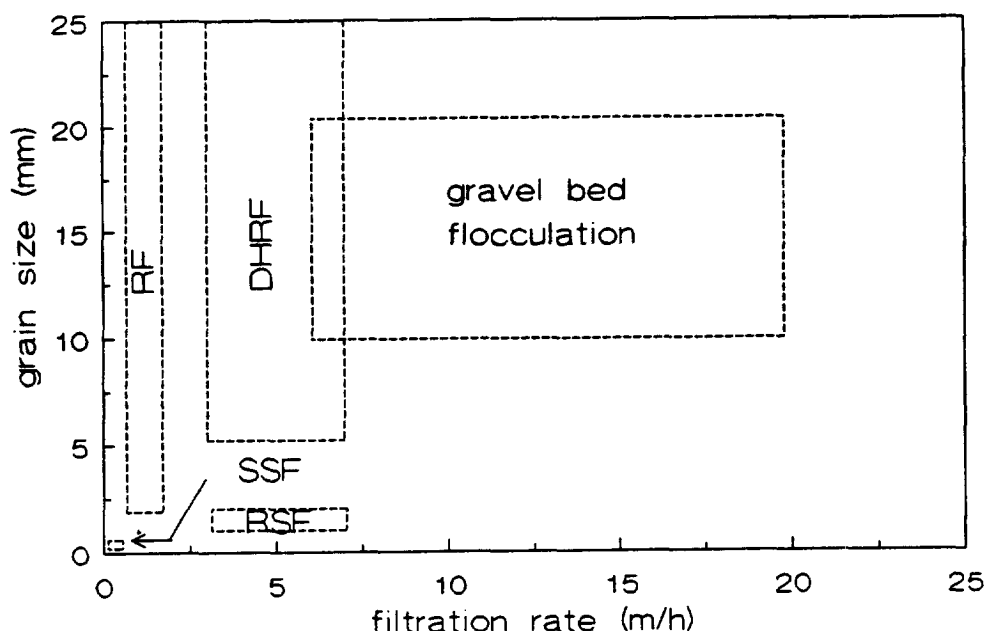


Fig. 2 Zones with dominant particle removal or agglomeration mechanisms in granular media. DHRF = direct horizontal-flow roughing filtration, RF = roughing filtration, RSF = rapid sand filtration and SSF = slow sand filtration.

The particle removal in a filter can be separated into two steps: (i) *transport* which carry the particles from the bulk of fluid to regions close to grain surface, and then (ii) *attachment* when the forces of attraction causes a capture of the particles and its attachment to the grains. The important transport mechanism in water and wastewater are sedimentation, interception, diffusion and hydrodynamic action (Boyed and Ghosh, 1974; Ives 1975a). The single collector efficiency η of those mechanisms are given in Table 1. The total transport efficiency η_0 is considered to be the summation of the individual efficiencies. Ives presented the total transport efficiency in a general form as

$$\eta_0 = \text{constant} \cdot \frac{d_p^{\alpha-\beta+2\gamma}}{\mu^{\beta+\gamma-\delta} d_s^{\alpha+\beta+\delta} v^{\beta+\gamma+\delta}} (KT_k)^\beta \frac{(\rho_s - \rho)^\gamma}{\rho^\delta}, \quad (3)$$

where α , β , γ and ρ are positive exponents. Eqn. 3 shows that a minimum efficiency can occur at a certain particle size; this was at $d_p = 1 \mu\text{m}$ and was experimentally verified (Yao et al, 1971). At $d_p < 1 \mu\text{m}$ diffusion is increasingly dominant whereas at $d_p > 1 \mu\text{m}$ sedimentation is increasingly dominant. Eqn. 3 also shows that smaller grains, and smaller filtration rates improve collection efficiency, which is well-known in practice.

The total particle removal efficiency is a product of two factors: the transport efficiency η_0 , and the attachment efficiency α_0 . α_0 may be equal to 1 or less depending on the degree of particle destabilization.

Table 1 Filter transport mechanisms.

mechanisms	efficiency	reference
sedimentation (gravity)	$\eta_G = \frac{(\rho_s - \rho)gd_p^2}{18\mu v}$	Ranz and Wong (1952)
interception	$\eta_I = \frac{3}{2} \left[\frac{d_p}{d_s} \right]$	Agrawal (1966)
diffusion	$\eta_D = 0.9 \left[\frac{kT}{\mu d_p d_s v} \right]^{2/3}$	Yao et al. (1971)
hydrodynamic action	$\eta_H = \left[\frac{\mu}{v d_s \rho} \right]^c$	Ives (1975a)

c is a constant

Sedimentation

In sedimentation particles are separated from the suspension under the influence of gravity. Particles which have little or no tendency to flocculate settle at a constant rate; this is called discrete settling. Flocculating particles collide with each other during settling and form bigger agglomerates with increasingly higher settling velocities. This process is referred to as flocculant settling. Regular settling basins or multiple plate settlers are commonly used for sedimentation. Gravel beds can also be treated as sedimentation units as similar processes occur.

The concept of sedimentation in the gravel bed was used i.a by Einstein (1968), Wegelin (1987), and Ahsan et al. (1991) to describe the processes in analogous to those occurring in a multiple plate settler. A similar model has been developed in order to describe the sedimentation in a horizontal-flow gravel bed and also to predict its removal efficiency

(Chapter 3). Settling is considered as the main removal mechanism, and the effect of simultaneous flocculation is also considered there. According to the sedimentation theory (Camp, 1946), and as also assumed in the model, the particles are considered removed when they touch the bottom of the tank or the top of a gravel; resuspension is not considered. In filtration terms, this means that the attachment efficiency $\alpha_0 = 1$.

Gravel Bed Flocculation

A granular medium acts as a gravel bed flocculator when particle agglomeration dominates the particle capture mechanisms and produces settleable or filterable flocs. Flocculation is achieved as a combined result of the flow through a tortuous pathway thus creating head loss and inter-particle collision opportunities, and of surface contact between coagulated suspension and the surface of the grains. Hydraulic flocculators such as gravel bed flocculators have the advantage that they do not require mechanical mixing equipment; they are relatively less costly than conventional mechanical flocculators (James Montgomery Inc., 1985). The gravel bed flocculator is compact and offers good process control since the head loss across the gravel bed can be controlled (Bhole, 1993). They are also applicable over a wide range of flow rates (Schulz et al., 1994).

Gravel bed flocculators are simple and low cost. They have proven to be effective for small water treatment plants (Kardile, 1981) and low cost package treatment plants (Bhole, 1981) in India. They are also being used in modular plants in Latin America (CEPIS, 1982). The main drawback of gravel bed flocculators is fouling by the deposited particles (Schulz and Okun, 1984). Therefore, for long operation the deposition within the bed should be limited, and provision of sludge removal facilities (such as backwashing) are an important design criterion.

The head loss across the gravel bed induces the velocity gradient necessary for flocculation. The theory of flocculation in a gravel bed is discussed in Chapter 3. There the expression for G in a typical gravel bed flocculator ($5 < Re < 100$) is derived and it is rewritten as

$$G = \left[\frac{0.115(1 - \epsilon)^{1.34}}{\nu^{0.45} \epsilon^2 d_s^{1.34}} \right] v^{1.48} \quad (3)$$

The two important control parameters for flocculation are G and Gt . As seen from eqn. 3 the G value mainly depends on the grain size and flow velocity. For a particular G (and flow rate), one way to obtain the desired Gt value is by adjusting the detention time (i.e. bed length). The grain size in gravel bed flocculators varies over a wide range from 2 mm to 75 mm in practice and in research (Bhole and Mhailaskar, 1977; Mishra and van Breemen, 1978; Bhole and Potdukhe, 1983; Ayoub and Nazzel, 1988). Commonly used are grain sizes of 10-20 mm and with flow rates of 6-20 m/h; this operational area is shown in Fig. 2. The depth of the bed varies from 1.5 to 3 m (Schulz and Okun, 1984). The G and Gt values range 10-100 s^{-1} and 2,500-30,000, respectively. The gravel bed flocculator is a plug-flow reactor which is more efficient in flocculation than a continuously stirred tank reactor (CSTR) (Harris et al., 1966): short-circuiting of flow is minimal and the flocculation time

can be considerably reduced. 3 to 5 min flocculation time in a gravel bed typically is equivalent to 15 min in a jar test and 25 min in a single compartment flocculation basin (Wagner, 1982; Schulz and Okun, 1984).

The different regimes flow (i.e. laminar, steady inertial, turbulent transition and fully turbulent) in a granular bed as is function of Reynolds number Re and was investigated by Wright (1968) (also see Chapter 3). The author defined Re as

$$Re = \frac{0.6 d_s v}{\mu(1 - \varepsilon)} \quad (5)$$

The factor 0.6 taken following Rose (1945) to make the porosity expression equal to unity when $\varepsilon = 0.4$.

Working definitions of processes within a granular media

Several processes such as flocculation, sedimentation and filtration can occur simultaneously within a granular bed. The granular medium is generally termed after the dominant process i.e. as a flocculator or a filter. In a gravel bed flocculator some particles may still be retained along the bed length by sedimentation or filtration. On the other hand, during deep bed filtration in-pore flocculation also occurs which assists in subsequent better removal (Willson et al., 1980; Graham, 1986, 1988).

Sedimentation in a gravel bed and gravel bed flocculation may again fall under the terminology of filtration as they are also involved in "separation of suspended solids by passing it through a porous medium". The particle transport to the grains is mainly achieved by one of the filtration mechanisms, namely sedimentation. Therefore, they can be considered as a special case within the general filtration process.

In a settling tank where particles are removed by sedimentation. Here differential flocculation and also flocculation due to G caused by the residual turbulence at the inlet and by the drag on the walls and floor occurs (Camp, 1946). When the gravel bed is used as a settler, sedimentation is the dominant process mechanism as determined by the media and the suspension characteristics. Nevertheless, a limited degree of flocculation can also happen. Sedimentation within a gravel bed and sedimentation in a settling tank are both initiated by the same force - gravity.

The mathematical expression for the sedimentation mechanism in filtration is the same as that in a sedimentation tank (following Stokes' law) after necessary correction for local velocities near the grain surface (Ives, 1975a). For particle size larger than $1 \mu\text{m}$ sedimentation becomes increasingly dominant transport mechanism in a filter. Analysis of flow equation around a filter grain indicates that the tangential velocity rapidly diminishes to zero at the surface (Ives, 1975a). This low-velocity region around a grain surface enhances the particle transport towards the grain surface whereas in the bulk of the fluid the particle transport is similar to a settling tank. Larger grains have lower surface area (and lower low-velocity region around it) per unit filter volume, the effect of which can be low and thus the particle transport is comparable to a settling tank; particle separation in larger grains is modeled in

analogy to a multiple-plate settler (Chapter 3). However, smaller grains have increased proportion of low-velocity region and thus particle separation will be higher than predicted by the model.

Therefore, in gravel beds usually both the terminologies and the processes can not be sharply separated. An overlap exists between the flocculation, sedimentation and filtration. Also, no clear definition or demarcation was encountered in literature regarding these processes. The following working definitions are used in this study for the different processes. The present investigation is limited to coagulated highly turbid suspensions (200 NTU). However, the concepts and criteria can be extended or modified, if necessary, to other water types.

Deep Bed Filtration: The particle removal along the bed length follows the exponentially declining relationship as proposed by Iwasaki (eqn. 2). The dominant particle transport mechanism may or may not be sedimentation, and other filtration mechanisms (such as diffusion, interception, etc.) can be effective. During filtration in-pore flocculation may occur.

Sedimentation (in gravel bed): The dominant particle transport mechanism is sedimentation and particle settling onto the different layers of gravel surface is similar to that in a multiple plate settler. Gravel bed flocculation can also occur. Hence, the removal efficiency for various flow rates and grain sizes can be predicted by the sedimentation model developed for similar conditions (Chapter 3).

Gravel Bed Flocculation: The gravel bed is considered to have characteristics that can be described by a conventional flocculator in terms of G and t . The limits of G are taken as recommended by Camp i.e. G is preferably between 10 and 100 s^{-1} . The desired detention time t , may be obtained by adjusting the bed length. For practical purpose (long operation time), the amount of particle deposition should be limited. An additional criterion of limiting particle deposition to less than 25% per meter (or $\lambda < 0.29 m^{-1}$) has been considered for a feasible gravel bed flocculator operation.

MATERIALS AND METHODS

The highly turbid raw water of 200 NTU was prepared from presettled kaolin suspension. The specification and procedure to prepare the kaolin suspension and experimental set-up is described in Chapter 2.

Series of standard jar tests were carried out to study the coagulation process. The jars were 2 litre pyrex beakers, and are mixed by a rectangular flat blade (75 x 25 mm) driven by an adjustable speed motor. Aluminium sulphate stock solution was prepared before each sequence of experiments by dissolving 2.5 g analytical grade $Al_2(SO_4)_3 \cdot 18H_2O$ (Merck) per litre. However, for convenience in comparing with Fig. 1 and other similar studies, the aluminium sulphate concentration is converted and hereafter expressed as $Al_2(SO_4)_3 \cdot 14.3H_2O$ (alum). In all coagulation experiments, alum was added directly from the stock without intermediate dilution.

The jar test procedure was performed similar as that of the earlier researchers on the coagulation diagram (Amirtharajah and Mills, 1982; Dentel and Gossett, 1988). The required

equilibrium pH (after alum addition) was adjusted by adding predetermined amounts of 0.1 M NaOH or 0.18 M HCl to the raw water suspension before alum addition. Then the solution was rapidly mixed at $G = 200 \text{ s}^{-1}$ for 1 min. It was then flocculated for 20 min at $G = 15 \text{ s}^{-1}$, and allowed to settle for 15 min. Samples were drawn from a level 30 mm below the liquid surface and were analyzed for turbidity (Dr. Lange Trübungssphotometer LTP4, Germany). Electrophoretic mobility EM (REPAP zeta potential meter from Tom Lindstrom AB, Sweden) was measured after 20 min flocculation and also after settling. The EM and residual turbidities were determined for different aluminium dosage [0-30 mg Al(III)/l] and pH values (5-9) to cover a wide spectrum of the parameters in order to ascertain the extent of different mechanisms involved.

The turbidity and EM based criteria for demarcating different zones of coagulation mechanisms are specified in Table 2. 2 m length filter beds consisting of 20, 13, 8 and 4 mm grain sizes were used for the clarifying process experiments. The water turbidity in all cases was 200 NTU (except when otherwise mentioned). It was coagulated with 1 mg Al(III)/l and with $G = 200 \text{ s}^{-1}$ and $t = 1 \text{ min}$. Details of the experimental set-up and procedure are given in Chapter 2.

Table 2 Criteria for mechanisms in the coagulation diagram (Fig. 5) applied for highly turbid water.

Mechanism	Criterion	Reference
Adsorption-destabilization	EM of the particles/flocs between -1 to +1 $\mu\text{m/s/V/cm}$.	Dentel and Gossett (1988)
Restabilization	EM of the particles/flocs $> +1 \mu\text{m/s/V/cm}$	Amirtharajah and Mills (1982); Dentel and Gossett (1988)
Sweep coagulation	EM of most of the particles/flocs ≈ 0 , only a few particles have EM = -1 to +1 $\mu\text{m/s/V/cm}$. In addition, the residual turbidity $\leq 1 \%$ of initial turbidity (i.e. $\leq 2 \text{ NTU}$).	Amirtharajah and Mills (1982) This study
Optimum sweep coagulation	EM of the particles ≈ 0 , only a few particles have positive or negative EM (-0.5 to +0.5 $\mu\text{m/s/V/cm}$). Maximal reduction of turbidity occurs. In addition, residual turbidity $< 1 \text{ NTU}$.	Amirtharajah and Mills (1982) This study
Combination (sweep & adsorption-destabilization)	Residual turbidity is not as low as in sweep coagulation zone, however, some degree of charge reduction and coagulation occurs. Residual turbidity in jar test $< 10 \text{ NTU}$. This condition would produce acceptable DHRF effluent (3-5 NTU).	Amirtharajah and Mills (1982) This study

RESULTS AND DISCUSSION

Coagulation

Fig. 3a shows the effect of alum concentration on EM for pH 5-9. A more or less linear relationship of EM with alum dose can generally be observed till the neutral EM value. This stoichiometry is thought to be related to the adsorption-destabilization mechanism in that range (O'Melia and Stumm, 1967; O'Melia, 1972). It can be observed that at lower pH the $|EM|$ changes more sharply than at higher pH. The higher sensitivity of EM to alum dosage at lower pH is likely to be related to the higher availability of multi-charged Al cations at low pH. At higher coagulant dose (beyond the dose required to achieve $EM \leq 0$) the EM is slightly positive and generally constant with increasing coagulation dose. Here the dominant mechanism is enmeshment of particles within $Al(OH)_3$ flocs or precipitation of $Al(OH)_3$ onto the surface of original particle (Alaerts and Van Haute, 1981; Chowdhury and Amy, 1991). The particle surface coating with $Al(OH)_3$ determines the surface potential and the EM which, therefore, does not change due to further addition of precipitates. At lower pH (e.g. pH = 6) a higher positive EM than at higher pH (e.g. pH = 8) is also observed. It has been reported that the $Al(OH)_3$ precipitates have increasingly strong positive charge from pH 8 towards pH 6 (e.g. Hayden and Rubin, 1974; Alaerts and Van Haute, 1981; Amirtharajah, 1988).

The data points in Fig. 4a represent the EM values at different alum dose and pH for all jar tests. Here the zones of EM values < -1 , $> +1$ and in between -0.5 and $+0.5$, which correspond to the criteria for different coagulation mechanisms (Table 2) are drawn by interpolation between the measured values. A similar graph for residual turbidities together with contour lines for 1, 2 and 10 NTU is given in Fig. 4b.

According to the criteria in Table 2 and with respect to the contour lines for EM and turbidity (Fig. 4a and 4b), the coagulation diagram for highly turbid water is drawn (Fig. 5) showing zones with different mechanisms. It can be observed that, in general, the boundaries of the zones for different coagulation mechanisms are shifted towards higher doses as compared to those in Fig. 1. The lower boundary of the destabilization zone is around 10-15 mg/l alum whereas for low turbidity it is in the order of 3 mg/l. The lower boundary of adsorption-destabilization and combined (sweep and adsorption) zone are also shifted to a similar extent. The shifting of the lower boundaries towards higher coagulant dose is likely to be related to its stoichiometric relationship with colloidal concentration (O'Melia and Stumm, 1968; O'Melia, 1972; Alaerts and van Haute, 1982). The upper boundaries of these mechanisms are also shifted towards higher coagulant dose (about 50%) due to the presence of high colloidal concentration. The shift is, however, not as pronounced as the lower boundary. The stoichiometry between the particle concentration and coagulant dose for adsorption-destabilization or sweep coagulation qualitatively confirms the concept of their chemical interactions.

The residual turbidities for different alum dose at pH 5-9 are shown in Fig 3b. The residual turbidity decreases with increasing alum concentration and remains almost constant for alum concentration > 50 mg/l. These results also indicate that the kaolin particles aggregate even at low alum dose (5-10 mg/l) when they are partially destabilized and still have a net negative charge. At higher alum concentration the solution is increasingly supersaturated and $Al(OH)_3$

precipitates can be formed in the bulk of the solution.

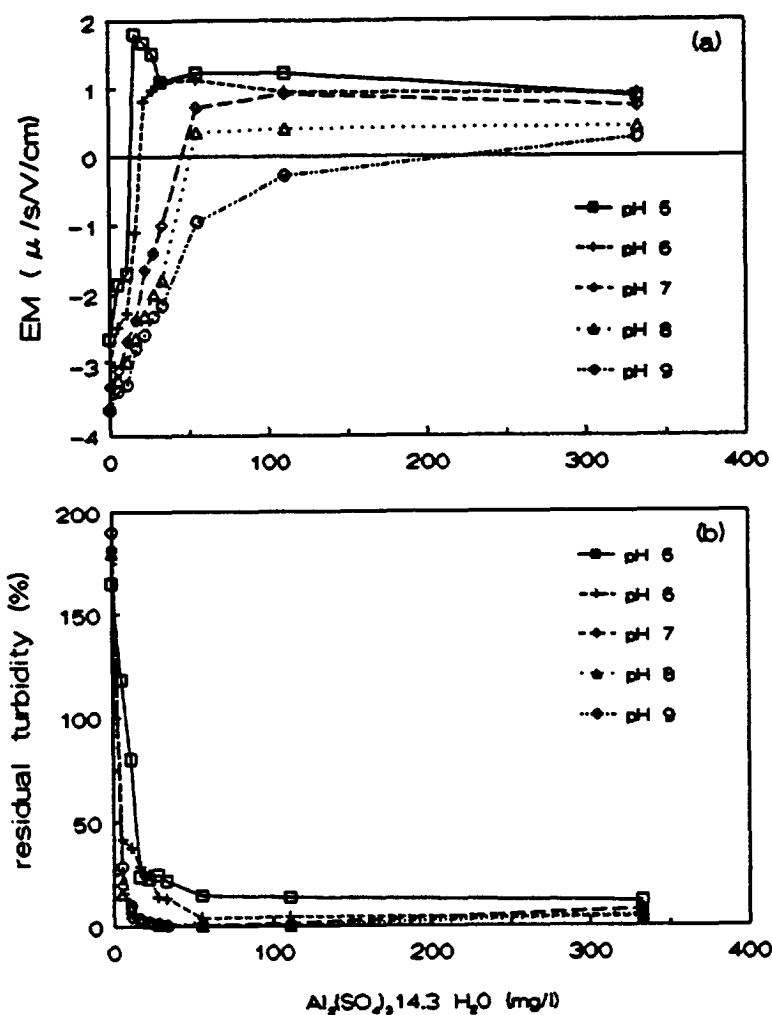


Fig. 3 Relationship of (a) EM and (b) residual turbidity with increasing alum dosage at pH 5-9.

The zones of sweep coagulation and optimum sweep coagulation are in the same order of magnitude for both the higher and the lower turbidities. Here the dominant mechanism appears to be the enmeshment of particles by the $\text{Al}(\text{OH})_3$ precipitates formed in the bulk of suspension when sufficient alum is added to cause oversaturation. Therefore, the colloidal concentration has relatively little influence except for the possible adsorption or precipitation of some additional Al species onto the larger colloid surface before enmeshment. Regarding the combined (sweep and adsorption) zone configuration (and partially the sweep coagulation zone), the boundaries are stretched at the right side beyond the stability line of $\text{Al}(\text{OH})_4^-$. It is commonly believed that the floc formation (for low turbidity water) in this area is poor (Rubin and Kovac, 1974). However, increasing colloid concentration shows a linear relationship with better coagulation performance (Harris et al., 1966). Thus the presence of high colloidal concentration may have supported the coagulation in this zone.

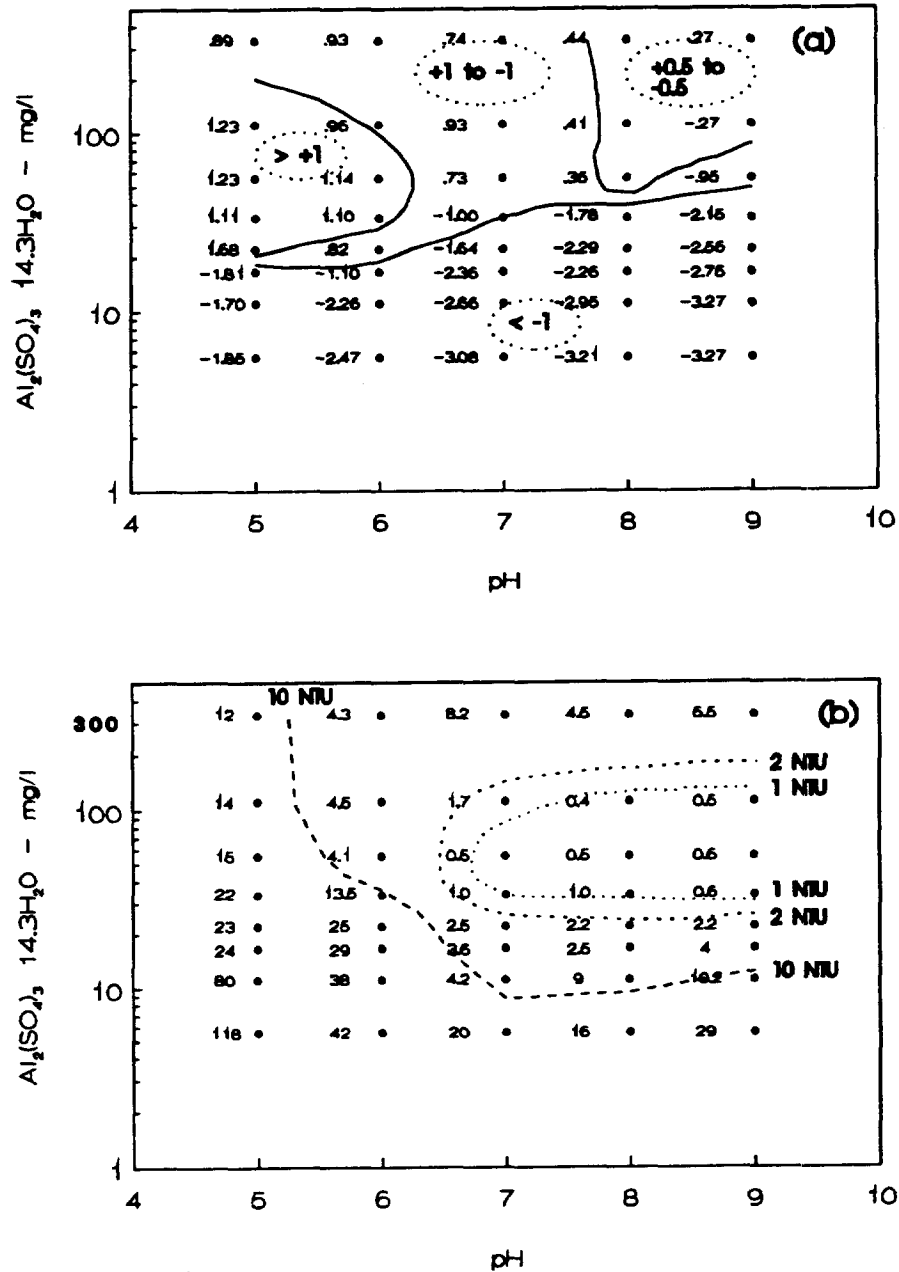


Fig. 4 Data points and corresponding (a) EM and (b) residual turbidity. Different zones of coagulation mechanisms according to Table 2 are also indicated in the graphs. Initial turbidity = 200 NTU.

The lower boundary for combined (sweep and adsorption) zone (residual turbidity < 10 NTU) has also been included given the feasibility of its application in roughing filters. Previous experience showed that this residual turbidity obtained by this particular jar test procedure would produce an acceptable average effluent turbidity < 3 NTU in a DHRF (Chapter 8). The alum requirement for roughing filters is in the order of 10 mg/l and is valid for wide pH range of 6.5 to 9. In a conventional coagulation-flocculation process coagulation is usually performed in the sweep or optimum sweep coagulation zone. In these cases, the

alum requirement is 2 to 3 times higher than that required for DHRF (Fig. 1). Water treatment plants, e.g. in the USA, with raw water turbidity in the range of 200 NTU, are reported to use 35-60 mg/l alum (Hudson, 1973). The coagulation in a DHRF (and possibly in other types of roughing filters also) with high turbidity water, in contrast with conventional flocculation-sedimentation, can therefore be operated along the lower boundary of combined (sweep and adsorption) zone. Thus the coagulant requirement can remain at least 2 to 3 times lower than in the conventional flocculation-sedimentation process. This interesting advantage of DHRF mode is further optimized in Chapter 5.

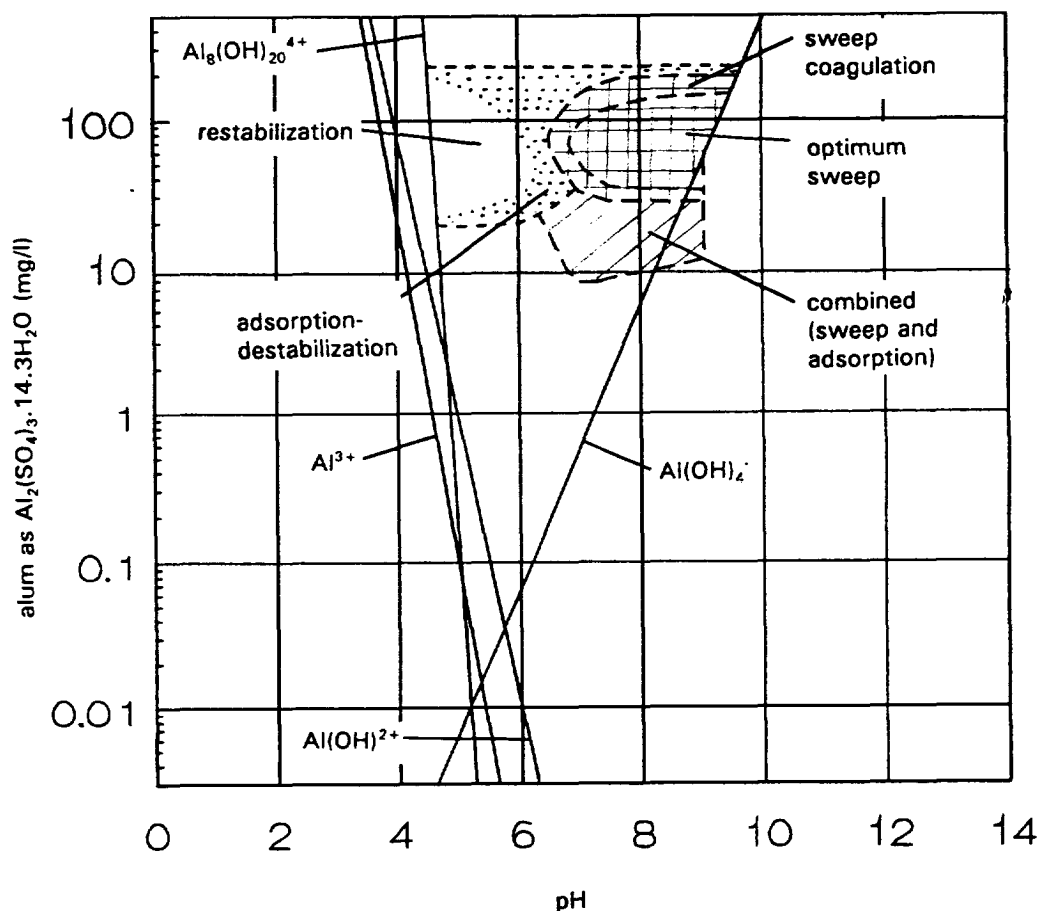


Fig. 5 Design diagram for alum coagulation with high turbid waters. Raw water turbidity = 200 NTU.

Processes in a Gravel Bed

The mean particle size of the coagulated suspension is about $5\ \mu\text{m}$ (Chapter 2). The dominant particle transport mechanism in this study is therefore sedimentation.

Table 3 shows the values of λ (average in the working period) and the respective R^2 for a 2 m bed lengths and for various grain sizes (4-20 mm) and flow rates (1-20 m/h). It can be observed that the kinetic filtration equation (eqn. 2) fits very well ($R^2 = 0.9$ to 0.99) with the filtration results for all values of grain sizes and filtration rates investigated. This suggests

that the concept of deep bed filtration can be extended to higher filtration rates and bigger grain sizes with coagulated suspensions. Again the typical characteristics of filter operation i.e. filter ripening, breakthrough and headloss development was also observed in 20 mm and 8 mm grains sizes in DHRF (Chapter 2).

Table 3 The filter coefficient λ (m^{-1}) with coagulated suspension and for different filtration rates and grain sizes. Figures between parentheses represent the variance R^2 of the fit of the experimental values (number of observation = 11) with the kinetic equation.

Filtration rate (m/h)	Grain sizes		
	8 mm	13 mm	20 mm
1	1.64 (0.97)	1.07 (0.97)	0.71 (0.93)
3	1.43 (0.93)	-	-
5	1.19 (0.97)	0.66 (0.95)	0.35 (0.88)
7	1.06 (0.97)	-	-
10	-	0.58 (0.96)	0.30 (0.90)
15	-	0.26 (0.98)	0.25 (0.92)
20	-	0.22 (0.98)	0.20 (0.98)

HRF (uncoagulated water) with low filtration rate (0.5 to 1.5 m/h) showed that it fit well with the kinetic filtration equation (Wegelin, 1987). Limited investigation was done with high turbidity (200 NTU) uncoagulated suspension on a 2 m horizontal filter bed consisting of the 20 mm grain size (largest in the investigated range). Filtration rates varied 1-20 m/h. Similar good fit was also revealed (Table 4). However, when coagulant is not used but other process conditions similar, λ is about 4 times lower.

In order to compare the λ of roughing filters with that of rapid sand filters investigations were carried with initial raw water turbidity of 20 NTU in a downflow filter (grain size = 0.8 mm; bed depth = 85 cm) and at filtration rate = 5 m/h. In direct filtration mode (without flocculation) coagulated with 1 mg $\text{Al(III)}/\text{l}$ the average λ is 11.6 m^{-1} . When no coagulant is used and other process remaining similar λ is reduced to 3.9 m^{-1} . The use of coagulants in rapid sand filters increases λ to a similar magnitude as in roughing filters. Although λ depends on several factors like water quality (pH, temperature, etc.), particle concentration, charge and size, presence of organic matters or other constitutes the values of these λ for rapid sand filters shows its order of magnitude. λ is e.g. 10-30 times lower,

depending on the grain size, in DHRF than rapid sand filter using coagulants. The lower λ in roughing filters are partly compensated by their longer filter lengths.

Table 4 The filter coefficient λ (m^{-1}) for 20 mm grain size with uncoagulated suspension. Figures between parentheses represent the variance R^2 of the fit of the experimental values.

Filtration rates (m/h)				
1	5	10	15	20
0.170 (0.90)	0.075 (0.94)	0.070 (0.90)	0.060 (0.95)	0.050 (0.96)

λ is an inverse function of flow rate and grain size (eqn. 3). The trend that λ decreases with increasing filtration rate is shown in Fig. 6a for coagulated suspension passed through beds of different grain sizes. A horizontal dotted line representing $\lambda = 0.29 \text{ m}^{-1}$ corresponding to 25% deposition of influent concentration per meter is marked in the graph. This will be used afterwards as reference criterion indicating a limit to the deposition rate in the case of the gravel bed flocculator.

The G value (calculated according to eqn. 4) in function of the filtration rates is shown in Fig. 6b for different grain sizes. A horizontal dotted line representing $G = 10 \text{ s}^{-1}$ is shown in the graph as minimum velocity gradient criterion for effective flocculation. The increasing Reynolds number Re (calculated according to eqn. 5) with increasing filtration rate is shown in Fig. 6c. The boundaries of different flow regimes are also shown in the figure; under the investigated grain size and filtration rate all the flow in the gravel bed was generally in the steady inertial regime. This means that the flow is essentially laminar, however, the relationship of headloss with filtration rate is not linear as in case with laminar flow regime.

The predicted residual turbidity by the sedimentation model for 2 m bed lengths with 20, 13 and 8 mm grain sizes, as well as the experimental values are shown in Fig. 7. It can be seen that for 20 mm grains the model corresponds very well for all filtration rates. For 13 mm grains the model corresponds well (< 5% deviation) up to 10 m/h, but deviates (> 5-20%) with further increase in filtration rate. In 8 mm grains the remaining concentration is lower (5-10%) than the model prediction in all cases. At smaller grain sizes (< 8 mm) the proportion of low-velocity flow in the vicinity of the grain surface compared to the bulk of flow becomes appreciable and increases the removal efficiency. The other filtration mechanisms like intersection, inertia, hydrodynamic action and diffusion are inverse functions of grain size (Table 1); they may also become noticeable at smaller grain sizes. These results in higher removal efficiency for smaller grains than predicted by the sedimentation model.

It had been observed that deep bed filtration as mathematically defined by the kinetic equation is applicable over all the grain sizes and filtration rates investigated. The boundaries of sedimentation and flocculation are, therefore, drawn within the overall spectrum of deep bed filtration. Applying the criteria for filtration, flocculation and sedimentation in a gravel bed in the calculated and experimental results (Fig. 6a, 6b and 7), the boundaries of their

dominant zones in terms of flow rate and grain size can be determined. Accordingly the boundaries of the different zones are outlined in Fig. 8.

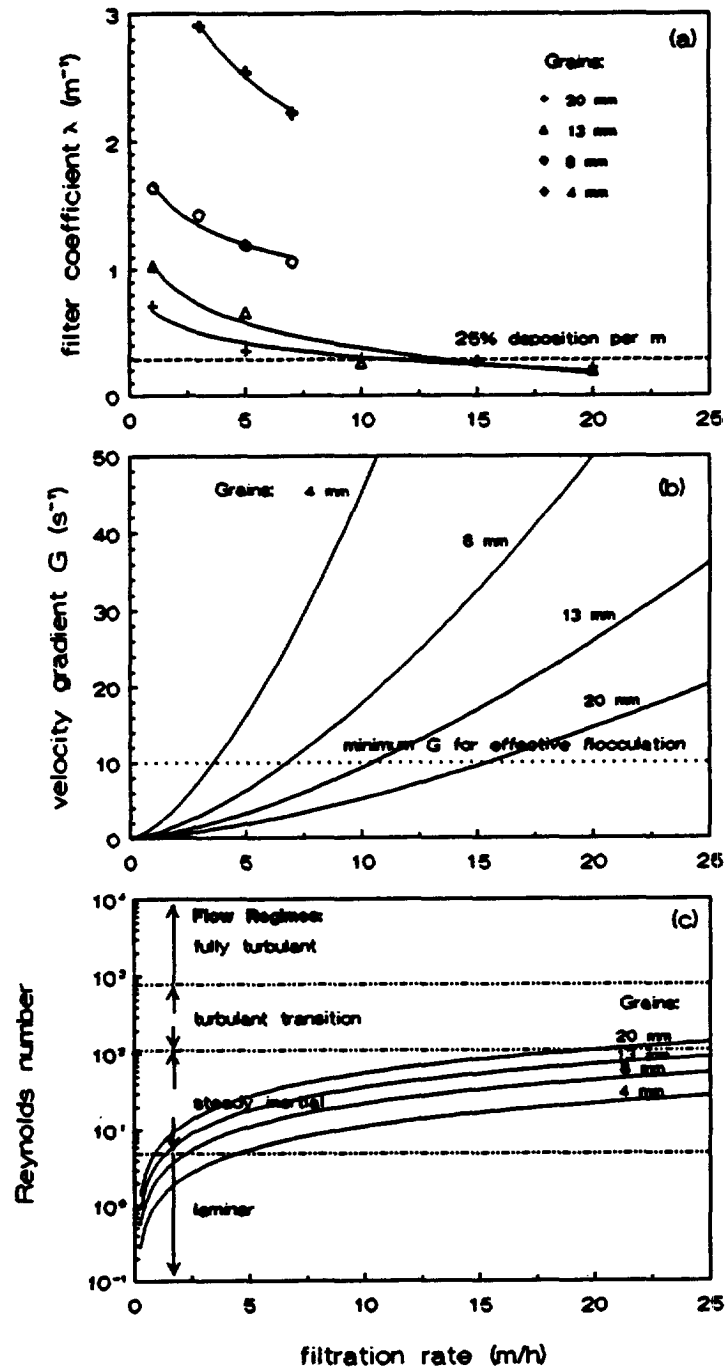


Fig. 6 Gravel bed parameters as a function of filtration rate and grain size: (a) filter coefficient λ ; initial turbidity = 200 NTU, coagulant dose = 1 mg Al(III)/l, (b) velocity gradient G (calculated according to eqn. 4), and (c) Reynolds number (calculated according to eqn. 1); temp = 20 °C.

In the sedimentation zone the residual turbidity can be predicted by the sedimentation model. The lower boundary of sedimentation is derived from Fig. 7 (applicability of sedimentation model) and below which the sedimentation model is not applicable. In the flocculation zone effective flocculation occur with limited deposition. This zone is determined both by the required G value and the amount of deposition. The lower boundary is governed according to the maximum allowable deposition criterion (Fig. 6a). Some experimental points are extrapolated to obtain filtration rates corresponding to the criterion of 25% deposition. The upper boundary is limited by the minimum G value required for effective flocculation (Fig. 6b).

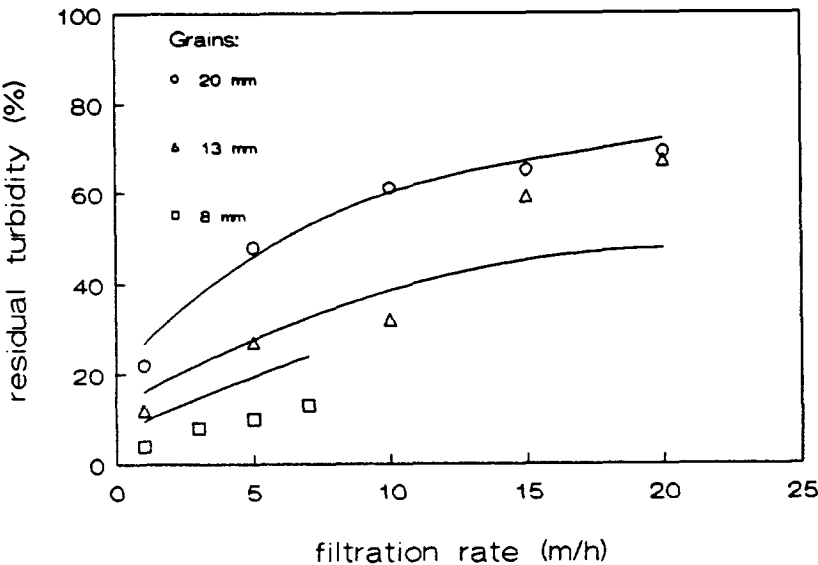


Fig. 7 Residual turbidity as predicted by the sedimentation model (firm lines) and experimental values (symbols) for different grain sizes and filtration rates.

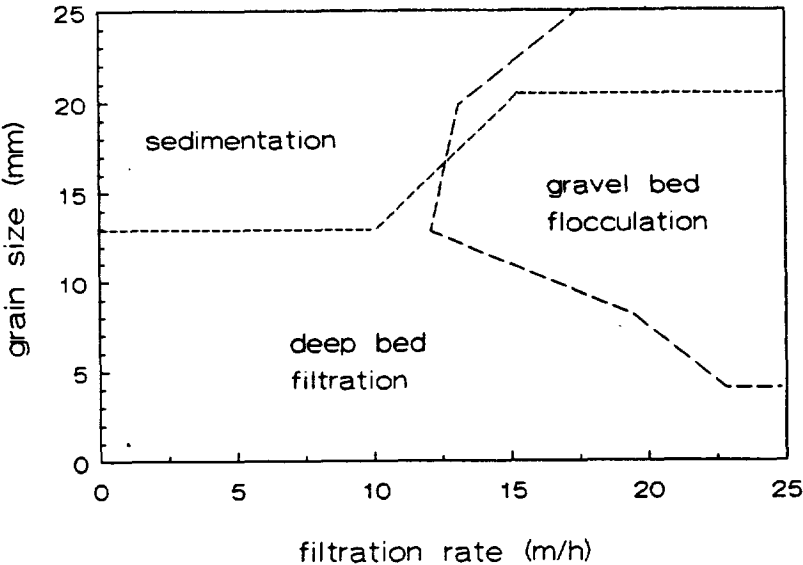


Fig. 8 The ranges of grain sizes and flow rates for optimal application of different granular media

units.

When Fig. 8 is compared to Fig. 2 it can be observed that, as expected, the mechanism in rapid sand filters is conventional deep bed filtration. The dominant mechanism in DHRF, on the other hand, is sedimentation for generally grain size > 13 mm and deep bed filtration for grain size smaller than it. When comparing the favourable zones for gravel bed flocculation in Fig. 2 and the zone in common practice in Fig. 8 it can also be understood that certain design practices such as filtration rate < 12 m/h or combinations of grain sizes > 10 -15 mm and filtration rate < 12 m/h should be omitted. They do not generate enough velocity gradient or the deposition rate is too high to achieve a feasible operating time. The combinations of larger grain sizes (15-20 mm) and higher filtration rate (< 15 m/h) can produce good gravel bed flocculator performance. Larger grains are desirable in terms of filter cleaning as they can easily and economically be washed by water from the top (described in Chapter 7).

The flow in gravel bed flocculator is generally in steady inertial regime (Fig. 6c). Therefore, the mechanism of flocculation is not due to turbulence but, for the combined effect of (i) velocity gradient in the filter pores, (ii) change of flow direction along its tortuous flow paths - the effect being similar to flocculation in coiled pipes (Elmaleh and Jabbouri, 1991), and (iii) contact opportunities between particles and grain surface.

The boundaries for the process mechanisms of roughing filters which usually use uncoagulated waters are not elaborately covered in this study. The proportion of colloidal particles is higher in uncoagulated waters and hence the diffusion mechanism in filtration can be expected to be important in addition to sedimentation (Chapter 2). Therefore, it can be predicted that the boundary between the zones for dominant filtration and sedimentation in the roughing filters will shift towards larger grain size as compared to the case with coagulated water.

CONCLUSIONS

An integrated alum coagulation diagram for highly turbid water (200 NTU) was developed in which the zones of the dominant coagulation mechanisms are demarcated. These zones are, in general, shifted towards higher coagulant dosage as compared to the case of low turbidity water.

Roughing filters for pretreatment of highly turbid waters can preferably operate along the lower boundary of the zone of combined sweep and adsorption coagulation which covers a wide pH range from 6.5 to 9 at low coagulant dose (1 mg Al(III)/l). The coagulant dose required for roughing filters is approximately 2 to 3 times lower than for the conventional flocculation-sedimentation process. Thus, considerable saving in chemical cost and sludge handling can be achieved in roughing filters.

The use of coagulant increases the filter coefficient to about 4 folds in roughing filters as well as in rapid sand filters.

The deep bed filtration kinetic equation of Iwasaki is also applicable over a wider range of

grain size (up to 20 mm) and at higher filtration rates (up to 20 m/h) under the conditions of this study, and thus cover the DHRF operation. This suggests that the deep bed filtration concept can be extended to gravel bed with coagulated and uncoagulated water, at least to the range investigated.

Boundaries of flocculation, sedimentation and filtration processes in a granular bed has been demarcated in terms of grain size and filtration rate (Fig. 8). This diagram suggests ways for optimal utilization of the granular bed.

The feasible zones a gravel bed flocculator have been specified also in terms of grain size and filtration rate. Some common practice of using low filtration rates (< 12 m/h) or combinations of smaller grains (< 10 -15 mm) and higher filtration rates (> 12 m/h) is inappropriate as they do not generate required velocity gradient or clogs too frequently. Satisfactory gravel bed flocculator performance can be achieved by the combination of bigger grains (10-20 mm) and higher filtration rates (< 12 m/h).

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NOTATIONS

- C_0 = initial particle concentration, NTU
- c = constant, -
- d_g = diameter of filter grain, m
- d_p = diameter of particle, m
- DHRF = direct horizontal-flow roughing filtration
- G = velocity gradient, s^{-1}
- EM = electrophoretic mobility, $\mu m/s/V/cm$
- HRF = horizontal-flow roughing filtration
- K = Boltzman's constant (1.38×10^{-23} joules/ $^{\circ}k$)
- L = length, m
- NTU = Nephelometric Turbidity Unit
- T_k = temperature, $^{\circ}k$
- v = approach velocity, m/h
- $\alpha, \beta, \gamma, \delta$ = positive exponents in eqn. 3.
- α_0 = particle-to-grain attachment efficiency, -
- ε = gravel bed porosity, -
- λ = filter coefficient, m^{-1}
- η = single collector efficiency i.e. the rate at which the particles strike the collector divided by the rate at which particles flow towards the collector, -
- $\eta_D, \eta_G, \eta_H, \eta_I$ = η with respect to diffusion, sedimentation, hydrodynamic action and interception respectively, -
- η_0 = total single collector efficiency ($= \eta_D + \eta_G + \eta_H + \eta_I$)
- ν = kinematic viscosity, m^2/s

APPENDIX I Reactions and equilibrium constants for hydrolysis of Al(III) (Dental and Gossett, 1988)

species	reaction	equilibrium constant
Al^{+3}	$\text{Al}(\text{OH})_3(\text{s}) + 3\text{H}^+ \rightarrow \text{Al}^{+3} + \text{H}_2\text{O}$	$\log K_{\text{so}} = 9.15$
AlOH^{+2}	$\text{Al}^{+3} + 1\text{H}_2\text{O} \rightarrow \text{Al}(\text{OH})^{+2} + 1\text{H}^+$	$\log \beta_1 = -4.97$
$\text{Al}(\text{OH})_2^+$	$\text{Al}^{+3} + 2\text{H}_2\text{O} \rightarrow \text{Al}(\text{OH})_2^+ + 2\text{H}^+$	$\log \beta_2 = -9.30$
$\text{Al}(\text{OH})_3^0$	$\text{Al}^{+3} + 3\text{H}_2\text{O} \rightarrow \text{Al}(\text{OH})_3^0 + 3\text{H}^+$	$\log \beta_3 = -15.0$
$\text{Al}(\text{OH})_4^-$	$\text{Al}^{+3} + 4\text{H}_2\text{O} \rightarrow \text{Al}(\text{OH})_4^- + 4\text{H}^+$	$\log \beta_4 = -21.7$

Chapter 5

Direct Horizontal-flow Roughing Filtration

Part I: Optimization of Process Parameters¹

ABSTRACT - The optimum process parameters of direct horizontal-flow roughing filtration are optimized with lab-scale pilot-plant. The parameters investigated are G and t in rapid mixing unit, coagulant dose and pH, filtration rate and, grain sizes and bed length of the second compartment.

INTRODUCTION

Horizontal-flow roughing filtration (HRF) has been proposed as a viable pretreatment process prior to slow sand filters in developing countries (Schulz and Okun, 1984; Wegelin et al., 1987; CINARA, 1993). They are typically 8-12 m long and divided into 3 or 4 compartments with the first containing coarse grains (15-25 mm diameter) followed by compartments with increasingly finer (down to 3-5 mm diameter) grains. The design guideline for HRF based on laboratory tests suggests that for influent turbidity up to 200-400 NTU and filtration rate 0.5-1 m/h, the filtrate quality is 3-5 NTU (Wegelin, 1986). However, field experiences show that the removal efficiency is 80-90%. For high turbidity influent (200 NTU) effluent can remain as high as 50 NTU (Basit and Brown, 1986; Wegelin, 1991; IWACO, 1991). Other major limitations of HRF are that it operates at low filtration rates and consequently needs large filter volumes and leads to higher construction cost. Particle removal efficiency would be also low when influent water contains organic substances and colloidal particles of high stability (see Part II). This makes HRF unattractive for applications in larger towns and more urbanized areas, for unfavourable water types, or when higher removal efficiency is required.

To overcome some of these difficulties the HRF technology has been modified earlier by applying a small constant amount of coagulant, typically 1 mg as $Al(III)/l$, prior to the gravel bed filtration (Ahsan et al., 1991). The modified process is called direct horizontal-flow roughing filtration (DHRF) in analogy with direct filtration (Fig. 1). The coagulated

¹Paper by Ahsan T., Buiteman J. P. and Alaerts G. J. submitted to *J. Water SRT-Aqua*.

suspension is then passed on to the gravel bed (approx. 7-9 m long). It is usually divided into 2 compartments. Filters are recommended to be cleaned by surface washing (Ahsan, 1995). Construction cost of DHRF would be 25-35% less and operating cost about half than conventional flocculation sedimentation pretreatment process. It can also be applied prior to rapid sand filters and has a good potential to replace conventional pretreatment process in small and medium towns in developing countries.

Lab scale pilot plant studies showed that on the same water DHRF systematically performed better than HRF in terms of higher filtration rates and hence smaller filter dimension, better effluent quality, and easier sludge removal (Ahsan et al., 1991). The turbidity removal profiles along the length of the gravel bed indicate that the first gravel bed compartment functions as combined flocculator and clarifier; the subsequent gravel bed compartments with finer grains exhibit the characteristics of deep bed filters (Ahsan et al., 1991).

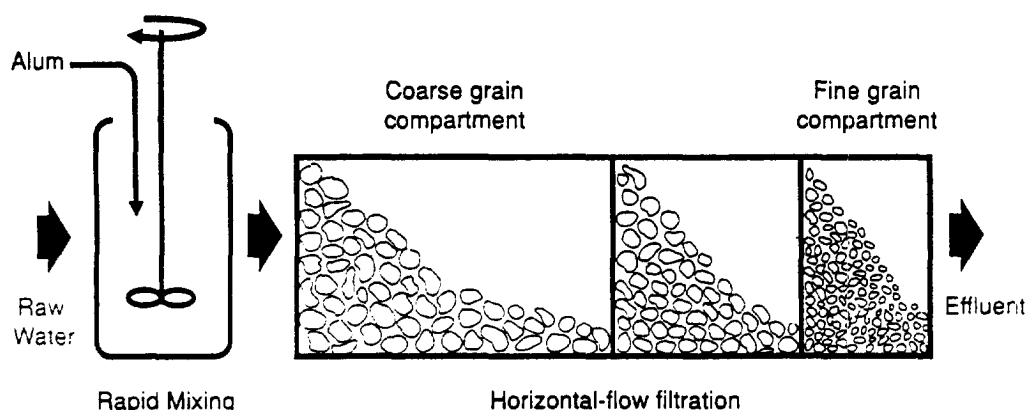


Fig. 1 Schematic diagram of DHRF with 2 compartments.

The physical layout of DHRF is divided into (i) the rapid mixing unit, and (ii) the roughing filter unit. In water treatment rapid mixing has two functions: coagulant dosing and rapid homogenization of the liquid (physical process), and the coagulation and initiation of flocculation (physio-chemical process). Coagulant is added to the raw water in the rapid mixing unit to facilitate particle separation in the subsequent granular media. Proper mixing ensures effective utilization of the coagulant added. Flocculation further takes place in the filter compartments and is determined by the physio-chemical and hydraulic conditions in those compartments.

Coagulation of suspended solids with Al salts is generally accomplished by adsorption of

oppositely charged hydrolysed coagulant species on the colloids and subsequent destabilization (adsorption-destabilization), enmeshment of the colloids within hydrolysed hydroxide precipitate (sweep coagulation), or both (Alaerts and van Haute, 1982). An optimal coagulation process which can form aggregates of appropriate physical and chemical properties is essential in water treatment. Amirtharajah and Mills (1982) developed a comprehensive design diagram for alum coagulation to ascertain the optimum coagulation conditions in terms of coagulant dose and pH for relatively low turbidity water (25–40 NTU). A similar coagulation diagram for high turbidity water (200 NTU) and DHRF application had been developed by Ahsan (1995) with the help of jar tests (Fig. 2); the zones of dominant coagulation mechanisms are demarcated. The feasible area of DHRF coagulation is argued to be in the combination of sweep and adsorption mechanism zone.

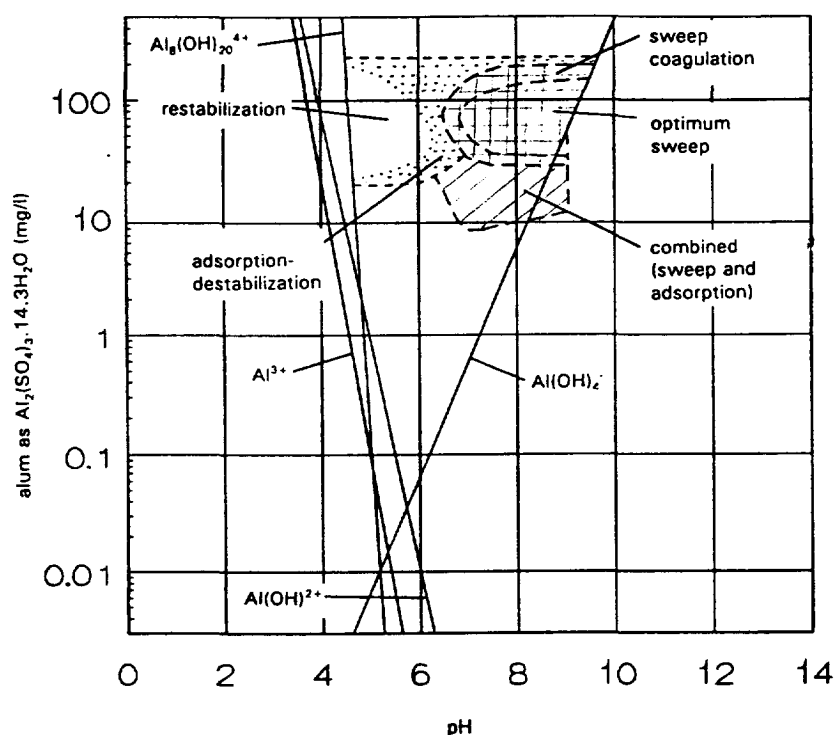


Fig. 2 Design diagram for alum coagulation with high turbid waters. Raw water turbidity = 200 NTU ($11.1 \text{ mg Al}_2(\text{SO}_4)_3 \cdot 14.3 \text{ H}_2\text{O} = 1 \text{ mg Al}$).

The rapid mixing stage is reported to be possibly the most important operation in the treatment process since here the destabilization reactions occur and micro-flocs are formed (Dentel and Gossett, 1988), the characteristics of which markedly influence the subsequent flocculation kinetics and other removal mechanisms (Guibai and Gregory, 1991). The rapid mixing process is governed by the velocity gradient and the detention time t (s) in the rapid mixing unit. The root mean square velocity gradient for turbulent condition G (s^{-1}) proposed

by Camp and Stein (1943) is the parameter widely used to describe the velocity gradient during rapid mixing (Cleasby, 1984).

The optimum detention time in the rapid mixing unit depends for a given raw water also on other factors e.g. G value, turbulence intensity at the dosing point, coagulant type, etc. (Kawamura, 1976). Amirtharajah and Mills (1982) explained the intensity of rapid mixing and time required in terms of different mechanisms of alum coagulation and showed that high intensity rapid mixing with short detention time ($G = 16,000 \text{ s}^{-1}$, $t = 1 \text{ s}$) is desirable in the adsorption-destabilization zone whereas in the sweep coagulation zone the mixing conditions are more indifferent. In DHRF, the appropriate coagulation is preferably performed in the combination of sweep and adsorption zone. Therefore, the influence of mixing intensity and duration is of importance.

The dominant particle removal mechanism(s) in a filter (e.g. sedimentation or diffusion) may influence the selection of coagulant dose and pH which would produce appropriate particle/floc characteristics and overall floc-forming and -removal conditions. High coagulant dose will assist in better particle removal by sweep coagulation but, on the other hand will also produce a large amount of floc volume and result in more rapid filter clogging. The downstream process can also set boundary conditions for the rapid mixing conditions. For flocculation-sedimentation the objective would be to produce larger settleable flocs which can be readily removed by sedimentation (e.g. in the first compartment of DHRF). Low G and longer t are then desirable as this produces less strong but bigger and easily settleable flocs. If the downstream process is direct filtration (like in the second compartment of DHRF), high G and short t are suitable as they produce relatively smaller but stronger and 'filterable' flocs (Amirtharajah and Trusler, 1986). Thus, in direct filtration the optimum mixing condition is also dependent on the filter grain size and filtration rate (Yeh and Huang, 1989).

For a particular suspension the main criteria which describe the performance of a filter are (i) removal efficiency, (ii) headloss, and (iii) filter run time. The filtration-related process parameters which largely determine them are (i) grain size, (ii) layer thickness, (iii) combination of grain size and layer thickness and number of layers, and (iv) filtration rate. With smaller grain size the removal efficiency increases but in the same time the headloss increases and the filter run time reduces. The depth of deposit penetration is often also reduced thus limiting the effective utilization of bed thickness. Increasing filtration rate lowers bed area requirement, but may decrease the removal efficiency and filter run time, and increase headloss. The headloss in DHRF increases the water level in the filter bed which has to be compensated by additional bed height, and increases the construction cost. It should, therefore, be minimal (preferably $< 20 \text{ cm}$). The allowable headloss in HRF is 20-30 cm (Wegelin, 1986). The lower removal efficiency in roughing filters with larger grains as compared to conventional rapid sand filters is partially compensates by their longer

bed length and low filtration rate, however, their effectiveness of increased length is limited. In roughing filtration the dominant process mechanism is sedimentation and thus particles $< 1 \mu\text{m}$ are not effectively removed (Boller, 1993), but the flocculation process in DHRF can overcome this limitation.

The particle/floc removal process in DHRF is a combination of conventional flocculation-sedimentation and direct filtration (Ahsan et al., 1991). The DHRF first compartment (typical grain size = 15-25 mm; length = 4-5 m) has been modeled in analogy to a parallel plate settler (Ahsan and Alaerts, 1995a). In order to further optimize the processes in all DHRF compartments, this paper discusses the chemical and physical processes occurring in DHRF with high turbidity water. The specific objectives are to determine:

- the optimum coagulation conditions in terms of coagulant dose and pH,
- the optimum G and t in the rapid mixing unit, and
- the influence of filtration rate and grain size in the second compartment.

MATERIALS AND METHODS

The schematic diagram of the experimental set-up is shown in Fig. 3. The model raw water was a kaolin suspension (kaolin fine powder from Riedel-de Haën, Germany) in Delft tap water ($\text{pH} \approx 8.1$, $\text{Ca}^{2+} \approx 50 \text{ mg/l}$, $\text{HCO}_3^- \approx 130 \text{ mg/l}$, total hardness $\approx 8.9^\circ\text{DH}$). The suspension was passed through a plain sedimentation tank (1 hour detention time) to eliminate coarse particles. The turbidity of this presettled suspension was 200 NTU ($\pm 2\%$) to represent highly turbid water after presettling in a plain sedimentation tank. The suspension was delivered to the rapid mixing unit where a predetermined constant dose of aluminium sulphate $\text{Al}_2(\text{SO}_4)_3 \cdot 18\text{H}_2\text{O}$ analytical grade (Merck) was injected by a volumetric dosing pump. The container of the gravel bed was made of perspex tubes, each 20 cm in diameter and 2 m long. Each tube unit could be connected in series to obtain desirable lengths of bed compartments. Water samples were collected iso-kinetically from the gravel bed at predetermined time intervals at the inlet, outlet and also from sampling ports at each 20 cm of bed length. Sample tubes were 10 mm diameter perspex rods with entrance diameter of 5 mm over 10 mm length then narrowed to 2 mm throughout the rest of the rod. The sampling tubes were placed vertically, and the high flow velocity in the narrow sections prevented any deposition within it. The sampling tubes were gently flushed before sampling. Only 3 to 4 samples were taken at a time so that no substantial reduction in filtration rate occurred.

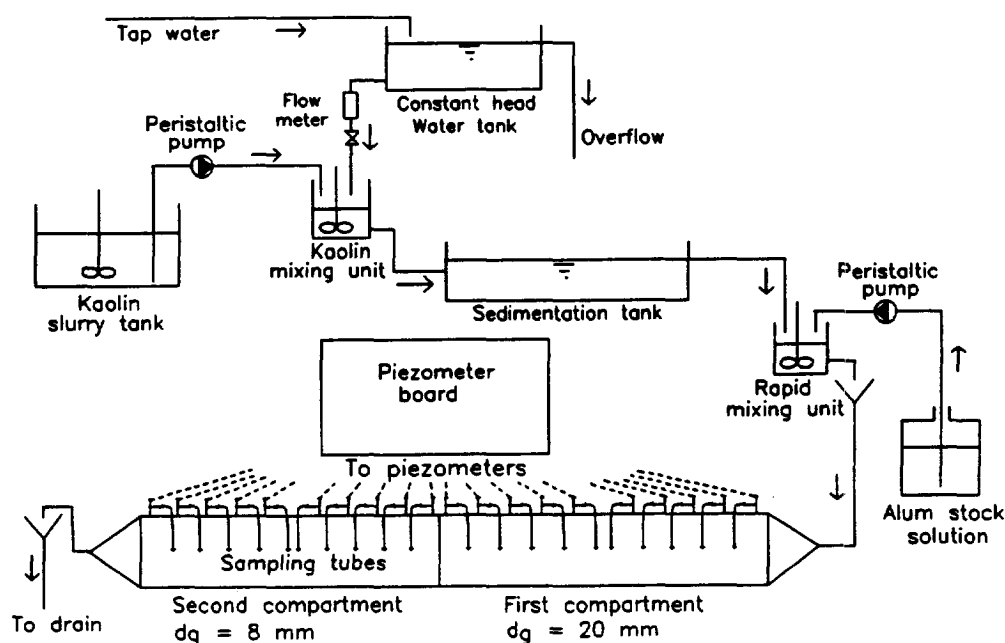


Fig. 3 Schematic diagram of the experimental set-up.

The filter tube diameter to grain diameter ratio (1:10) in the first compartment is lower than usually recommended for rapid sand filters (25-50:1). However, the wall effect is presumed to be low because the headloss in the gravel bed with large grains is so low (1 mm/m bed length) that it is comparable to the headloss along the straighter pathways along the wall. Also, earlier comparative study with a filter larger ratio (20:1) showed similar turbidity profiles. Care was taken during gravel packing to minimize any short-circuiting opportunities near the walls. No settling of gravel packing occurred during the filter run. Visual observation of particle velocity along the walls or top provided semi-quantitative evidence of the absence of short-circuiting along the walls.

Piezometric levels at the corresponding sampling sections were measured through another set of ports connected to the manometer board. Turbidity (Dr. Lange Trübungssphotometer LTP 4) was used as the main indicator of the removal performance and was expressed as NTU (Nephelometric Turbidity Unit). The average turbidity was calculated as the weighted time-average from end of filter ripening period till breakthrough. Filter breakthrough was defined to occur when the effluent turbidity sharply increased.

Fig. 2 suggests that coagulation for DHRF should preferably be achieved near the lower boundary of the zone of combined sweep and adsorption-destabilization (i.e. $Al(III) \approx 1$

mg/l; pH 6.5 - 9.0). To verify the effectiveness of this zone and also that of the sweep coagulation zone ($\text{Al(III)} > 2 \text{ mg/l}$) the coagulant dose was varied (0.5-4.0 mg Al(III)/l) in different runs. pH was kept constant at 8.1 by dosing the required amount of NaOH in the rapid mixing unit. G was 200 s^{-1} and t 1 min in the rapid mixing unit, filter length 8 m (4 m first compartment with grain size 20 mm and 4 m second compartment with grain size 8 mm) and filtration rate 5 m/h. The influence of pH was studied in a second series of DHRF runs by varying the pH from 6.0 to 8.1 and with constant 1 mg Al(III)/l dose.

Preliminary investigation on the mixing time t indicated that a detention time of about 1 min in the rapid mixing unit was technically and economically suitable. In order to assess the influence of G more rapidly a shorter 4 m filter was used initially which composed of a 2 m compartment with coarse size grains (20 mm) and a 2 m compartment with medium size grains (8 mm). G in rapid mixing was varied from 20 to 600 s^{-1} with 1 min detention time. Other process conditions were identical to the previous DHRF runs. Three runs were also performed with the 8 m filter length, and G was varied from 100 to 680 s^{-1} . The sludge volume for different coagulation conditions was measured in a 1 litre Imhoff cone. The simulating suspension (200 NTU) was prepared in a jar test apparatus with different alum dose and pH 8.1; coagulant was rapidly mixed at $G = 180 \text{ s}^{-1}$ for 1 min and then flocculated at $G = 15 \text{ s}^{-1}$ for 20 min. The suspension was then poured into the Imhoff cone and sludge volume was measured after 30 min. However, in case when no coagulant was used the sludge volume measurement was taken after 24 h settling time to achieve comparable suspended solids removal from the suspension. Particle size distribution was measured by HIAC PC-320 (sensor type CMB-60).

The effect of grain size, in the typical range of 4-13 mm in the second filter compartment of DHRF was investigated for filtration rates of 3-7 m/h. The first compartment was 4 m long with grain size was 20 mm, and the second compartment was 4 m. Raw water of 200 NTU was coagulated with 1 mg Al(III)/l .

RESULTS AND DISCUSSION

Coagulation

Fig. 4 shows the effluent turbidity as a function of alum dose. The results of the experiments with different alum dose are summarized in Table 1. Filter breakthrough is taken when there was a sharp increase in effluent turbidity. The average turbidity values were calculated from the end of ripening period till the breakthrough. At coagulant dose $> 0.5 \text{ mg Al(III)/l}$ the effluent turbidity was good to excellent. When dose $> 2.0 \text{ mg Al(III)/l}$ the turbidity is very low (0.2-0.3 NTU), however, filter run time decreases ($< 60 \text{ h}$). At 1 mg Al(III)/l alum dose the effluent turbidity is low (2.1 NTU) and the filter run time longest (95 h). The filter

run time generally decreased with increasing coagulant dose. Headloss in all cases developed linearly and was within a reasonable limit of 20 cm.

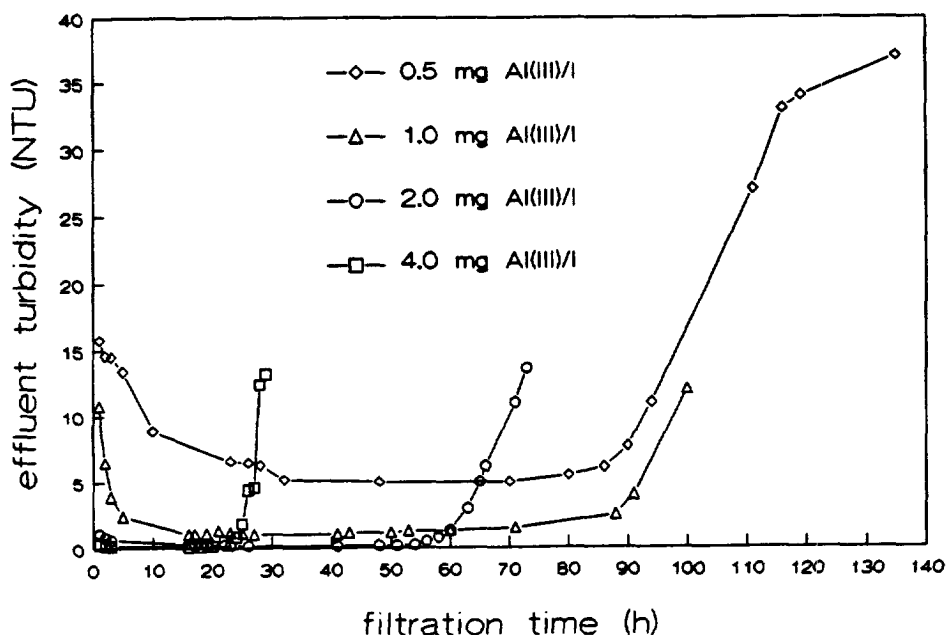


Fig. 4 Effluent turbidity and filtration time as a function of alum dose. Initial turbidity = 200 NTU, pH = 8.1, DHRF filter length = 8 m, filtration rate = 5 m/h.

According to Fig. 2 sweep coagulation occurs at alum dose ≥ 2 mg Al (III)/l and, therefore, excellent particle removal happens in DHRF as the particles are entrapped by the hydroxide precipitate. At 0.5-1.0 mg Al(III)/l dosage the effluent turbidity is higher (2-6 NTU). At still lower coagulant dose (e.g. 0.5 mg Al(III)/l) the particles are insufficiently destabilized resulting in lower filter efficiency.

Deposit volumes were measured in order to determine the relationship of deposit volume with coagulant dose. Hudson (1973) found the deposit volume to be linearly proportional to the coagulant dose. This proportionality was generally observed here (Fig. 5); increase of turbidity with increasing coagulant dose was, however, insignificant ($< 3\%$). Although the processes and floc deposition pattern in a filter and those during a jar test or in an Imhoff cone are different, Fig. 5 is assumed to give a good qualitative relationship between the deposit volume in the filter and the alum dose. The effective floc formation begins at coagulant dose > 0.5 mg Al(III)/l (Fig. 2). Hence the deposit volume did not increase appreciably initially from 0-0.5 mg Al(III)/l. The slope increased linearly from 0.5 to 2 mg Al(III)/l and also linearly beyond 2 mg Al(III)/l but with a marginally different slope. The different slopes may be due to the different coagulation mechanisms involved, namely,

combination of sweep and adsorption, and sweep coagulation, respectively.

Table 1 DHRF performance as a function of coagulant dose. Filter length = 8 m, initial turbidity = 200 NTU, pH = 8.1 after coagulation, filtration rate = 5 m/h.

alum dose	ripening period	filter run time	total headloss	average turbidity after 1st compartment ± sd ¹	average effluent turbidity ± sd ¹
(mg Al(III)/l)	(h)	(h)	(cm)	(NTU)	(NTU)
0.5	10	86	13.7	80 ± 5	6.0 ± 1.0
1.0	5	90	16.5	50 ± 4	2.0 ± 0.6
2.0	2	60	15.9	20 ± 2	0.3 ± 0.1
4.0	<1	24	10.1	10 ± 2	0.2 ± 0.1

¹ standard deviation of turbidity between ripening period and filter breakthrough.

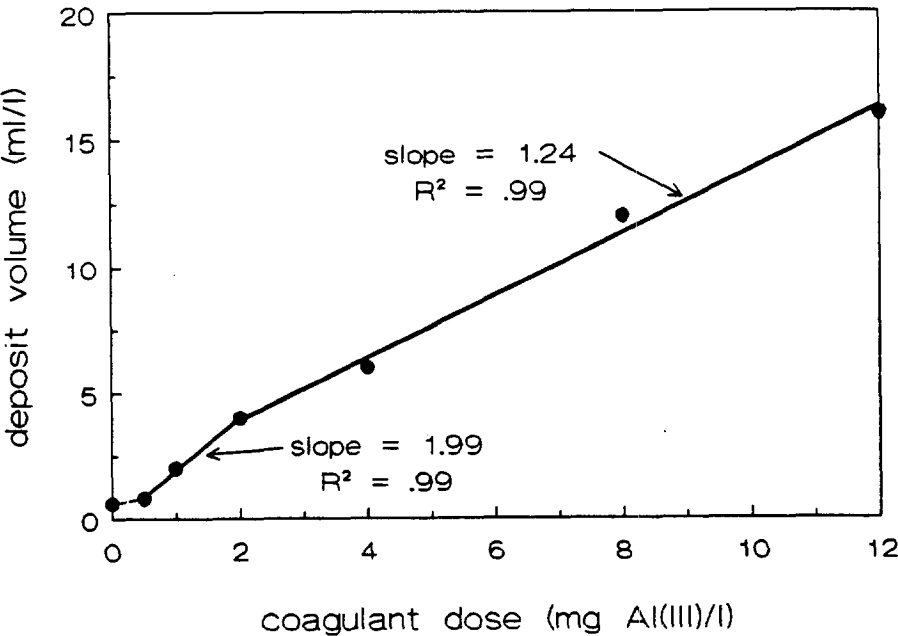


Fig. 5 Relationship of deposit volume as measured in an Imhoff cone with alum dose. Initial raw water turbidity = 200 NTU. The turbidity variation with coagulant dose was negligible (< 2%).

With increasing alum dose the first compartment of DHRF removes a higher portion of turbidity and thus accommodates higher a share of the total deposit (Table 1). The filter run time is supposed to be mainly the function of the deposit volume and the location of deposition (in the first or second compartment). The first compartment with coarse grains has a higher deposit retention capacity before breakthrough than the second compartment (Ahsan et al., 1991). In the case of the filter run with 0.5 mg Al(III)/l about 80% more turbidity load (including Al hydroxide) was passed on to and mostly retained in the second compartment than in the filter run with 1 mg Al(III)/l. However, the deposit volume per unit turbidity in the former case was about half the latter. As the deposit volume (in the second compartment in this case) mainly determines a filter breakthrough, the filter run time of the two filters was similar. The sharp reduction in run time at still higher doses (2-4 mg Al(III)/l) was probably mainly due to the higher deposit volumes in both the compartments.

Filter efficiency improved with increasing filtration time (Fig. 4) as previously captured particles began to serve as additional collector sites in the clean filter for the incoming suspended particles (filter ripening) (O'Melia and Ali, 1978; Tare and Venkobachar, 1985). The filter ripening period decreased with increasing alum dose (Table 1). The higher volume of deposits produced at higher dosages covered the filter grains faster thus creating additional opportunity for particle retention. One way to decrease the ripening period for DHRF operation with low coagulant dose e.g. 1 mg Al(III)/l would be to dose coagulant at a higher concentration during the initial hour(s) of operation. This would allow low turbidity effluent to be fed to the subsequent sand filters from the early beginning of DHRF operation.

Effluent turbidity substantially improved with increase in alum dose from 0.5 to 1.0 mg Al(III)/l as compared to the much more moderate improvement with dosage > 1 mg Al(III)/l. When an effluent turbidity of around 2-3 NTU is desirable, an alum dose of 1 mg Al(III)/l appeared to be optimum. Also, the filter run time was highest (4 days), whilst 3-4 days run time was adequate for an effective filter cleaning procedure (Ahsan, 1995). Further increase in coagulant dose produced lower effluent turbidity, however, it has marginal caused improvements. The reduction of headloss with higher dosages was also marginal. Moreover, such high quality effluent may not be desirable for pretreatment purpose at the expense of higher chemical cost and shorter filter run time. The optimum coagulant dose for direct filtration is reported to be in the range from 0.5 to 1.5 mg Al(III)/l (AWWA, 1980; Janssens et al., 1986; Amirtharajah, 1988). The coagulant dose required for DHRF was in the range of that for direct filtration but was lower than the conventional flocculation-sedimentation process which requires 3-6 mg Al(III)/l for this water type (Hudson, 1973).

The influence of pH on filter performance was investigated at coagulant dose of 1 mg Al(III)/l (Table 2). The variation of effluent turbidity during filtration is shown in Fig. 6. The results in the pH range 6.0-8.1 concurred with the coagulation diagram (Fig. 2):

acceptable effluent turbidity (< 3 NTU) for DHRF could be found along the lower boundary of the zone of combination of the sweep and adsorption. The trend of the results also suggests that the range of this acceptable performance can be extended over the pH range 6.5-9.0 which would also be in accordance with the coagulation diagram. At pH 6.0 the effluent turbidity was higher (6 NTU) and filter run time shorter (55 h). The amount of sludge volume measured was identical ($\pm 2\%$) irrespective of pH.

At the lower coagulant dose (e.g. 0.5 mg Al(III)/l) or pH (e.g. 6.0) insufficient particle destabilization occurred as also found from the coagulation diagram. The DHRF effluent would not have the appropriate properties (e.g. particle charge) for effective removal in the subsequent slow or rapid sand filters. Therefore, such low values may not be recommended except when necessary like in the case of coloured water which requires low coagulation pH (see Part II).

Table 2 DHRF performance as function of coagulation pH. Filter length = 8 m, initial turbidity = 200 NTU, alum dose = 1 mg Al(III)/l, filtration rate = 5 m/h.

pH	ripening period	filter run time	total headloss	average turbidity after 1st compartment \pm sd ¹	average effluent turbidity \pm sd ¹
(-)	(h)	(h)	(cm)	(NTU)	(NTU)
6.0	3	58	10.7	90 \pm 5	6.0 \pm 1.0
7.0	5	74	19.0	50 \pm 4	1.3 \pm 0.8
7.8	5	95	17.2	50 \pm 4	2.1 \pm 0.6
8.1	5	90	16.5	50 \pm 4	2.0 \pm 0.5

¹ standard deviation of turbidity between ripening period and filter breakthrough.

Rapid Mixing

Fig. 7a shows that the turbidity removal efficiency dropped drastically for G higher than 300 s^{-1} in the 4 m long filter. In contrast, the removal efficiency of the 8 m filter was insensitive to G within the investigated G range. The 4 and 8 m filters showed similar response of the

filter run time to G; filter run time increased for G up to 200-300 s^{-1} after which it decrease (Fig. 7b). The total headloss at breakthrough in all cases was low, less than 7 cm for the 4 m long filter and about 17 cm for 8 m long filter. The optimum G value as defined by the best removal efficiency and the longest filter run time was 200-300 s^{-1} .

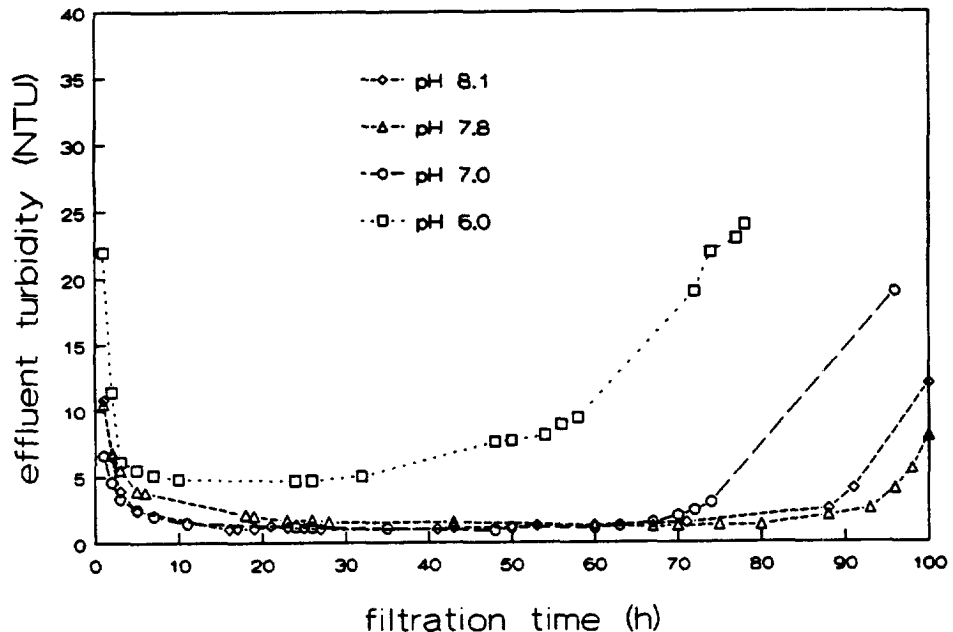


Fig. 6 Effluent turbidity as a function of filtration time and different pH. Initial turbidity = 200 NTU, alum dose = 1 mg Al(III)/l , DHRF filter length = 8 m, filtration rate = 5 m/h.

Below the optimum G range the removal efficiency was almost as good but the filter run time was about 15 % shorter, possibly because coagulants are not uniformly dispersed. Local overdose or underdose may occur or poorer flocs are formed. These flocs tend to be voluminous with low density and floc strength (Twort et al., 1985). During the initial stage of the filter run the amounts of deposits on the grains were low and the large voluminous flocs could easily settle and remain undisturbed causing good turbidity removal efficiency. With increasing filter run time the deposit volume increased causing higher shear stress on the deposits. The voluminous but low strength flocs formed at low G are supposed to detach more easily under mounting stress. In direct filtration the filter run time was also found to be inversely proportional to floc density (Shea et al., 1971; Hutchison and Foley, 1974; Hutchison, 1976). This may partly explain the short filter run time.

In the optimum G range for rapid mixing unit swift dispersion of coagulant is achieved and flocs of an optimum combination of size, strength and density are formed. Increasing the

mixing intensity would produce stronger but smaller flocs. Guibai et al. (1988) also observed that the settling velocity of flocs could be several times lower at above optimum G . Flocs beyond an upper limit formed might be too small for effective removal in the roughing filter; for G above the optimum 200-300 s^{-1} range the filter run time dropped significantly in both filters (Fig. 7b), probably due to poorer floc settling in the first compartment thereby overloading the second. The removal efficiency in a 4 m long filter dropped sharply at higher G . The removal efficiency in an 8 m long filter, however, hardly reduced.

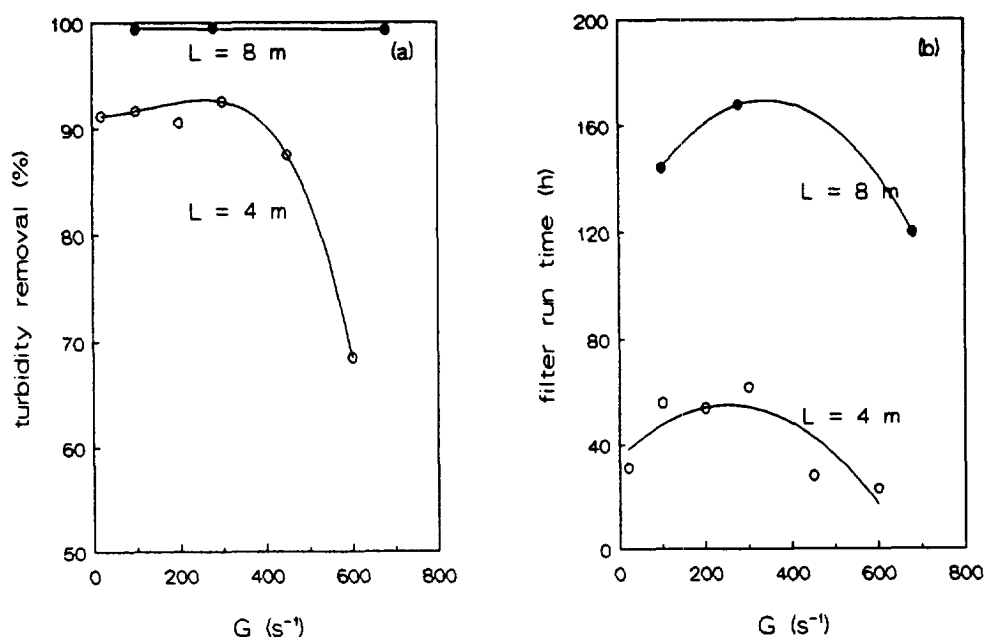


Fig. 7 (a) Turbidity removal and (b) filter run time as a function of G and filter length. Initial turbidity = 200 NTU; alum dose = 1 mg Al(III)/l; pH = 7.8; detention time in rapid mixing unit = 1 min; filtration rate = 5 m/h.

Fig. 8 shows the residual turbidity profile along the 8 m filter length for the three filter runs. The profiles for $G = 100$ and 280 s^{-1} followed each other very closely (deviation $< 2\%$) indicating comparable particle removal along the filter length. The profile for $G = 680$ s^{-1} showed about 10% less particle removal in the first coarse compartment (sedimentation dominated) than in the filter runs with lower G values. This difference was gradually reduced along the length of the second compartment (filtration dominated). The higher removal efficiency in the second compartment in the case of higher G may be due to the effect of the relatively higher initial turbidity at the entrance of the second compartment and to the effect of continued in-pore flocculation of the particles which did not grow to optimum size in the rapid mixing unit and during their travel along the first compartment.

The maximum floc size under agitation condition is reported to be inversely proportional to the power dissipation (Fair and Gemmell, 1964; Parker et al., 1972; Tambo and Hozumi, 1979; François, 1987; Mühle, 1993). This was also observed here (Fig. 9). For $G = 100$ – 300 s^{-1} the median particle size was $9 \mu\text{m}$; it reduced to $5 \mu\text{m}$ at $G = 1000 \text{ s}^{-1}$. It is likely that the larger flocs produced at lower G values (100 – 300 s^{-1}) will be preferentially removed in the filters by sedimentation in the first coarse compartment yielding good overall removal performance. It can also be observed that the particle size distribution are similar for G 100 to 300 s^{-1} (Fig. 6). As a result, the removal efficiencies in both the 4 and 8 m long filters within that G range were similar. In the case of the small flocs generated at high G there was a delay in the formation of sufficiently large flocs (formation partly caused by in-pore flocculation) that can be efficiently removed in the finer filter, as also reported for direct filters (Janssens and Buekens, 1993). This also suggests that a portion of smaller flocs are likely to escape from the short 4 m filter. At longer filter lengths, especially in the second compartment with smaller grains, remaining particles are progressively flocculated producing larger filterable flocs.

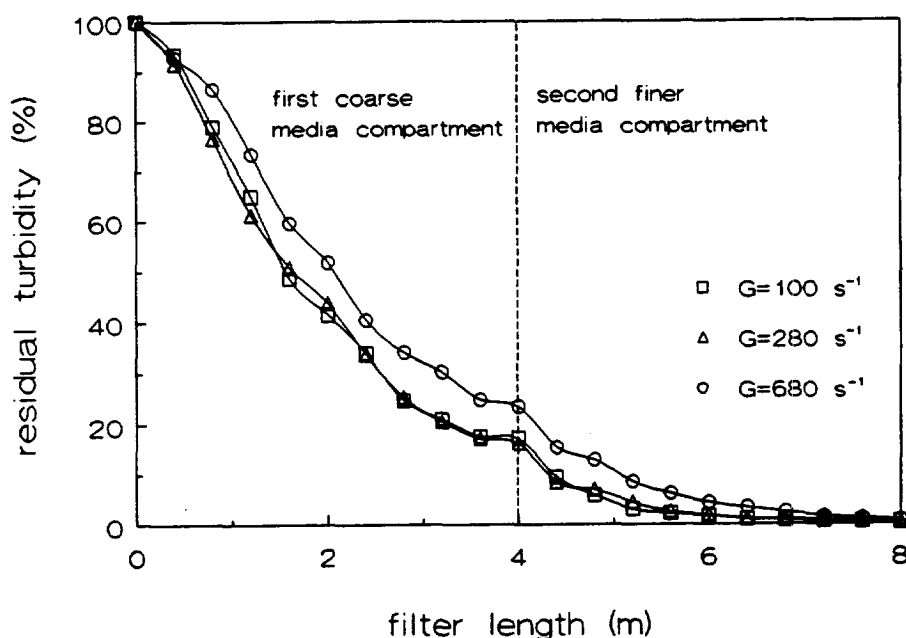


Fig. 8 Residual turbidity profile along the filter length for different G in rapid mixing unit. Initial turbidity = 200 NTU, alum dose = 1 mg Al(III)/l , pH = 7.8, filtration rate = 5 m/h and filter length = 8 m.

Filtration

The results of the filter runs with different grain sizes in the second compartment, and

different filtration rates are summarized in Table 3. Effluent turbidity was generally acceptable for pretreatment (average ≤ 3 NTU). The filter efficiency decreased with increasing grain size (in the second compartment) and filtration rate. In rapid sand filters the filter coefficient is reported to be inversely related to grain size or filtration rate (e.g. Boyd and Ghosh, 1974; Ives, 1975). DHRF with 4 mm grains in the second compartment and applying a filtration rate of or below 5 m/h produced low turbidity effluent of a level as in rapid sand filters (< 1 NTU). An increase in filtration rate appeared to have a larger effect on effluent turbidity increase in the case of smaller grains. With 4 mm grains effluent turbidity increased tenfold from 0.2 to 2.0 NTU as filtration rate increased from 3 to 7 m/h, whereas with 13 mm grain size it increased from 2.3 to 3.0 NTU for the same increase in filtration rate. The filter run time varied from 3-4 days for 7 m/h filtration rate to 10-15 days for 3 m/h. It was generally one third to one fourth shorter at higher filtration rates and smaller grain sizes. The total headloss was reasonably low (< 20 cm) for the 8 and 13 mm grains at all filtration rates but was higher (25-45 cm) for the 4 mm grains. In all cases the headloss development was linear.

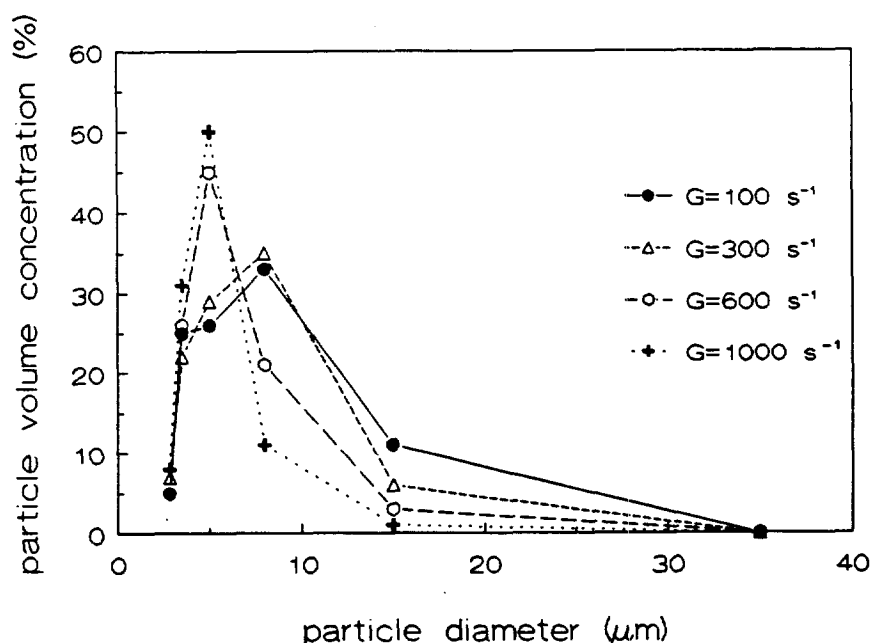


Fig. 9 Particle volume concentration distribution after rapid mixing for different rapid mixing G . Initial turbidity = 200 NTU, alum dose = 1 mg $\text{Al(III)}/\text{l}$, pH = 7.8 and mixing time = 1 min.

The specific deposit before breakthrough (ultimate specific deposit) depends on many factors such as raw water quality, coagulant dose and mixing intensity, grain size, filtration rate, etc. The presence of organic matter tended to reduce the ultimate specific deposit considerably

(see Part II). For the filtration rates of 3-7 m/h the ultimate specific deposit was 20-25 g/l in the first compartment (20 mm grains) and 7-10 g/l in the second compartment (4-13 mm grain size).

Table 3 Results of DHRF runs with different grain size (in the second compartment) and filtration rate. Raw water turbidity = 200 NTU, coagulant dose = 1 mg Al(III)/l, pH = 7.8, first compartment: grain size = 20 mm, length = 4 m; second compartment: length = 4 m.

grain size in 2nd compartment	filtration rate	filter run time	total headloss	average effluent turbidity ± sd ¹
(mm)	(m/h)	(h)	(cm)	(NTU)
4	3	240	25.1	0.2 ± 0.1
	5	90	34.4	0.7 ± 0.2
	7	75	43.4	2.0 ± 0.5
8	3	310	15.2	1.6 ± 0.3
	5	95	17.2	2.1 ± 0.6
	7	75	18.1	2.5 ± 0.7
13	3	350	14.5	2.3 ± 0.6
	5	130	12.8	2.5 ± 0.6
	7	95	12.0	3.0 ± 0.8

¹ standard deviation of turbidity between ripening period and filter breakthrough.

The time-average (after filter ripening till breakthrough) residual turbidity profiles along the second filter compartment for different grain size and flow rate are shown in Fig. 10. Generally the profiles reduced exponentially with the filter length i.e. follows the filtration kinetic (Iwasaki, 1937) and again suggesting that the removal mechanism is deep bed filtration. In the filter with 4 mm grain size turbidity was nearly completely removed in the first 1.5 m filter length, causing more rapid pore clogging and higher headloss in the initial sections. On the other hand filters with 8 and 13 mm grain sizes were characterized by deeper penetration of deposits and better utilization of the bed length. If a combination of smaller grains (e.g. 4 mm) and shorter length (e.g. 1.5 m), which would still produce effluent turbidity < 2 NTU, is used in the second DHRF compartment then the ultimate

specific deposit capacity will be reached within 2 days. Short second compartments are also sensitive to the smaller particle sizes (Fig. 7c). Thus such short time may not be feasible for economic and reliable operation of roughing filtration.

The improvements on effluent turbidity at lower filtration rate e.g. 3 m/h was generally insignificant for pretreatment purpose, however, the filter run time was longer. A filtration rate of 3 m/h produced a long filter run time of 10-14 days. On the other hand, long deposit retention within the filter may not be desirable because of possible microbiological growth and deposit hardening. Roughing filters are recommended to be cleaned at least once a week to avoid anaerobic conditions in the filter and difficulties in hydraulic cleaning (Ahsan, 1995). The cost-optimal design of the process parameters (e.g. coagulant dose, filter bed composition) should be considered in the framework of the total treatment scheme and local circumstances (raw water quality, unit capital and operating costs) (Lawer et al., 1980; Wiesner et al., 1987). This investigation is expected to offer a guideline for selecting the optimum combination.

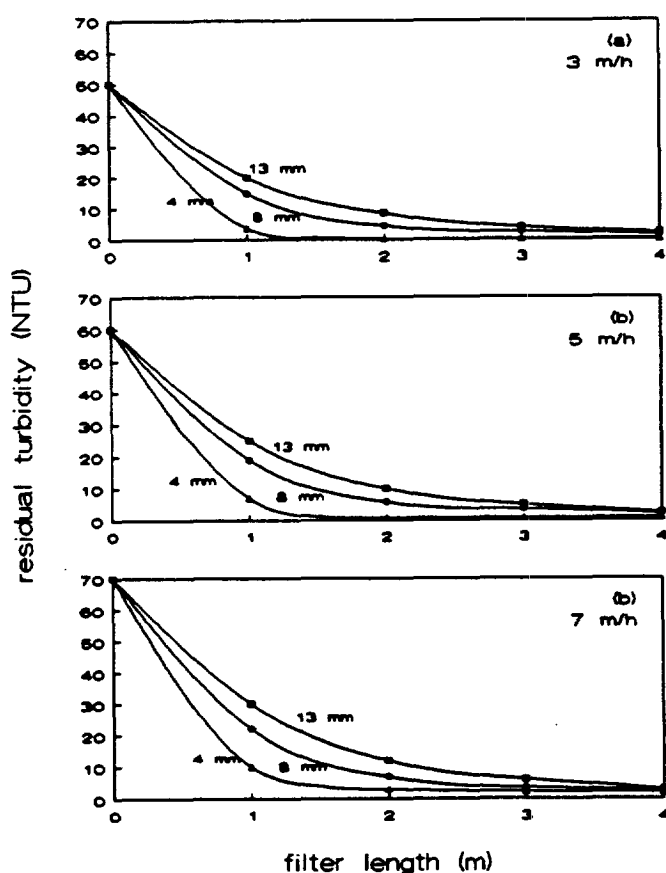


Fig. 10 Average residual turbidity profiles along the second compartment of DHRF with different grain size and at different filtration rate. First compartment: grain size = 20 mm, length = 4 m. Raw water turbidity = 200 NTU, coagulant dose = 1 mg Al(III)/l, pH = 7.8. Filtration rate (a) 3 m/h, (b) 5 m/h and (c) 7 m/h.

CONCLUSIONS

The optimum coagulant dose for DHRF with highly turbid water (~ 200 NTU) was about 1 mg Al(III)/l and was effective over a wide pH range (6.5-9.0). This experimental finding on a filter pilot plant was in accordance to the prediction made earlier by the alum coagulation in jar tests for high turbidity water (DHRF coagulation diagram).

The optimum G value in the rapid mixing unit under the study conditions was 200-300 s^{-1} with 1 min detention time. It is expected that this optimum condition will be valid for a wide of range raw water turbidity.

The effluent turbidity in longer DHRF plants (8 m filter length, 2 compartments) was less sensitive to variations in G in the mixing unit than short plants (4m). The difference in the particle removal efficiency became smaller along the long second compartment filter bed.

Larger grain sizes (8-13 mm) and filter length of 3-4 m in the second DHRF compartment, and filtration rate 5-7 m/h appeared to be suitable for low effluent turbidity (2-3 NTU), low headloss (12-18 cm) and reasonable run time (3-5 days).

Higher coagulant dose (e.g. 2 mg Al(III)/l) with the use of 8 mm grain size in the second compartment, or medium coagulant dose (1 mg Al(III)/l) with smaller grains (4 mm), could produce effluent turbidity (0.2-0.7 NTU) comparable to the treated water from direct filtration or conventional treatment plants. In certain cases of raw water composition DHRF has the possibility to offer a complete treatment scheme for particle separation.

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Chapter 6

Direct Horizontal-flow Roughing Filtration

Part II: Influence of Raw Water Variables¹

ABSTRACT - Direct horizontal-flow roughing filtration (DHRF) was found to be a versatile pretreatment process in handling wide fluctuation in raw water turbidity with operating conditions like coagulant dose, mixing intensity and time, filtration rate, etc., remaining unchanged. In presence of humic substances the turbidity and colour removal reduces drastically in horizontal-flow roughing filter and in DHRF. Satisfactory turbidity and colour removal can be achieved in DHRF by increasing the coagulant dose or decreasing pH, or both. The filter run time in presence of humic substances is however shortened to about half.

INTRODUCTION

River water quality

Rivers are one of the major sources of water supply. Important quality considerations for drinking water are the suspended solids (SS) content, the chemical composition (including natural organic matters) and the bacteriological quality. The most prominent source of SS in rivers is mechanical erosion of rocks and soil which results from combined efforts of various erosion agents, i.e. running water, wind, etc. (WHO/UNEP, 1989). SS consist mostly of clay minerals such as kaolinite, montmorillonite and chlorite, and also quartz, feldspars, oxides and organic matters (Drever, 1988).

The composition of a river water depends on several factors, including (UNICSO et al., 1992): (i) the proportion of surface run-off, sub-surface run-off, and groundwater, (ii) reaction within the river system governed by internal processes, (iii) inputs of pollutants, etc. During flood water quality shows marked variation due to change in proportions of water originating from different sources. Surface run-off is generally high turbid and carries large amount of SS; at high intensity rainfall the river water turbidity increases dramatically. For

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example, the Sabake River in Kenya has occasionally SS content of more than 15,000 mg/l during flood period (Ives, 1983). The Sebou River in Morocco have normal SS content of 2,000 mg/l but 5% of the annual time over 50,000 mg/l (Elmghari et al., 1993). The turbidity of Ganges River in West Bengal, India varies from 600 to 1200 NTU in the monsoon season (Adhikari et al., 1974). The average SS of the Cauca River in Colombia is about 100 mg/l but can rapidly go over 3000 mg/l at the rate of 300 mg per hour (CINARA, 1993).

Bicarbonate (HCO^-) and major cations (Ca^{2+} , Mg^{2+} , Na^+ , K^+) in river water originates from ground water, HCO^- mainly determines the buffering capacity and influences pH. During heavy rainfall the groundwater (flow remains almost constant) is diluted. The HCO^- concentration and buffering capacity reduces, pH goes down, and other properties also changes. The organic matter in soil and waters is composed of a mixture of plants and animal products in various stages of their decomposition together with substances synthesized bio-chemically. Sub-surface run-off mainly carry the organic matter to rivers.

SS cause turbidity and is undesirable in drinking water. More importantly is interferes disinfection during water treatment by creating a possible shield for disease-causing organisms. SS are also considered as a major pollutant carrier (Meybeck, 1982). Many toxic heavy metals, organic pollutants, pathogens and nutrients are also found in SS (WHO/UNEP, 1989). Substantial and rapid changes in water quality make operation and optimization of a water treatment plant difficult. Operators are often not able to respond sufficiently swiftly, and the treatment process can usually not be designed to accommodate water quality that varies more than one order of magnitude. Storage reservoirs or presettling tanks are generally provided to reduce high turbidity levels and dampen sudden changes. However, some (dampened) turbidity peaks may still arrive at the subsequent treatment units. This creates operational problems such as rapid clogging and early breakthrough of filters. The flocculation-sedimentation process is notably affected by the fluctuations in turbidity type and concentration, and colour. In addition, flocculation process also depends on water chemistry variables such as pH, hardness, alkalinity, ionic strength and buffer capacity.

Colour removal

The typical average colour of surface water in U. S. may vary between 5 and 200 mg Pt-Co/l (Narkis and Rebhun, 1977). At higher concentrations it causes a distinct brownish-yellow colour and sometimes noticeable taste and odour. In unpolluted surface water the main cause of colour are humic substances. Humic substances constitute 25 percent of the total world organic carbon budget and about 50 percent of the natural organic matter (NOM) in oceans and fresh waters (Vik and Eikebrokk, 1989; Aiken et al., 1985). Humic substances are often divided into humic, fulvic and humathomelanic acids and their salts based on their

solubility (Rebhun and Lurie, 1993).

Humic substances are the main precursors for the formation of carcinogenic trihalomethanes during water chlorination (Rook, 1976), and have a certain binding potentials for multivalent ions, heavy metals and many pesticides and insecticides (Fettig et al., 1988). The main reasons to remove colour (or humic substances) in water treatment are (Narkis and Rebhun, 1977; Dempsey et al., 1984; Randtke, 1988): (i) removal of associated hazardous compounds and toxic metals which may otherwise escape through the treatment steps into the finished water supply, (ii) removal of biodegradable material that might serve as a substrate for bacterial aftergrowth in the distribution system, (iii) control of taste, odour and for aesthetic reasons, (iv) corrosion control, (v) reduction of organic load to subsequent treatment processes, e.g. chlorination and activated carbon bed, and (vi) minimizing haloform production.

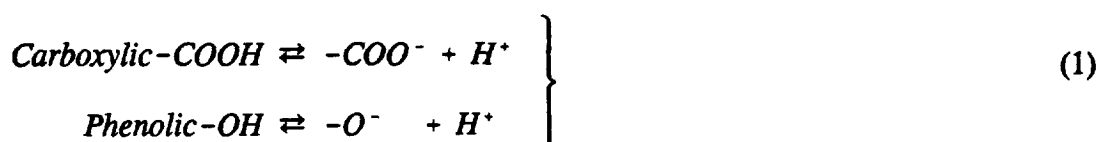
Until now there is no uniform standard for the permissible concentration of humic substances. Some have established standards for parameters which indirectly represent their concentration; the most common lump parameter is colour. The WHO guideline for colour is 15 mg Pt-Co which corresponds to UV absorbance (254 nm) of $6-8 \text{ m}^{-1}$ and TOC value of 1.5-2.0 mg/l (Fettig et al., 1988). The European Commission guide level is 1 mg Pt-Co/l and maximum allowable concentration 20 mg Pt-Co/l. In the USA the standard is formulated as maximum concentration level of 0.10 mg/l for total trihalomethanes (Semmens and Field, 1980).

The presence of humic substances increases the colloidal stability of mineral particles and affects coagulation forces (Narkis and Rebhun, 1975; Tipping and Higgins, 1982; Gibbs, 1983). A thick organic coating of adsorbed humic compounds on the mineral surface is able to separate the colliding particles by a large distance so that the attractive van der Waals forces are not effective any more. The particle-particle attachment efficiency (α_p) was therefore found to be reduced (Jekel, 1986a). Humic substance is also a dominant factor affecting the coagulation process. This is a result of their comparatively high reactivity and number concentration (reacting units concentration) as compared to the number concentration of mineral particles. For the same mass concentration, the number concentration of humic substances is 4 to 6 orders of magnitude higher than that of mineral particles (Rebhun and Lurie, 1993).

Two major mechanisms of humic substances coagulation with Al(III) (or Fe(III)) are recognized (Dempsey et al., 1984; Edwards and Amirtharajah, 1985; Randtke, 1988; Van Benschoten and Edzwald, 1990; Graham et al., 1992; Edzward, 1993; Rebhun and Lurie, 1993):

- i) *precipitation*: charge neutralization and co-precipitation of humic substances with hydrolysed monomeric and polymeric alumino complexes, and
- ii) *adsorption*: adsorption of humic substances on previously formed humic substances-alumino complexes onto the amorphous $\text{Al}(\text{OH})_3$ flocs.

Humic and fulvic acids are polyaromatic compounds with phenolic and carboxylic functional groups. The functional groups can be dissociated depending on pH (Rebhun and Lurie, 1993):



With increasing pH the ionization of the humic substances increase and this in turn increases the concentration of negatively charged functional groups. As pH influences the Al hydrolysis reactions and the extent of humic substance dissociation, the humic substance-Al reactions are, therefore, strongly pH dependent. The two mechanisms of humic substance coagulation were proposed for separate pH ranges (Dempsey et al., 1984, 1985; Dempsey 1989; Vik and Eikebrokk, 1989; O'Melia, 1991; Alaerts et al., 1991): low pH (generally 4-6) for precipitation and higher pH (> 6.5) for adsorption.

The association of humic substances with metal hydrolysed species is faster than mineral particles (Narkis and Rebhun, 1975, 1977; Rebhun and Lurie, 1993). The humic substances easily outcompete the mineral particles to engage in reactions with coagulant. Therefore, relatively higher coagulant doses are required in coloured turbid water to satisfy the humic substances first and then the mineral particles. A stoichiometric relationship between humic substance and coagulant dose to saturate the complexing sites on the organic molecule is reported to exist (Hall and Packham, 1965; Narkis and Rebhun, 1977; Edwards and Amirtharajah, 1985; Jekel, 1986b; Edzwald, 1993).

Poor floc settling and short filter runs are typical for highly coloured water. Alumino-humate flocs are more fluffy, voluminous and fragile than alumino-mineral particle flocs (Rebhun and Lurie, 1993). They also have much lower density than alumino-mineral flocs (Rebhun and Lurie, 1993; Tambo and Watanabe, 1979). Filter deposit and floc strength (expressed as filter deposit detachment constant and floc breakup constant, respectively) were quantified by Rebhun (1990) and Rebhun and Lurie (1993): the deposit detachment constant for

alumino-humate deposit was 15 times higher than for alumino-clay deposits, and the floc breakup constant for alumino-humate flocs was 2 times higher than for alumino-clay flocs. The breakup constant for the alumino-humic flocs was also reported by Bratby (1980) to be 10 times higher than alumino-kaolinite flocs.

The concentration and nature of NOM can (i) control coagulant dose and type, and (ii) can influence the water treatment process i.e. the coagulation and flocculation processes as well as the downstream solid-liquid separation processes (Edzwald, 1993). In many waters NOM effectively controls coagulant dose, rather than the mineral turbidity causing particles (Edzwald, 1993). Many researchers found good NOM removal of at pH range 4.5-7.5 depending on the raw water and process parameters, but general agreement exists on an optimum pH range of 5-6 (Randtke, 1988; Semmens and Field, 1980). On the other hand, the coagulation diagram for highly turbid water (see Part I- Chapter 5) showed that turbidity removal efficiency is less effective below pH 6.5. Therefore, the optimum pH for colour removal may not necessarily be the optimum for turbidity removal, especially in the case of DHRF which is designed for very high turbidity.

It is reported that HRF (i.e. roughing filtration without coagulation) is not effective in colour removal whilst DHRF can be surmised to be capable of removing it. The influence pH was assessed in part I; two other important raw water variables i.e. turbidity and colour is investigated in this paper. The specific objectives are to investigate:

- DHRF performance with change in influent turbidity (as kaolinite), and
- colour removal (as humic acid) in HRF and in DHRF.

MATERIALS AND METHODS

The pilot plant set-up and general experimental procedures were described in Part I (Chapter 5).

Commercially available humic acid (Aldrich, Germany) was used in the experiments to determine the influence of humic substance in raw water. The C, H, N and O content of humic acid provided by the same supplier were analyses to be 59, 5, 1 and 35% respectively (Zhou et al., 1994). These authors also found that the adsorption of this humic acid on mineral particles was higher and desorption more difficult than of natural humic and fulvic acid obtained from River Dodder (Dublin, Ireland). Thus the influence of Aldrich humic acid on mineral stability and coagulation forces could be more pronounced than that of many natural organic matters. The stock solution was prepared by dissolving 1.5 g/l humic acid in demineralized water and continuously stirring the solution for 24 h. In order to achieve

a raw water concentration of 10 mg/l humic acid (colour 80 mg Pt-Co/l) before the filter, the required amount of stock solution was pumped by a peristaltic pump to the raw water preparation unit (where also kaolin is added) of the pilot plant set-up to provide approx. 1 h contact time with the particulate material in order to achieve an equilibrium adsorption-desorption condition.

Colour was determined by the platinum-cobalt method and the true colour and expressed as mg Pt-Co/l (APHA et al., 1992). UV absorbance at 254 nm wave length is generally accepted as a more simple and equally effective measurement of humic substance concentration (Tambo, 1990; Edzwald et al., 1985; Eaton, 1995). In the present artificial raw water the added humic acid was the primary constituent to the UV absorbance and colour. UV was measured by a spectrophotometer (UV-VIS Spectrophotometer No. 554, Perkin-Elmer, Ct., USA) and the samples to be analyzed were first filtered through a 0.45 μm cellulose acetate filter membrane to remove interfering turbidity caused by particulate material. Since UV absorbance of humic acid is pH dependent, all measurements were taken after correcting pH at 7.8. In a calibration curve to correlate colour and UV absorbance values a linear relationship was observed within the investigated range; correlation was very high ($R^2 = 0.99$) (Fig. 1). The raw water without humic acid had colour of 3 mg Pt-Co/l.

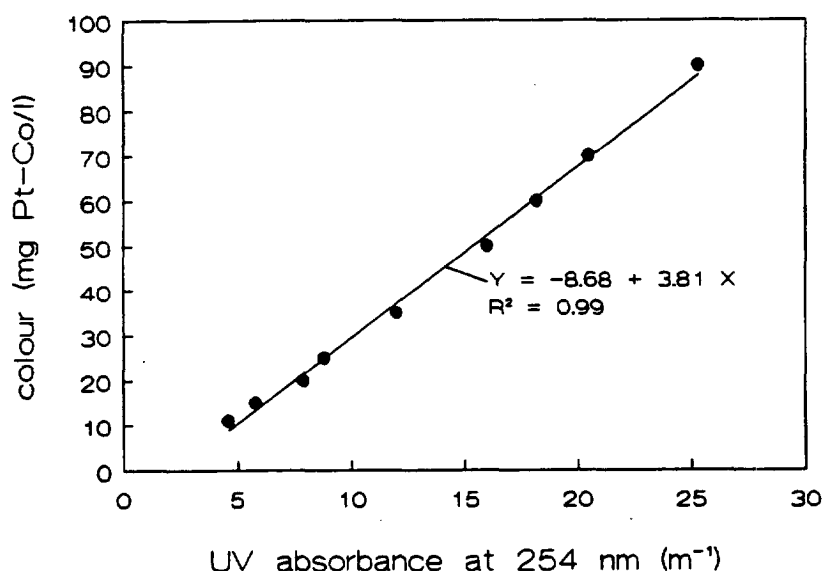


Fig. 1 Calibration curve for colour determination.

Jar tests were carried out in 2 litre pyrex beakers. The required equilibrium pH (after alum addition) was adjusted by adding predetermined amount of 0.1 M NaOH or 0.18 M HCl to the suspension. After adding alum the solution was rapidly mixed at $G = 180 \text{ s}^{-1}$ for 1

min. It was then flocculated for 20 min at $G = 15 \text{ s}^{-1}$ and allowed to settle for 30 min. Dissolved Al concentration was measured by a graphite furnace atomic absorption spectrophotometer (Model 1100 B, Perkin Elmer, Ct, USA). Samples were filtered through $0.45 \mu\text{m}$ cellulose acetate membrane to remove suspended solids which would otherwise interfere with the analysis.

In all filtration experiments an 8 m long filter was used which was divided into two equal compartments; the first compartment was filled with 20 mm and the second with 8 mm grains. As recommended optimum for high turbid water (see Part I - Chapter 5) a coagulant dose of 1 mg Al(III)/l was applied, and $G = 200 \text{ s}^{-1}$ and $t = 1 \text{ min}$ in the rapid mixing unit were used, in the experimental series to investigate the influence of varying turbidity (100-400 NTU). pH was 7.8 and filtration rate 5 m/h. The filter runs were operated till breakthrough. In the experimental series investigating the influence of humic acids, pH was adjusted with the necessary amount of 0.1M NaOH or 0.18M HCl before alum mixing. Filtration time in this case was restricted to 48 h as this period was sufficient to study the influence of humic acid on the ripening period and removal efficiency.

RESULTS AND DISCUSSION

Turbidity

The effluent turbidity of the filter run with an influent turbidity varying over time is compared with that of two filter runs with constant influent turbidity, the first with 200 NTU and the second with 400 NTU (Fig. 2). All other parameters remained identical. Effluent turbidity fluctuation in the filter run with the variable influent turbidity remained limited ($< \pm 1.5 \text{ NTU}$) despite the drastic influent turbidity change from 100 to 400 NTU. This effluent turbidity remained sufficiently small not to affect the function of a downstream slow or rapid sand filter. The average effluent turbidity was 2.4 NTU as compared to 2.1 and 3.5 NTU for the run with constant influent turbidity of 200 and 400 NTU, respectively. Even when the raw water turbidity increased to the very high 400 NTU the effluent was satisfactory. The abrupt changes in raw water turbidity were immediately absorbed and produced a rapid transition to the new steady-state situation within 1 h. When the turbidity level in the variable turbidity run was 200 or 400 NTU the corresponding effluent turbidity was comparable to that in the runs with constant 200 and 400 NTU with a variation of $\pm 10\%$. The headloss at breakthrough for the variable influent run was 14.4 cm, and for constant turbidity of 200 and 400 NTU run 17.2 and 15.5 cm, respectively. Comparable headloss ($\pm 10\%$) at the same filtration rate suggests similar deposit volume at the breakthrough.

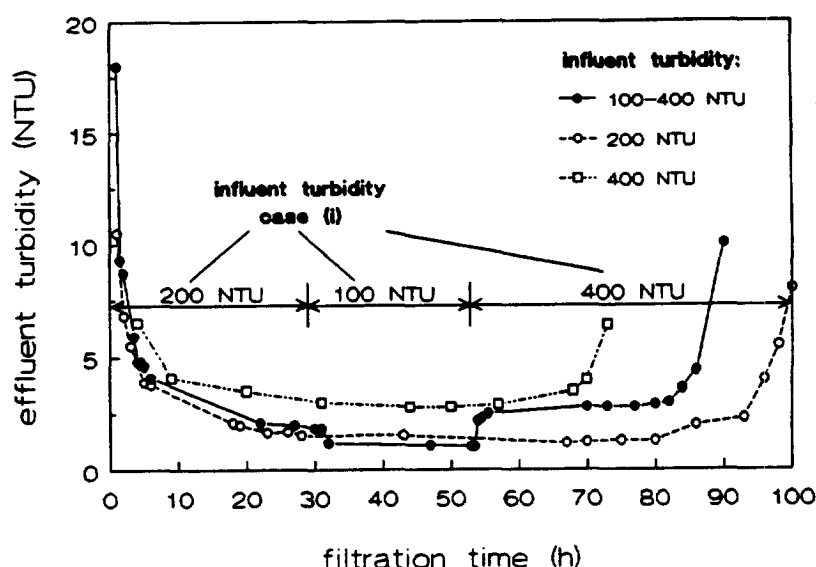


Fig. 2 Effluent turbidity of DHRF runs as a function of influent turbidity i.e. (i) variable 100-400 NTU, (ii) constant 200 NTU, and (iii) constant 400 NTU. Alum dose = 1 mg Al(III)/l, pH = 7.8, filtration rate = 5 m/h, total filter length = 8 m (first compartment L = 4 m and d_p = 20 mm, second compartment L = 4 m and d_p = 8 mm).

In direct filtration the effluent turbidity is often reported to not differ much with raw water turbidity fluctuation. Odira et al. (1987) studying upflow multilayer direct filtration (grain size 4.0-0.8 mm) with moderately turbid water (40-80 NTU) found that the optimum coagulant dose was about 1 mg Al(III)/l. They also observed that an abrupt increase of effluent turbidity up to 180 NTU for a short interval (2-3 h) had little influence on the filtrate quality. Cleasby (1984) also found that in common direct filtration with 0.7 mg Al(III)/l dose effluent turbidity remained nearly the same irrespective of fluctuations in raw water turbidity between 2 and 20 NTU. In addition, Guibai et al. (1988) showed that for the water with medium to high SS level (25-180 mg/l) the flocculation unit in a conventional process can be omitted, and in that case the optimum G value in the rapid mixing is also nearly independent of the SS concentration. Earlier investigation also showed that DHRF can produce effluent turbidity < 3-5 NTU when raw water pH varies in the range 6.5-9.0 (Part I - Chapter 5).

Filter runs were also carried out to assess the influence of filtration rate (3-7 m/h) at high influent turbidity (400 NTU) and under identical other process conditions (Table 1). The average effluent turbidity increased from 3 to 6 NTU for an increase in filtration rate from 3 to 7 m/h which, however, is still acceptable for a pretreatment process. With this high turbidity the filter run time was shortened substantially at higher filtration rate. For example,

with 200 NTU influent turbidity the filter run time for filtration rate 7 m/h was 75 hours (see Part I - Chapter 5) whereas it was 36 hours in the case of influent turbidity 400 NTU. The faster rate of floc deposition resulted in earlier breakthrough. Therefore, when very high influent turbidity (> 200 NTU) is expected for long times, filtration rates > 5 m/h should be avoided in DHRF. The headloss at the end of the filter run time also increased by 1-2 cm with the filtration rate but the magnitude of this increase does not make significant difference for applications.

Table 1 Performance of DHRF runs with 400 NTU influent turbidity. Coagulant dose = 1 mg Al(III)/l, pH = 7.8, filter length = 8m: first compartment L = 4 m and $d_0 = 20$ mm, second compartment L = 4 m and $d_0 = 8$ mm.

filtration rate	average turbidity after 1st compartment	average filter effluent turbidity	filter run time	total headloss at end of run time
(m/h)	(NTU)	(NTU)	(h)	(cm)
3	84	3	172	12.2
5	111	3.5	70	15.5
7	127	6	36	16.4

Turbidity reduction parallels reduction of bacteria and viruses (Hudson, 1962; Robeak et al., 1962). From the trend of turbidity and faecal coliform removal in different types of roughing filters (CINARA, 1993), and turbidity removal in DHRF it can be estimated that there would be 2-3 log removal of faecal coliform in DHRF.

Colour

The performance of HRF and DHRF in removing colour was investigated by comparative filter runs consisting of raw water with and without (additional) humic acid. In HRF mode filtration was carried out at filtration rates of 5 m/h (a proposed DHRF filtration rate) and 1 m/h (a typical HRF filtration rate). The residual colour of the HRF and DHRF runs are summarized in Tables 2 and 3, respectively, and compared in Fig. 3. In HRF, at filtration rate of 1 m/h the raw water colour of 80 mg Pt-Co/l declined gradually over the filter length to an effluent colour of, initially, 29 mg Pt-Co/l. The effluent colour, however, increased over time to about 57 mg Pt-Co/l after 48 h. At filtration rate of 5 m/h the colour removal was insignificant, it lowered only to 74 mg Pt-Co/l.

Table 2 Residual colour values along HRF filter length. Influent turbidity = 200 NTU and average colour = 80 mg Pt-Co/l, pH = 7.8, filter length = 8m (first compartment L = 4 m and d_0 = 20 mm, second compartment L = 4 m and d_0 = 8 mm).

filtration rate	filtration time	residual colour (mg Pt-Co/l) ¹	
		after 1st compartment	effluent
(m/h)	(h)		
1	2	56	29
	6	58	31
	24	66	41
	48	67	57
5	2	78	74
	6	78	74
	24	78	74
	48	78	74

¹ measured via UV adsorption

The presence of humic acid drastically increased the effluent turbidity (Fig. 3). In the HRF runs at filtration rate of 5 m/h and without humic acid the initial effluent turbidity was 105 NTU (influent turbidity 200 NTU); after a ripening period of 24 h this was reduced to 58 NTU. In similar runs but with humic acid present, the initial effluent turbidity was comparable (123 NTU) but, increased gradually to 135 NTU after 48 h. At 1 m/h filtration rate the removal efficiency of HRF is higher: initial effluent turbidity in presence of humic acid was similar (~25 NTU) to the filter run without, but it gradually increased to 35 NTU after 48 h. No filter ripening was observed in either of the HRF runs.

The colloidal stability of the kaolin particles appeared to be increased by the adsorption of humic acid molecules. During the initial filtration period the filter grains are clean and particle removal efficiency is influenced by the particle-grain attachment efficiency (Spielman and Fitzpatrick, 1973). The observed initial particle removal efficiency was similar for HRF runs with and without humic acid. Over time the filter grains are thought to be increasingly

covered with deposited particles. After the initial period particle removal is then influenced by the particles-deposit attachment efficiency (Vreeken et al., 1978; O'Melia and Ali, 1978; Graham, 1988). When humic acid is absent (e.g. Fig. 3, line c) the deposits act favourably as additional sites for deposition as is usually observed in deep bed filtration. The particle removal efficiency increases (filter ripening). In the presence of humic acid the particle-particle attachment efficiency α_p (similar to particle-deposit attachment efficiency) is reported to reduce (Jekel, 1986a). Thus the deposits with humic acid have lower particle-deposit attachment efficiency as compared to deposits without humic acid and possibly also to clean filter grains. As a result, in the HRF runs with humic acid no filter ripening was observed and the particle removal efficiency in fact gradually decreased over time due to progressing exhaustion of favourable deposition sites (clean grain surface).

Table 3 Residual colour along the DHRF. Initial raw water turbidity = 200 NTU, colour = 80 mg Pt-Co/l. Filtration rate = 5 m/h, filter length = 8m (first compartment L = 4 m and d_0 = 20 mm, second compartment L = 4 m and d_0 = 8 mm). results of the same experiments as Fig. 5.

coagulant dose and pH	filtration time (h)	residual colour (mg Pt-Co/l) ¹		
		after rapid mixing	after 1st compartment	effluent
1 mg Al(III)/l pH = 7.8	2	32	29	29
	6	32	33	31
	24	33	33	34
	48	34	34	30
1 mg Al(III)/l pH = 6.0	2	7	7	7
	6	7	7	7
	24	7	7	7
	48	7	7	7
2 mg Al(III)/l pH = 7.8	2	7	7	7
	6	7	7	7
	24	7	7	7
	48	7	7	7

¹ measured via UV adsorption

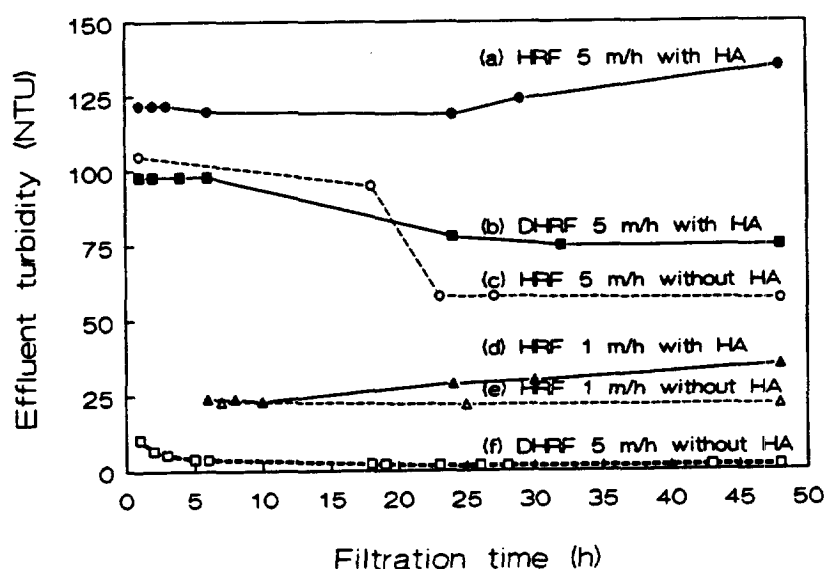


Fig. 3 Effluent turbidity of HRF and DHRF with and without humic acid added to the raw water. HA = humic acid, HRF = horizontal-flow roughing filtration, DHRF = direct horizontal-flow roughing filtration. Influent turbidity = 200 NTU and colour = 80 mg Pt-Co/l.

In the DHRF mode (coagulant dose = 1 mg Al(III)/l, pH = 7.8, filtration rate = 5 m/h) the influence of humic acid was drastic; the average effluent turbidity increased from 2 to 75 NTU when humic acid was present (Fig. 3). This was conceivably because of the competition between kaolin particles and humic acid molecules for Al molecules, in which the association of humic acid with Al is kinetically faster than that of inorganic particles (Narkis and Rebhun, 1975; Rebhun and Lurie, 1993).

Jar tests were carried out in order to analyze the influence of pH and alum dose on the turbidity and colour removal. Fig. 4a shows residual turbidity and Fig. 4b residual colour at different pH (4.5-7.8) and alum dose (1-12 mg Al(III)/l). The turbidity and colour removal at 1 mg Al(III)/l were lower than at higher doses. For all alum doses, the maximum colour reduction occurred in the pH range 5.5-6.5. This range is similar to what is generally recommended for optimal removal of humic substances. In the case of 1 mg Al(III)/l coagulant dose the turbidity and humic acid removal were becoming lower for increasing pH above 6.5. In this pH range the amounts of positively charged dissolved Al species and amorphous Al(OH)₃ decrease, but the amount of negatively charged humic acid functional group increases due to increased ionization (eqn. 1). When pH increased the lower coagulant dose became unable to satisfy the humic acid ionized groups as well as the kaolin particles. However, at higher coagulant doses the abundance of active Al species resulted in good humic acid and turbidity removal. The turbidity and colour increased also with decreasing

pH below 5.5. In this pH range the formation of $\text{Al}(\text{OH})_3$ is minimal as well as the amount of trivalent (Al^{3+}) and other positive divalent and monovalent Al species. The ionization degree of humic acid, however, decreases. The charge neutralization by positive Al monomers or polymers formed at pH 5.5–4.5 may be overall a less effect, and together with absence of sweep coagulation may cause the lower humic acid and kaolin particle removal.

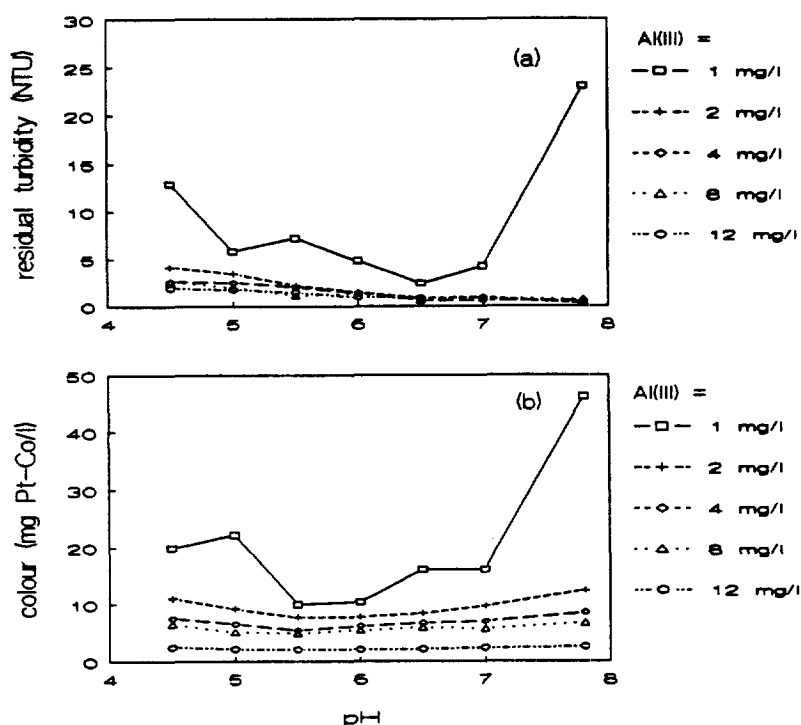


Fig. 4 Jar test results as a function of pH with (a) residual turbidity and (b) colour. Initial raw water turbidity was 200 NTU and colour 80 mg Pt-Co/l.

For optimal colour and turbidity removal, adjustment of pH and coagulant dose will depend on raw water quality and local circumstances. From the jar tests two combinations of alum dose and pH were selected reducing turbidity and colour to acceptable levels: (i) representing low alum dose and low pH (1 mg $\text{Al}(\text{III})/\text{l}$ at pH 6.0), and (ii) representing higher alum dose and high pH (2 mg $\text{Al}(\text{III})/\text{l}$ at pH 7.8). The variation of effluent turbidity of the two filter runs is shown in Fig. 5. The filter run with low coagulant dose (1 mg $\text{Al}(\text{III})/\text{l}$) and high pH (7.8) and in presence of humic acid (i.e. line b in Fig. 3) is also compared in Fig. 5. The colour reduction of the three filter run is shown in Table 3. Effluent turbidity did improve significantly after adjustment of pH (down to 6.0) or alum dose (up to 2 mg/l). The effluent turbidity at 2 mg $\text{Al}(\text{III})/\text{l}$ and pH 7.8 is lower (average 7 NTU) than at 1 mg $\text{Al}(\text{III})/\text{l}$ and pH 6.0 (average 13 NTU), possibly due to the formation of a higher amount of hydroxide precipitate at this higher alum dose.

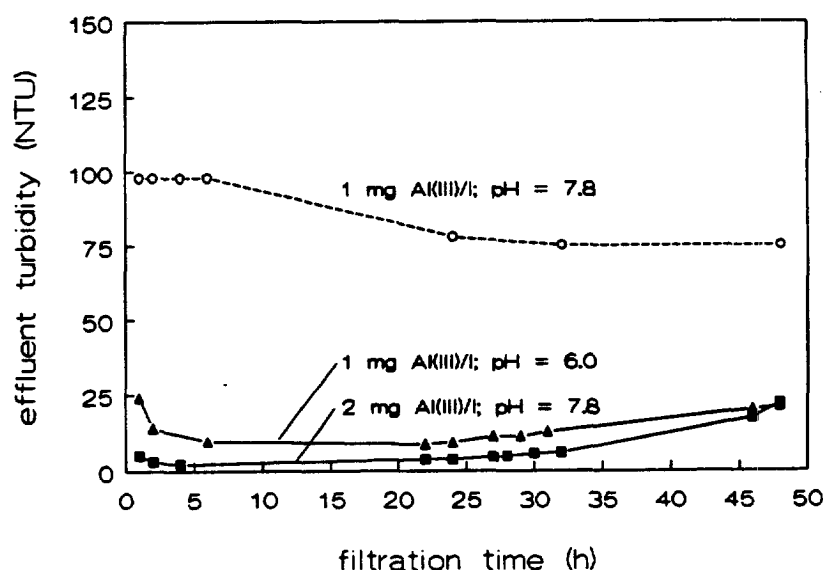


Fig. 5 DHRF effluent turbidity at different Al(III) dose and pH in presence of humic acid. Raw water turbidity = 200 NTU, colour = 80 mg Pt-Co/l.

The effluent colour for both the runs with (i) low pH and low alum dose, and (ii) high pH and high alum dose were reduced to 7 mg Pt-Co/l. This is below the WHO guideline for drinking water of 15 mg Pt-Co/l. A further decrease in colour (20-50%) can also be expected to take place in the subsequent slow or rapid sand filter. For the filter run with 1 mg Al(III)/l and pH 7.8 the influent colour of 80 was reduced to only about 30 mg Pt-Co/l. The reduction of (true) colour took place almost entirely in the rapid mixing unit meaning that the association of the originally dissolved humic acid with Al is almost complete within the first minutes. According to Semmens et al. (1980) and Edzward (1986) the association of organic matter and Al species is complete within 1 sec. The flocculation seems not to play any additional role in dissolved organic matter removal except that it promotes floc formation for subsequent easier removal from the suspension.

The filter run time for both optimized filter runs was about 40 h. This is markedly lower (30-50%) when compared with similar filter run without the presence of humic acid (filter run times were 90 and 60 h and final headloss 16.5 and 15.9 cm, for a dosage of 1 and 2 mg Al(III)/l at pH 8.1, respectively (see Table 1 in Part I - Chapter 5). When humic acid is present the headloss at breakthrough was only 8.5 and 9.4 cm, respectively. As headloss is indicative of the deposit volume accumulated within a filter, the lower headloss in the presence of humic acid suggests that breakthrough in filter runs occurred when the porosity of the filter was not reduced to the same extent as compared with filter runs with colour-free raw water. Consequently, the shear stress was not as high as in the latter case. A lower shear

strength of deposits was reported to occur when humic acid is present by Rebhun (1990), and could be correlated with early breakthrough in DHRF.

CONCLUSIONS

DHRF showed to be a versatile pretreatment process in handling wide fluctuation in raw water turbidity while the process conditions remained unchanged. When the effluent turbidity fluctuates 100-400 NTU, DHRF could still produce stable effluent turbidity (2-4 NTU) and ensure satisfactory operation. If high turbidity (> 200 NTU) is expected for a prolonged time then high filtration rates (> 5 m/h) should be avoided as this would result in short operating time (typically < 2 days).

The optimum coagulant dose (1 mg Al(III)/l), rapid mixing intensity ($G = 200-300 \text{ s}^{-1}$) and time (1 min) determined for raw water of 200 NTU was found also to be applicable for a wider range of turbidity (100-400 NTU) and pH (6.5-9.0). In case when the operational parameters such as coagulant dose and filtration rate require to remain unchanged for reasons of simple and reliable operation DHRF will be an effective pretreatment process to produce a stable effluent quality even if the raw water quality fluctuates widely or shock loads are received. Less operational control, such as adjusting coagulant dose, pH, and rapid mixing intensity and time, is necessary in DHRF over a wide range of the raw water quality turbidity and pH. The risk of sudden clogging and effluent quality deterioration of the subsequent slow or rapid sand filters can therefore be avoided or minimized. Simple DHRF operation suggests that it could be an appropriate pretreatment technology for towns and small cities in developing countries.

HRF was not effective in removing colour. At 1 m/h filtration rate (raw water turbidity = 200 NTU and colour = 80 mg Pt-Co/l) colour was reduced 60% initially, however, after 48 h operation the colour removal efficiency gradually dropped to 30%. The turbidity removal efficiency dropped simultaneously from 90% to about 80% after 48 h. At higher filtration rate of 5 m/h HRF performance was very poor, only 8% colour removal and 35% turbidity removal occurred. No filter ripening was observed in HRF.

Presence of 80 mg Pt-Co/l colour deteriorated the average effluent turbidity drastically from 2 NTU to about 75 NTU for a typical DHRF run (influent turbidity = 200 NTU, filtration rate = 5 m/h, coagulant dose = 1 mg Al(III)/l and pH = 7.8). The colour reduction was about 65%. Good performance, however could be achieved in DHRF by increasing coagulant dose or lowering pH, or both, depending upon the raw water quality and the local circumstances. At coagulant dose 2 mg Al(III)/l and pH 7.8, or at 1 mg Al(III)/l and pH 6.0, the colour removal was 90% and effluent turbidity about 10 NTU.

The presence of humic acid markedly reduced the DHRF run time to about half. The low shear strength of the deposits in the presence of organic matter is believed to have caused early breakthrough.

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Chapter 7

Development of Roughing Filter Cleaning Procedure¹

ABSTRACT - Development of appropriate roughing filter cleaning procedure is explored and the underlying mechanisms were investigated. Surface washing was found to be an appropriate filter cleaning technique. Use of coagulants produced deposits which are more easy to clean. Investigation also suggest that deposits become increasingly difficult to clean with increasing deposit age and thus roughing filter should be cleaned at least once a week.

INTRODUCTION

Roughing filters are used for pretreatment of highly turbid water (100-400 NTU) especially prior to slow sand filters in developing countries (Schulz and Okun, 1984; CINARA, 1993). They can be up-, down- or horizontal-flow depending on the flow direction and usually consist of 2 or 3 subsequent compartments with coarse to fine filter grains. The filter grain size varies typically 25 mm in the first compartment to a minimum of 2 mm in the last compartment. Filtration rate is low, generally less than 1 m/h. In conventional roughing filter no chemicals (coagulants) are added; removal efficiency is generally 80-90% (Wegelin et al., 1991, 1988; Collins et al., 1991). In order to improve the performance of horizontal-flow roughing filtration (HRF) the process has been modified earlier by applying a small constant coagulant dose, typically 1 mg Al(III)/l, prior to filtration (Chapter 2). The modified process is called direct horizontal-flow roughing filtration (DHRF). Such filters are typically 8 m long and divided into 2 compartments, the first with coarse grains (15-25 mm diameter) and the second with finer grains (4-15 mm). On raw water of 100-400 NTU DHRF can be applied at a higher filtration rate (3-7 m/h) and the removal efficiency is 98-99%.

Rapid sand filters are usually cleaned by backwashing (at a velocity of 40-60 m/h) to cause fluidization of the bed (10-20% expansion). The upflowing water causes dislodgement of deposits from the grain surface. The forces acting on deposits and grains include fluid shear, collision and abrasion; the fluid shear force is reported to be the dominant (Amirtharajah,

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1993; Fitzpatrick, 1993). Fluidization ensures that all grain surfaces are exposed to the shearing effect and that all loosened deposits are flushed out. However, backwashing is not feasible for roughing filters because it would require an excessively high water velocity to achieve bed expansion especially in the first compartment. Also, such filters are conceived to be of simple construction which would call for manual cleaning.

Hydraulic flushing

The filter run time in roughing filters varies from a few days in up- or down-flow filters to several weeks and even months in HRF. The run time of DHRF can be 3-15 days depending on influent turbidity, filtration rate and grain size. Cleaning of the first generation HRF was done by manually removing the filter grains, and washing and replacing them. Hydraulic cleaning has been proposed as an improvement over manual cleaning (Wegelin, 1986; Galvis and Visscher, 1987; Pardon, 1992; CINARA, 1993). The drainage system consists of perforated pipes or troughs at the filter bottom and connected to outlet valves. In hydraulic flushing the filter box is filled with water and drained by opening the underdrainage valves quickly. This cycle is repeated several times till clear washwater is obtained. In HRF all the filter compartments are built in one structural unit and hydraulically connected; only one or two drainage valves are opened at a time in order to generate enough scour velocity near the drainage conduits connected to those valves. The inlet section with coarse grains is recommended to be cleaned first as it retains most of the deposits and in the next cycles progressively other valves towards the end of the filter are opened. A minimum grain size of 4 mm is recommended in HRF to allow effective filter cleaning (Wegelin, 1986).

The hydraulic flushing is schematized in Fig. 1. Deposits are flushed down to the filter bottom (step 1); effectiveness of this step is determined by the cohesive strength of the deposits and their attachment to the grains, and the shear stress created by the downward flowing water. Deposit is then transported from the filter bottom to the drainage system (step 2); this is influenced by the density and viscosity of the deposit and the filter bottom construction (slope, spacing of drains and openings). Deposit thereafter is washed out through the underdrainage system (step 3) which depends on the water velocity through the drainage openings pipe. The operation of steps 2 and 3 can be ensured by suitable structural and hydraulic design of the filter bottom. In step 1 the downward drainage velocity v_d (interstitial velocity) in the filter bed depends on (i) mainly the hydraulic resistance of the underdrainage system, and (ii) to a certain extent on the grain size. The cohesive strength of the deposit, however, is not affected by design considerations. Deposit cohesive strength is influenced mainly by physio-chemical characteristics of the suspended solids, presence of natural organic matter and the deposit age (Chapter 5).

In practice cleaning efficiency by hydraulic flushing has been reported for HRF to be

insufficient (Wegelin, 1991; Pardon, 1992). The downward drainage velocity v_d is reported to vary from 5-15 m/h in Colombian HRF to 50-90 m/h in Peruvian installations (IRCWD, 1989). In Peru troughs with high proportion of open areas were used in underdrains to achieve the high drainage velocity. However, physical investigation in Peru revealed that with this velocity only 30-50% of the deposit could be removed from the inlet section with 20 mm grains. With other grain sizes (down to 5 mm) the removal was reported to be 80-90% (IRCWD, 1988). Nevertheless, a moderate rate of 10-30 m/h and a perforated pipe as underdrainage were recommended for HRF (IRCWD, 1989). The unremoved deposit in the filter pores accumulates over successive cleaning operations and is likely to gradually increase headloss and decrease filter run time to a limit where economic filter operation cannot be continued. After prolonged operation, periodic hydraulic flushing of the roughing filters may not be sufficient to reestablish filter efficiency. Manual cleaning will then be required (Wolters et al., 1989). This is a costly and labour intensive procedure. During such manual cleaning of up-flow roughing filters in India the accumulated filter deposits are found to be black and petrified (Haryana Public Health Branch, 1995).

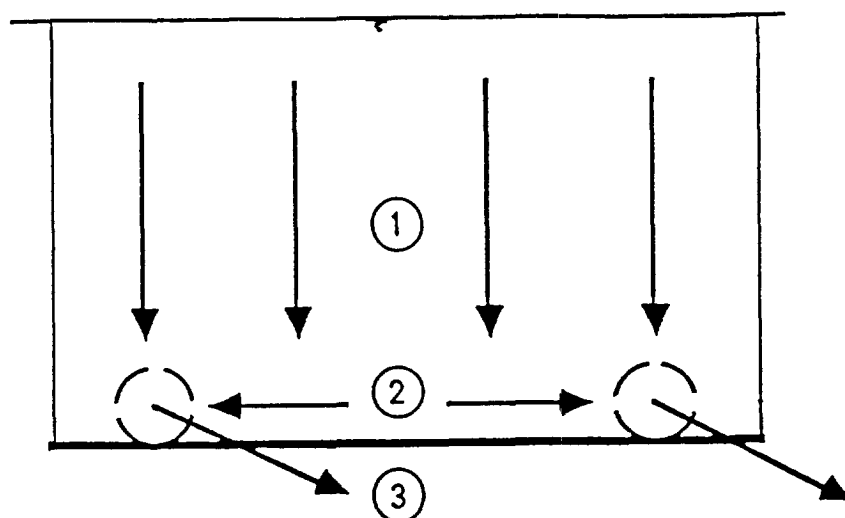


Fig. 1 Schematic representation of sludge transport during hydraulic cleaning (IRCWD, 1988). step 1: flushing of deposits downwards, step 2: horizontal transport towards the drainage conducts, and step 3: transportation through drainage conducts.

Several measures have been suggested to improve the hydraulic cleaning. Alternating fast opening and closing of drainage valves was advocated to destabilize the accumulated deposit on the grains by creating shock waves (water hammer) (Wolters et al., 1988). However, this method had little effect in practice and could endanger the hydraulic installations (IRCWD, 1989). In order to achieve higher drainage velocity the drainage pipes' outlet can be placed at lower elevation (to obtain higher head) or the number of underdrainage pipes and outlets

points can be increased (IRCWD, 1988). Surface washing of HRF with installed showers has also been tried in Peru and in Switzerland with but limited success (IWCWD, 1989).

Experience with an optimized roughing filter cleaning procedure is limited. Cumbersome and ineffective filter cleaning is regarded as one of their major drawback. This Chapter aims at investigating an appropriate roughing filter cleaning procedure. Special attention is given for removal of specific floc deposits in DHRF.

Three cleaning procedures were compared in a pilot filter box: (i) hydraulic flushing procedure by filling and draining, (ii) surface washing by first draining till empty the filter bed and followed by washing over the bed by a hose pipe, and (iii) air scour by first draining the filter and then again filling full and providing air scour, followed by another draining. All the washing procedures were repeated till the washwater from the underdrains were clear (turbidity < 2-3 NTU).

MATERIAL AND METHODS

The shear strength of deposits and flow velocity required to destabilize and resuspend them under different conditions were assessed in jar tests. The stirrer blade (75 mm x 25 mm) was placed 2 cm above the 2 litre jar bottom. A layer of 4 mm sand was glued to the jar bottom to resemble a filter grain surface. Raw water was made of kaolin suspension (200 NTU). It was rapidly mixed with predetermined amount of Al(III) at 150 rpm for 1 min and then flocculated for 20 min at 30 rpm. The solution was allowed to settle and kept undisturbed for different periods (1-33 days) to reflect aging of deposits. After each test period the water was stirred at different speeds (5-150 rpm) and turbidity measurements were taken after 1 min stirring time. Experiments were also done without coagulant to determine the scour velocity required for roughing filters without coagulants. Duplicate jar test results were taken; standard deviation was always < 2%.

It is assumed that the velocity required for deposit detachment (i.e. effective cleaning) in the filter is comparable to that required in a jar test for total resuspension of deposits. During mixing the water velocity in a container and at the tip of the blade is lower than the angular velocity of the blade tip due to water slip; the velocity during jar tests is taken as 0.75 times the tip blade velocity (Metcalf and Eddy, 1991). The vertical free fall v_d with different grain sizes was assessed experimentally. 1 m high and 0.2 m diameter perspex columns were filled with grain sizes of 0.8 mm, 2 mm, 4 mm, 6 mm, 13 mm and 20 mm, respectively. 2 litre water was poured over the gravel bed at a rate equal v_d (determined experimentally for each grain size). The time required for 50% of the poured water to reach the column bottom was taken for calculating the mean free falling v_d .

Filter cleaning experiments were carried out in a rectangular DHRF of 5 mm thick perspex material in a metal frame. It was 2 m long, 0.2 m wide and 0.4 m high and was filled with a different gravel size for different runs. The underdrainage system consisted of perforated collecting pipes (10 cm apart and 3 nos. 5 mm openings per cm) perpendicular to the flow direction and spaced 20 cm apart. Each underdrainage pipe was connected to a valve. The discharge end could be placed at a lower elevation (through flexible pipes) to achieve higher v_d within the gravel bed; v_d was adjusted to 20-30 m/h. Perforated PVC pipes were also placed along the filter bottom to provide air scour. Also, a perforated (perforations of 5 mm diameter, 25,000 nos./m²) false filter bottom could be placed at a height of 5 cm from the bottom. Baffle walls were provided (20 cm apart) below the false bottom in the horizontal direction to prevent short-circuiting. Kaolin suspension of 200 NTU was coagulated with 1 mg Al(III)/l and filtered through the gravel bed at 5 m/h. Filtration was continued for 5-7 days, specific deposit before cleaning was 15-20 g/l.

RESULTS AND DISCUSSION

Deposit strength and aging effect

The effect of the deposit age on the scour velocity required for resuspension of deposits is shown in Fig. 2a. A coagulant dose of 1 mg Al(III)/l was used. The velocity required for resuspension increased with the deposit age. The scour velocity required for 100% resuspension was the same for 1 to 7 days sludge age. For sludge age up to 7 days scour velocity of 145 m/h was required; for sludge age of 12 days 190 m/h was needed. With further increase in deposit age the deposits further consolidated and the required velocity increased exponentially. When the sludge was about one month old only approximately 10% of the sludge could be resuspended with a scour velocity of 240 m/h.

The effect of deposit age in the case where coagulants are not used (e.g. HRF) is shown in Fig. 2b. A similar pattern of increasing scour velocity requirement with deposit age was observed. The scour velocity requirement was, however, higher than in the case with coagulant. After deposit age of 7 days 50% resuspension was possible at scour velocity of 240 m/h as compared to 110 m/h when 1 mg Al(III)/l was applied. Clearly, the use of coagulants produces deposit which can be detached more easily. The comparative jar tests suggest that the required scour velocity for effective cleaning is in the order of 2 to 3 times less when coagulants are used. Fitzpatrick (1993) visually observed deposit detachment during backwashing by endoscope. Coagulated kaolin deposits broke away in larger chunks than the uncoagulated kaolin; deposit characteristics related to detachment changes due to coagulation. In present practice HRF and other roughing filters are often cleaned after long intervals of weeks or months. The results suggest that such prolonged deposit retention may

be the main reason for reported difficulty in complete filter cleaning. Moreover, long sludge retention will encourage biological growth within the filter which tends to further consolidate deposits and make them even more difficult to detach from the grains.

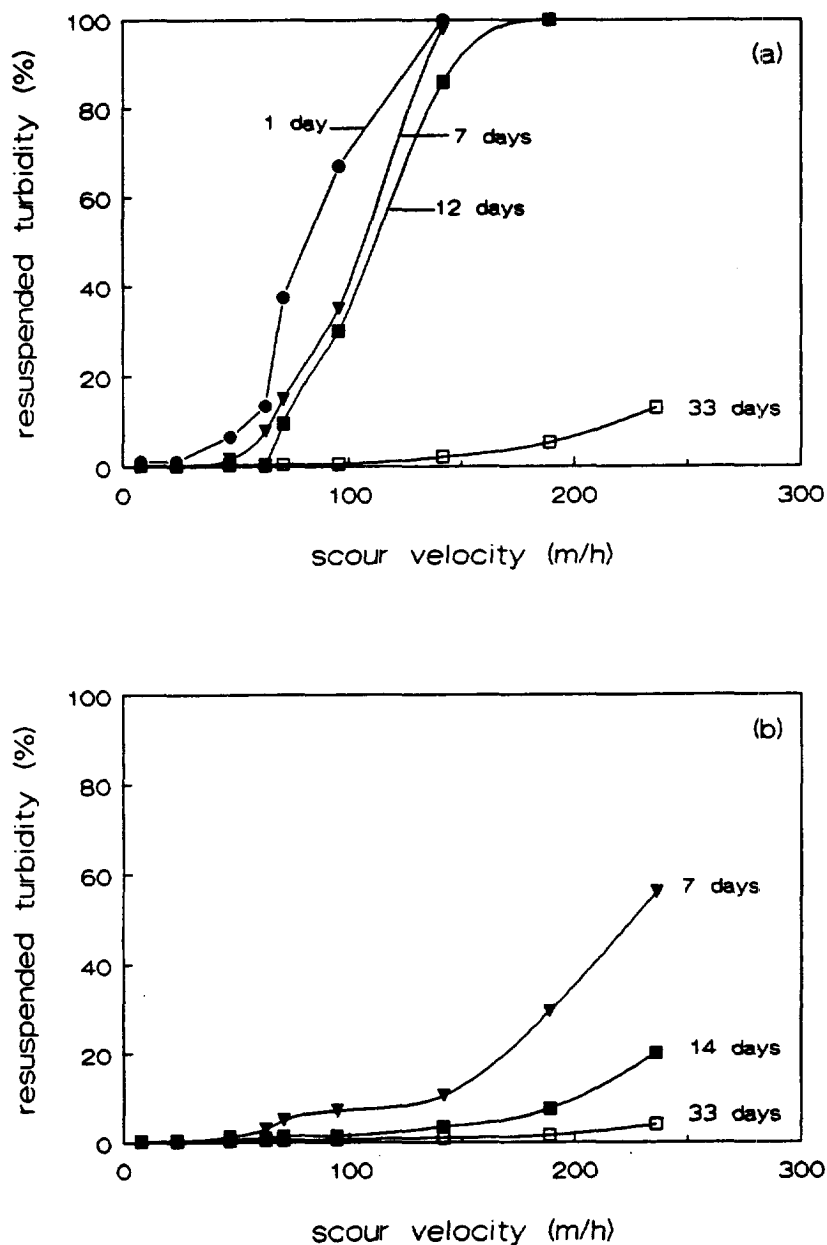


Fig. 2 Scour velocity required in jar tests for resuspension of deposit of different age. Initial turbidity = 200 NTU. (a) With coagulant dose of 1 mg Al(III)/l; (b) without coagulant.

The effect of the coagulant dose (0-4 mg Al(III)/l) on required scour velocity is shown in

Fig. 3. A dose of 1 mg Al(III)/l produced sludge deposit that required minimum scour velocity. Coagulants produced flocs of hydrolysed products that have lower density and internal resistance to shear stress than only mineral particles. Insufficient alum dose (e.g. 0.5 mg Al(III)/l) caused less domination of flocs by the Al hydrolysis products, and flocs are likely to behave partly like pure mineral agglomerates. With a further increase in coagulant dose (2-4 mg Al(III)/l) higher scour velocity was required, which was likely related to the increasing volume of deposit produced at higher coagulant dose (Chapter 5).

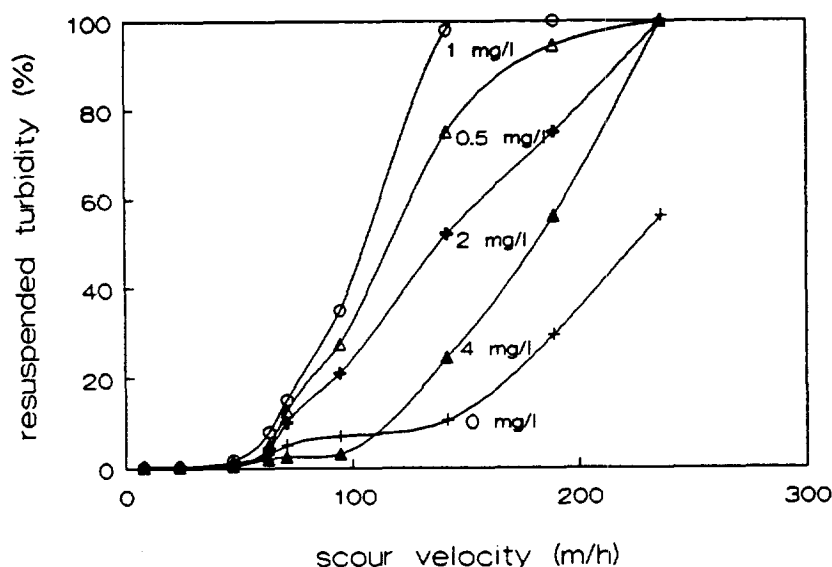


Fig. 3 Required scour velocity for resuspension of deposits in a jar test for different coagulant dose. Initial turbidity = 200 NTU, deposit age = 7 days.

The roughing filters cleaning procedure under immobile grain condition (as in the case with hydraulic flushing) has an inherent weakness as not all the grain surface is equally exposed to the water shear. The corners between the contact points of grains would be less effectively exposed. Such corners of large grains are more remote from the main stream flow. The relationship of turbidity resuspension with scour velocity in jar tests (Fig. 2) was carried out with a bottom 4 mm grain layer. The actual scour velocity required for larger grains (say 20 mm) may be higher than predicted by the jar tests.

Surface washing

An alternative way to generate a high drainage velocity at low cost is to take the advantage of gravity and washing from the top of the filter (surface washing). Washwater is delivered by a hose pipe that slowly moves along the gravel bed surface. After the bottom drainage

valves have been opened and surface washed one to two times most of the deposits are easily removed from the gravel bed. Further removing the remaining deposits are critical. The hydraulic resistance of the filter bed with few remaining deposit can be approximated as that of a clean bed; an expression for free falling velocity under clean bed condition is developed below.

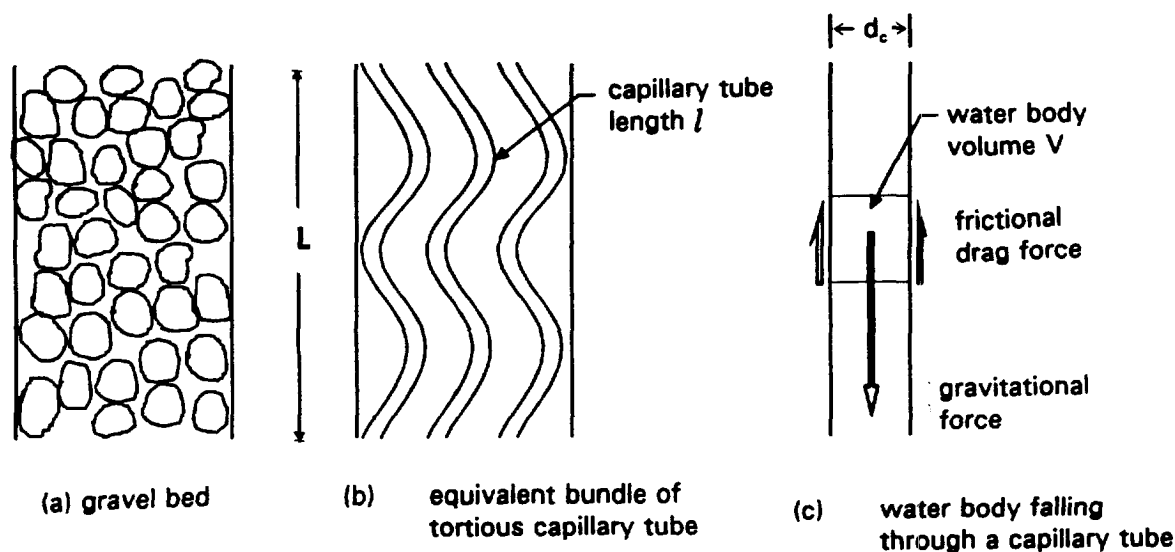


Fig. 4 Free falling velocity through gravel bed.

In gravel bed the washwater will be accelerated downward under gravity but flow will experience resistance because of shear against the grains and energy loss by its tortuous paths through the bed. The frictional loss of water flow through a gravel bed can be determined using Carman-Kozeny's approach (Kozeny, 1927; Carman, 1937). The gravel bed (Fig. 4a) is schematized as a bundle of capillary tubes (Fig. 4b). The length of the capillary tube l is larger than the gravel bed length L to incorporate the tortuosity of the flow path through the gravels. The capillary tube diameter is d_c and a same flow rate as the interstitial velocity in the filter is considered. The pore volume and total surface area of the gravel bed are equated with those of the capillary tubes bundle; the tube diameter d_c then can be expressed as:

$$d_c = \frac{2}{3} \left[\frac{\varepsilon}{1 - \varepsilon} \right] d_s, \quad (1)$$

where ε is porosity and d_s drain diameter. Let us consider a water volume V falling through such a capillary tube with increasing velocity under gravity (Fig. 4c). This will be resisted by the increasing frictional force between the water and tube surface. After a certain distance a steady state velocity will be reached. The gravitational force is given by

$$\text{gravitational force} = \rho g V, \quad (2)$$

where ρ is water density and g acceleration due to gravity. The frictional drag force depends on the velocity of the body v_c , fluid density, and drag coefficient C_D and the contact surface area A_c between the water body and the capillary tube. Additional resistance due to the tortuous water path through the gravel bed is to be considered; the tortuosity factor (l/L in Fig. 4b) is estimated to be $\sqrt{2}$ by Carman-Kozeny. Multiplying C_D with the tortuosity factor frictional drag force can be expressed as

$$\text{frictional drag force} = \frac{(\sqrt{2} C_D) A_c \rho v_c^2}{2}. \quad (3)$$

By equating the gravitational force and the frictional drag force the steady state falling velocity v_s can be obtained:

$$v_s = \left[\frac{2g}{\sqrt{2} C_D} \frac{V}{A_c} \right]^{1/2}. \quad (4)$$

The ratio V/A_c equals the ratio between the tube cross-section and its perimeter. This relationship also holds for flow condition when tube cross-section is partially filled. For full and partial flow conditions, the hydraulic diameter d_h is defined as four times the ratio between the cross-sectional area and the wetted perimeter (Morries and Wiggert, 1972) i.e.

$$\frac{V}{A_c} = \frac{\text{cross-sectional area}}{\text{wetted perimeter}} = \frac{d_h}{4}. \quad (5)$$

Using eqn. 5 in 4

$$v_s = \left[\frac{g}{2\sqrt{2} C_D} \cdot d_h \right]^{1/2} \quad (6)$$

Eqn. 6 is now a general expression for the steady state velocity of falling water in a bed of

any grain size. It shows that the falling velocity is independent of the length of pores filled with water, however, dependent on the degree of pore cross-section filled. Eqn. 6 is also applicable for full or partially filled flow conditions. The hydraulic diameter d_h corresponding to a grain size and for different flow conditions can be determined by using eqns. 1 and 5. The drag coefficient C_D is a function of Re and the shape of the object. The drag force for an object moving through a fluid is mathematically equivalent to the fluid flowing passed a stationary object: C_D for the water body flowing through the capillary tube is equivalent to a cylinder moving through water. The relationship of C_D with Re for cylinders has been experimentally determined (e.g. Metcalf and Eddy, 1991).

The calculated steady state falling velocity (eqn. 6) as a function of grain size and flow condition is given in Fig. 5. The percentage of filled flow refers to the percentage of tube cross-sectional areas filled with water. The value of C_D is approximately constant for $Re > 100$ and up to 10^5 (turbulent flow). At Re less than 100 the viscous forces govern (laminar flow) and C_D increases as a function of decreasing Re . At grain diameter less than 4 mm generally $Re < 100$ and the falling velocity sharply reduces due to smaller hydraulic diameter and also higher C_D . The falling velocity decreases with decreasing grain diameter due to decreasing capillary hydraulic diameter (pore resistance). For a particular grain (or tube) diameter the hydraulic diameter decreases for flow condition below 50% water filled, however it increases for above 50% filled. Thus, for a particular grain size the falling velocity is proportional to the flow condition: falling velocity for 80% full flow is higher and for 20% full flow lower compared to 100% full flow (Fig. 5).

The calculated falling velocity in Fig. 5 gives a good qualitative relationship during surface washing. The actual falling velocity may deviate from the calculations. The delivery of water over the a particular gravel bed section is discontinuous and may not achieve steady state initially. A flux of water is rushed downward under gravity and fills the pore cross-section partly full and partly not-full. The degree of filling depends on the washwater flow and the time of holding the hose over the bed section, and possibly also on grain size. Part of the downflow water can also spread laterally which is like to flow under partially filled (possibly $< 20\%$) condition. When the first flux of water is flowing over dry grains (i.e. not submerged) this creates impact and will cause additional resistance to flow. The magnitude of all these factors can only be estimated. In order to incorporate those losses and deviations in the model a C_D value 2 to 3 times higher than that used corresponds with the observed falling velocity in the 4-20 mm grain size range.

The falling velocity decreases more rapidly with grain size below 4 mm. For rapid sand filters (grain size 0.8-2.0 mm) surface washing would generate a velocity of only 25-50 m/h (equivalent to backwash rate of 10-20 m/h). Therefore, surface washing is not a feasible method for rapid sand filters.

A scour velocity of 145 m/h sufficed for 100% resuspension of 7 days old flocculated deposit (Fig. 2a). This velocity can be correlated with the observed falling velocity (Fig. 5). Thus, for surface washing to be appropriate for DHRF cleaning, the minimum grain size should be 3-5 mm. Wegelin (1986) also observed that HRF becomes increasingly difficult to hydraulically clean below 4 mm grains. When coagulants are not used it can be extrapolated from Fig. 2b that about 2 times higher scour velocity was required and the minimum grain size would then be 15-20 mm. Therefore, smaller grain sizes can only be afforded when coagulants are used.

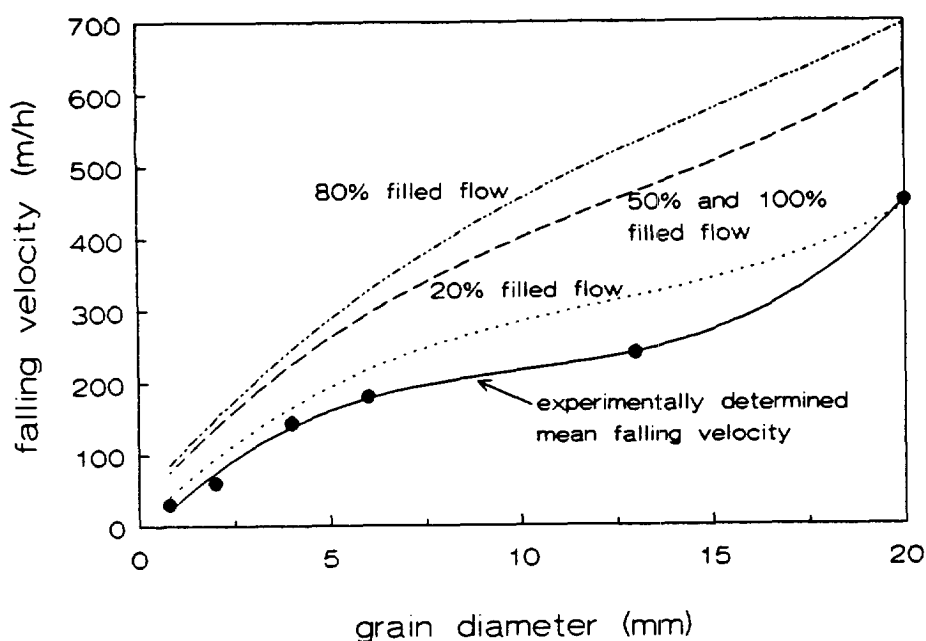


Fig. 5 Mean vertical free flow velocity of water in the pores of different grain size filters, as a function of grain size and percentage of the pore cross-sectional area filled with water.

Drainage velocity during hydraulic flushing

During hydraulic flushing the water level (head) in the filter box gradually reduces which decreases the drainage velocity v_d . The amount of water in a gravel bed is the bed volume multiplied by its porosity ϵ . A gravel bed can be schematized as the a container with same height as that of the bed height but horizontal cross-sectional area of the container (S_c) is p times the filter box cross-sectional area (S_f). Thus, the water volume and available head are equal in both the cases i.e. they are hydraulically equivalent in static condition. The drainage time for a container with constant horizontal cross-section and an opening at its base is given as (Degremont, 1991):

$$t = \frac{2S_c (\sqrt{h_1} - \sqrt{h_2})}{k_1 s \sqrt{2g}}, \quad (7)$$

where s area of opening, h_1 and h_2 are the initial and final height of water above the opening, respectively, k_1 the contraction coefficient and g is the acceleration due to gravity. The average drainage velocity between a section can be determined by dividing the section length $(h_1 - h_2)$ with t required to travel that distance.

During hydraulic flushing majority of the flow resistance occur at the openings of the underdrainage and is proportional to the total opening area. Other relative minor losses are frictional losses in the grains and in the drainage conducts and valves; they can be expressed in terms of equivalent reduction in total open area. If the total area of openings in the underdrainage is b percent of the gravel bed surface area then the equivalent open area s in the container will be $b.S_f.k_2$, where k_2 is the correction factor for minor losses (< 1). Using the value of s and $S_c = \varepsilon.S_f$, eqn. 7 becomes

$$t = \frac{2\varepsilon (\sqrt{h_1} - \sqrt{h_2})}{k_1 k_2 b \sqrt{2g}}. \quad (8)$$

Using eqn. 8 the drainage velocity v_d and the total time which the grain are exposed to water is shown in Fig. 6. Here $b = 0.1\%$, $k_1 = 0.62$ (for circular openings), $k_2 = 0.8$ (assumed) and $\varepsilon = 0.4$ are considered. The bed height is 1.5 m and 10 cm sections are considered in the calculations. v_d reduces as the water level moves to lower bed heights; the grains above the water level is no more exposed to water shear. At a section 25 cm above the filter bottom the total exposed time is 300 s. But during this time v_d reduces from 24 to 10 m/h. Another section 25 cm below the top surface is only exposed for 45 s during which v_d reduces from 24 to 22 m/h.

Comparative tests of cleaning procedures

The relative effectiveness of the three filter cleaning procedures for with different grain size was assessed by measuring the required washwater volume (Table 1). The air scour with 20 mm grains showed that air passed through preferential paths leaving other parts untouched from its flow. Similar observation was also made by Amirtharajah (1984) in case of rapid sand filters and air scour alone (i.e. without simultaneous backwash water flow). Visual

observation showed that the cleaning by air scour was not very effective and further tests with air scour were discontinued.

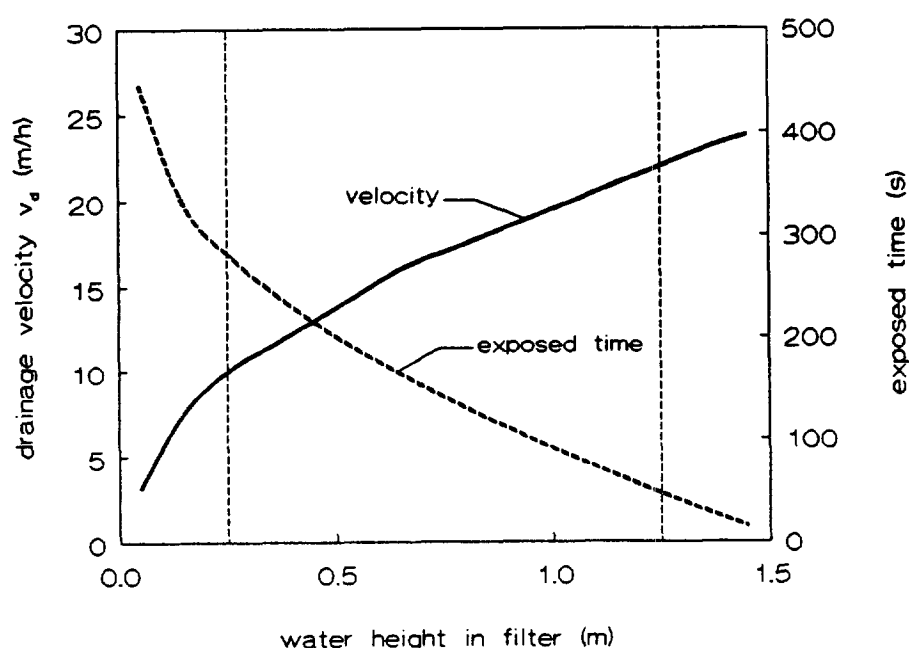


Fig. 7 Drainage velocity v_d and exposed time of water to grains as a function of reducing water level in a filter during hydraulic washing (calculated according to eqn. 8: $b = 0.1\%$, $k_1 = 0.62$, $k_2 = 0.8$ and $\epsilon = 0.4$).

The grains were excavated after the filter cleanings. After hydraulic flushing small amounts of deposit ($\sim 1\%$) remained in the corners between grains. No significant remaining deposits were observed after surface washing. As compared to hydraulic flushing surface washing required 3 to 4 times less wash water and all grain sizes could be effectively cleaned.

Surface washing could generate about 5 times higher v_d than the usual hydraulic flushing rate (10-30 m/h). This high velocity ensured effective cleaning. Re in the pores (see Chapter 4, eqn. 5) for the hydraulic flushing varied from 100 for 20 mm grains to 27 for 4 mm grains (at 20 °C). Whereas, in case of surface washing Re based on observed free falling velocity (Fig. 5) was estimated to be 1,500 for 20 mm and 100 for 4 mm grains. Turbulent flow is defined as the fluctuation of micro-velocity. Wright (1968) studied the flow pattern through gravel bed and found that at $5 < Re < 90$ the linear relationship between headloss and filtration rate does not hold, however, the micro-velocity does not fluctuate. At $Re > 90-120$ the flow is in turbulent transition (micro-velocity fluctuation in regular frequency) and at $Re > 800$ it is fully turbulent (micro-velocity fluctuation is vigorous and random). In case of surface washing highly fluctuating micro-velocity may be expected to occur and vigorous

eddies possibly developed behind each gravel layer assisting in deposit detachment in the next gravel layer.

Table 1 Required wash water (m^3 water/ m^2 filter bed surface) to achieve effective deposit removal in different filter cleaning procedures. First with filter water is not included.

grain size (mm)	filter cleaning procedure		
	hydraulic flushing	surface washing	air scour
4	1.5	0.5	-
8	1.0	0.3	-
13	1.3	0.4	-
20	1.5 [‡]	0.4	0.9 [‡]

[‡] could not be fully cleaned

In hydraulic flushing the washwater is not efficiently utilized: all the vertical sections are not exposed to total drainage time and furthermore, v_d gradually reduces and becomes increasingly less effective. Whereas, in surface washing the washwater remains in a constant high velocity and the exposed time is same for all sections; washwater is efficiency used. Moreover, during flushing the flow can be short-circuited and flow towards drainage openings leaving some "dead zones" in between the openings.

During surface washing part of the downflow water is in full flow condition. Part above and below the full flow section is likely to be partially filled. The water in the partially filled section ($> 50\%$ filled) and above the 100% filled section will have relatively higher falling velocity than the latter section (Fig. 5). As a result the water from the upper section will move in to the fully filled section and create turbulence; this will assist in deposit detachment. The impact of rushing wash water on dry grains and filter bottom adds to its better cleaning. Moreover, the amount of washwater flowing through a underdrainage system is low and thus possible water logging (prevention of quick discharge) at the filter bottom is avoided. The downflowing wash water can impact directly on un-submerged filter bottom covered with deposits in order to detach them easily.

To expose all the grain corners to water scour part of the washwater should flow under 100% filled condition so that water scour and turbulence reaches the remote corners. Therefore

during surface washing the hose pipe should be held over a bed section long enough so that a 100% filled water column can reach the filter bottom but, not too long to cause water logging; holding the hose 2-3 s over a 10 cm^2 surface section could be adequate. Operator experience is further required for optimizing this technique.

In order to ascertain the usefulness of a false filter bottom in better filter cleaning and deposit removal (especially at the filter bottom) filter runs were carried out with a false bottom and with different grain sizes. In all cases much short-circuiting of the horizontal flow in the compartments below the false bottom was observed despite baffle walls in the horizontal flow direction were provided. As a result the effluent quality deteriorated. Short-circuiting was more pronounced in smaller grains ($< 8 \text{ mm}$) because of their higher flow resistance. Similar short-circuiting was also observed in HRF with false filter bottom (Mbwette, 1992). Therefore, horizontal-flow filters with false bottom is not advisable. A bottom slope towards the drainage conducts should be provided for easy deposit movement.

CONCLUSIONS

This investigation addresses the underlying mechanisms important for roughing filters cleaning. The study was limited to synthetic suspension of mineral particles. The presence of natural organic matters and biological activities may exhibit different deposit adhesive characteristics. However, the findings are believed to be qualitatively applicable to natural waters. The surface washing technique should be field tested to determine its effectiveness for different conditions and for further development. The following conclusions can be made:

Surface washing was found to be an appropriate procedure for cleaning roughing filters. The mean free falling velocity during surface washing can be approximated by modelling the bed as a bundle of capillary tube according to Karman-Cozeny (eqn. 6). However, the drag coefficient C_D determined for ideal conditions used should increased 2-3 times to incorporate deviations.

Use of coagulants produces deposit which are which are easier to clean and required about 2-3 times less wash water velocity than when coagulants are not used. Thus in DHRF smaller grain size can be afforded; the minimum grain size is 3-5 mm.

Long deposit retention time, both with and without coagulants, makes deposit increasingly difficult to clean: roughing filters should be cleaned at least once a week. The present practice of several weeks operating time in HRF should be avoided.

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NOTATIONS

- b = percent of gravel bed surface area, -
- C_D = drag coefficient, -
- d_c = diameter of capillary tube, m

- d_g = diameter of filter grain, m
 d_h = diameter of filter grain, m
DHRF = direct horizontal-flow roughing filtration
 g = gravitational constant (9.81), m/s²
HRF = horizontal-flow roughing filtration
 h_1, h_2 = initial and final height of water above a drainage opening, m
 k_1 = contraction coefficient, -
 k_2 = correction factor for minor losses, -
 l = length of capillary tube, m
 S_c = equivalent container cross-sectional area, m²
 S_f = gravel bed surface area, m²
 s = area of open in a container, m²
 t = time, s
 V = water volume, m³
 v_d = interstitial drainage velocity, m/s
 v_s = steady state falling velocity, m/s
 ε = gravel bed porosity, -
 ρ = water density, kg/m³

DESIGN AND OPERATION

ABSTRACT - A design guideline for direct horizontal-flow roughing filtration (DHRF) is proposed which is applicable for predominantly mineral particles and for the suspension in optimized conditions. For other water types column settling tests and jar tests are useful tools for predicting DHRF performance. A comparative analysis between horizontal-flow roughing filtration, DHRF and conventional flocculation-sedimentation process suggests that DHRF is an appropriate pretreatment technology for towns and small cities in many developing countries.

INTRODUCTION

The effluent concentration of a filter shows a characteristic variation during its run featuring initially a steep decrease (filter ripening) and, after a stable main period (working period), a final increase (breakthrough). The headloss across the bed increases steadily with filtration time. If t_c and t_h are the times in which the allowable limits in effluent concentration and headloss, respectively, are reached, Mintz (1966) showed that the filtration process is optimum when

$$t_c = t_h. \quad (1)$$

Commonly, rapid sand filter operation terminates when headloss reaches a certain maximum value at t_h . In order to avoid turbidity breakthrough and ensure acceptable effluent quality it is generally desirable that $t_c > t_h$.

The objective of process optimization would be to maximize the volume of filtrate and to minimize the volume of wash water required, subject to some boundary conditions such as required effluent quality, allowable headloss, etc. (Adin and Hatukai, 1991). Process optimization can identify several effective alternative combinations. However, only one design combination can produce the desired effluent at the least cost. This is the economic optimum design.

Researchers have attempted to optimize filter in various fashions. Huang and Baumann (1971) used an empirical filtration model for iron floc removal and showed that various combinations of grain sizes and filter depths can produce a desired effluent quality. Letterman (1980) considered the total cost to be a function of filter bed area and concluded that higher filtration rates are generally economical even if filter runs are shorter i.e. (direct) rapid sand filtration would be economically attractive even for comparatively high turbid water.

Ives (1980) systematically discussed the effect of the four main process parameters on the process optimization for a uniformly graded single medium filter i.e. (i) filtration rate, (ii) filter run time, (iii) filter depth, and (iv) grain size. Sambi and Ives (1983) considered the four process parameters and determined maximum production of water between backwashes. Wiesner (1985) proposed an economic optimization model considering only filtration rate and run time, and simple cost functions. Adin and Hatukai (1991) developed an economic optimization model for multilayered filters considering the process variables including coagulation, and using pilot-plant data. They also determined that filtration rate is the most significant cost determining factor. In a similar but more elaborate approach Dharmappa et al. (1992) proposed an economic optimization model and algorithm, but using laboratory data and a computer simulation. They demonstrated the strong effect of particle size distribution of the raw water on total cost. Other important considerations that are reported to influence the optimal design parameters and hence overall cost are the desired filter effluent quality, the type of pretreatment (coagulation, flocculation, etc.) and the filter cleaning procedure (Adin et al. 1979; Adin and Hatukai, 1991).

The optimum design is achieved by two successive steps: process optimization followed by economic optimization. In the economic optimization models, costs are determined by cost functions for different construction and operating units (e.g. Clark, 1982; Clark and Dorsey, 1982). These costs are, however, valid for a specific locality and time. The different cost components such as construction cost, labour cost, operation and maintenance cost, interest rate and availability of materials and supplies vary from country to country and over time. Optimum design for one country may not necessarily be economically optimum for others.

DHRF DESIGN GUIDELINES

In DHRF coagulants are added to the raw water in a rapid mixing unit. The coagulated water is delivered to the roughing filter unit. Filtration is achieved in constant rate mode. The filtrate water is then passed to the subsequent slow or rapid sand filters.

Coagulation conditions

The coagulation and rapid mixing processes were optimized in Chapter 5. It was found that a coagulant dose of 1 mg Al(III)/l is applicable for a wide turbidity range (100-400 NTU) and pH range (6.5-9.0) (Chapter 6).

Filter configuration

Investigation on DHRF has shown that two filter compartments are sufficient to produce systematically good effluent water quality (turbidity < 2-3 NTU) and achieve its objective as pretreatment. A third filter compartment with finer grains (< 3 mm) to produce still lower turbidity may not be required in a pretreatment mode. Moreover, the smaller grains compartment will be disproportionately more difficult to clean (Chapter 7), will increase headloss and adds to the construction and maintenance costs. The remaining turbidity from the two-compartment DHRF can effectively be removed in subsequent e.g. rapid sand filter and where the deposits can be removed by backwashing.

The structural requirements of the DHRF filter box and its dimensions can be in principle similar to those proposed for HRF as recommended in the design and operation manual of Wegelin (1986). A bed height of 1-1.5 m is recommended for HRF for the convenience in periodic manual cleaning. Such manual cleaning would not be required for DHRF with appropriate cleaning procedure (Chapter 7), at least not as frequent as in HRF, and thus the bed height could be increased. A height of about 2 m can be proposed if structural cost and manual filter cleaning considerations allow so. The filter width can be 2-5 m, similar to what recommended for HRF for convenient filter cleaning and maintenance. Two parallel filter units at least should be provided to allow treatment continuity during maintenance of one unit.

With increasing headloss with filtration time the free water level in the filter bed rises. This has to be compensated by additional bed height over the effluent weir. To keep the construction cost down the total headloss in DHRF should preferably be limited to 20 cm. Moreover, an extra height of 10 cm of the filter bed is recommended. This will minimize sunlight intrusion and prevent algal growth; similarly, it will protect against contamination of the water in the filter bed from the surroundings.

The headloss in DHRF was investigated in Chapter 5 with 4 m long first compartment (grain size 20 mm) and 4 m long second compartment (grain size varied 4-13 mm). Headloss primarily occurred in the second compartment, the headloss in the first compartment being < 5 mm. The total headloss in DHRF with grain sizes from 8-13 in the second compartment and filtration rate 3-7 m/h varies from 12 to 18 cm and is within the headloss limit. A smaller grain size i.e. 4 mm, would produce a total headloss of 25 to 43 cm for filtration rate from 3 to 7 m/h. Thus, grain size < 8 mm is not desirable in terms of headloss limitation.

Filtration rate > 7 m/h would also increase headloss to more than 20 cm. Moreover, in such cases t_c may be less than the necessary cleaning interval t_{clean} (Chapter 7). Thus, a filtration rate up to 7 m/h is feasible for DHRF. The recommended process conditions and filter configuration are summarized in Table 1A.

Table 1 Tentative design and operation guideline for DHRF.

A. Process conditions and filter configuration

Raw water turbidity	100-400 NTU
Process condition	
pH	6.5-9.0
coagulant type	Alum, Ferric chloride, etc.
coagulant dose	e.g. 1 mg Al(III)/l, 1 mg Fe(III)/l
rapid mixing time (t)	1 min
rapid mixing intensity (G)	100-300 s ⁻¹
filtration rate	3-7 m/h
Filter configuration	
no. of compartments	2
filter height	1.5-2 m
filter width	2-5 m
minimum grain size	3-5 mm for effective filter cleaning
allowance in bed height (over the effluent overflow weir)	30 cm (20 cm for headloss, plus 10 cm to protect against sunlight and other contamination)
filter underdrainage	Perforated pipes or troughs connected to washout valves. Filter bottom slope towards the drainage pipes.

B. Expected residual turbidity (as % of influent turbidity) and headloss in DHRF.

grain size range	filtration rate (m/h)	filter length (m)					terminal headloss (cm/m filter)
		1	2	3	4	5	
<u>1st compt:</u> 20-25 mm	3	80.0	49.4	29.7	25.2	21.0	negligible
	5	83.9	64.6	40.4	28.9	25.5	negligible
	7	85.5	72.7	50.6	34.3	27.7	negligible
<u>2nd compt:</u> ¹ 10-15 mm	3	40.0	17.2	8.0	4.6	4.0	3.6
	5	41.7	16.7	8.3	4.2	4.0	3.2
	7	42.9	17.7	8.6	4.3	4.2	3.0
6-10 mm	3	30.0	9.0	5.0	3.2	3.0	3.8
	5	31.7	9.8	5.8	3.3	3.0	4.3
	7	31.4	10.0	6.5	3.6	3.5	4.5
3-5 mm	3	7.2	0.8	0.6	0.04	0.04	6.3
	5	11.7	1.3	1.2	1.2	1.1	8.6
	7	14.3	4.3	3.1	2.9	2.8	10.9

¹ after 4-5 m first compartment

C. Filter operations

Cleaning procedure	<ol style="list-style-type: none">1. Open all the underdrainage valves starting from the inlet.2. Drainage valves remain open. Apply wash water e.g @ 0.5 m³/h by a 5 cm diameter hose pipe over the filter surface. Move the hose pipe over all the surface @ 2-3 s for each 10 x 10 cm section. More than one hose pipe may be used in distant sections for faster procedure.3. Stop washing when the wash water from the underdrains is clean. Close all underdrainage valves.
Cleaning interval	Maximum 7 days.
Filter commencement	In the first hour double coagulant dose can be applied to shorten the ripening time (from 5 h to < 1 h).
Filter termination	At predetermined day(s) in a week based on turbidity breakthrough.

Filter length

The expected time-average (during the filter working period) residual turbidity (%) at different filter lengths for a two compartment filter and for filtration rate of 3-7 m/h is shown in Table 1B. The Table is primarily based on results obtained from lab-scale pilot plant with a raw water turbidity of 200 NTU (without organic matter) and coagulated with 'optimum' coagulant dose of 1 mg Al(III)/l and pH 7.8 (Chapter 5). The grain size ranges in the Table are representative for the average grain diameter with which experiments were carried out. The values for 5 m length are based as extrapolation. Experimental results showed that the Table can generally be used for raw water turbidity of 100-400 NTU and pH 6.5-9.0 within $\pm 10\%$ accuracy (Chapter 6). The terminal headloss is also shown for different grain size and filtration rate. The headloss in the first coarse compartment is negligible (~ 1 mm/m length) and can be omitted in calculating the total filter headloss.

The residual turbidity after the first compartment can be found by multiplying the efficiency number in the Table by the raw water concentration. The turbidity after the second compartment can be determined by multiplying the effluent concentration of the first compartment by the efficiency ratio.

The maximum allowable headloss in DHRF is low (20 cm) compared to rapid sand filters (1-2 m) but increases at a low rate (3-5 cm/day) over 3 to 7 days; filter operation termination on the basis of allowable headloss as in rapid sand filters would not be feasible for DHRF. Moreover, with a relatively larger grain size in the second compartment ($> 3-5$ mm) the total headloss may remain below 20 cm even when there is a turbidity breakthrough (Chapter 5). In an optimum DHRF design the headloss not necessarily has to reach its maximum allowable limit, as is common in the case of rapid sand filters (eqn. 1). However, it is an important design criterion that the total headloss should be within its limits.

Long filter operating time beyond 7 days is required to be restricted as the deposit becomes more difficult to be removed during cleaning (Chapter 7); this is a consideration for the optimization. Thus, in DHRF the filter run time is mainly limited by two boundary conditions: (i) the time to reach the allowable limit of effluent concentration (t_c), and (ii) the maximum filter cleaning interval (t_{clean}). The optimization expression for DHRF thus becomes, with t_{opt} the optimal filter run time

$$t_{opt} = t_c = t_{clean}, \quad (2)$$

and under the boundary condition that total headloss \leq its maximum limit. In order to have a safety in effluent turbidity $t_c > t_{clean}$ is preferred. This means that DHRF filter operation

is best terminated on the basis of predetermined cleaning interval. Rapid and slow sand filters commonly are terminated on another criterion i.e. predetermined headloss limit.

The filter breakthrough time t_c can be estimated on the basis of the ultimate specific deposit of the two compartments. The first compartment removes the major portion (about 75%) of the incoming turbidity load but also has a higher ultimate specific deposit capacity. A design example is given below.

DHRF operation

Filter control can be comparable to what is recommended in the HRF design manual (Wegelin, 1986). A distribution box can divide the flow to different filter units. Water level in the filter are controlled by a downstream effluent weir.

When commencing a filter run the rate of coagulant dose can be increased, e.g. doubled, for the first one hour to shorten the filter ripening time, e.g. from 5 hours to less than 1 hour (Chapter 6). The initial water production will then not need to be wasted, and the transfer of higher turbidity water to the subsequent sand filter can be minimized during that period. The filter cleaning interval t_{clean} was determined to be maximum one week (Chapter 7). To be on the safe side, or when t_c is expected to be less than a week, more frequent cleaning (up to twice a week) may be desired. In those cases fixing one or two particular days in a week (say every monday and thursday) will be convenient for a semi-skilled operator in a developing country. Filter cleaning procedure is discussed in Chapter 7. DHRF operation is recommended in Table 1C.

Design example

Consider raw water turbidity of 200 NTU and a desired DHRF effluent turbidity of 3 NTU. As a first estimate, the length of the first compartment is considered 4 m and filtration rate 5 m/h. From Table 1B the effluent of the first compartment will be $(200 \times 0.289 =) 59$ NTU. If 5-10 mm grain size range is selected then at a filtration rate of 5 m/h a length of 3 m will produce $(59 \times 0.58 =) 3.5$ NTU and 4 m will produce $(59 \times 0.33 =) 2$ NTU effluent turbidity. By interpolation 3.5 m second compartment length would produce about 3 NTU. For an conservative estimate it can be considered that the ultimate specific deposit capacity of the first compartment is 20 g/l and second compartment 7 g/l (Chapter 5). When a filter cleaning frequency of twice a week is considered then there will maximum 4 days cleaning interval, i.e. $t_{clean} = 96$ h. The first compartment would reach its ultimate specific deposit capacity in 105 h (t_c of 1st compartment) and the second compartment (t_c of 2nd compartment) in 85 h; but t_c (2nd compartment) $< t_{clean}$. Increasing t_c by e.g. increasing the second compartment length can be considered. For 4 m second compartment length t_c will

be 100 h, total headloss is below 20 cm. The process can be considered to be reasonably optimized (eqn. 2) for a DHRF with a 4 m first compartment (grain size range 15-25 mm) and a 4 m second compartment (grain size range 5-10 mm) and which would produce effluent turbidity of 2 NTU.

Alternatively, if in the second compartment a larger grain size range e.g. 10-15 mm is used then 4 m length will produce effluent turbidity of 3 NTU. The difference in effluent turbidity may not be significant for pretreatment purpose, however, bigger grain sizes may be desirable for more convenient filter cleaning. Thus, process optimization can be done with different combinations of filtration rate, filter length and grain sizes to obtain the desired effluent quality under the boundary conditions for optimization. Considering local costs and other conditions, a least cost solution can be determined from the alternative process-optimum configurations.

PERFORMANCE INDICATORS

Table 1 predicts the DHRF performance for mineral particles and under optimum coagulation conditions for these circumstances. This would be representative of many river systems (Chapter 2). However, when the raw water is highly coloured then higher coagulant dose and adjustment of pH will be required (Chapter 6). In these cases DHRF performance can be estimated by some simple indicator tests with actual water and coagulation conditions.

For water composition or operational condition deviating from those in the pilot experiments, column settling test data can be used in the model developed for the first compartment of DHRF to determine adequate compartment length (Chapter 3). Thus, for various raw water and process conditions (coagulant dose, mixing time and intensity, filtration rate) the residual turbidity profile can be predicted to an accuracy of $\pm 10\%$. Fig. 1 shows the residual turbidity profile of the first compartment of DHRF for filtration rate 1-7 m/h. For example, if 25% residual turbidity after the first compartment is desired (shown as dotted line in Fig. 1), the required compartment length corresponding to different filtration rates can be determined. Higher filtration rate will require longer compartment length for the same degree of removal but on the other hand will need smaller vertical cross-sectional area. The performance of the DHRF first compartment strongly influences that of the subsequent compartment. Therefore, optimum design of the first compartment is an essential criterion for optimum design of the total filter.

Jar tests are generally used to predict the performance of water treatment plants under various water quality and process conditions (Hudson, 1979; Hudson and Wagner, 1981). A general correspondence between optimum coagulation identified in jar tests with optimum

floc characteristics both for colour and turbidity removal has been reported (Bache et al., 1995). The performance of DHRF can be well predicted qualitatively by jar tests for various water quality (e.g. when organic substances is present) and coagulation conditions (coagulant dose, pH). Fig. 2a compares the residual turbidity of jar tests and DHRF effluent for various coagulant doses (0.5–4 mg Al(III)/l). The DHRF filter consisted of a 4 m first compartment (20 mm grains) and a 4 m second compartment (8 mm grains), and filtration rate 5 m/h. In this set of jar tests the process condition were similar as that in Chapter 4: the solutions were rapidly mixed at $G = 200 \text{ s}^{-1}$ for 1 min and flocculated at $G = 15 \text{ s}^{-1}$ for 20 min. However the settling time was 30 min in order to have comparable residence time as in the filter (38 min).

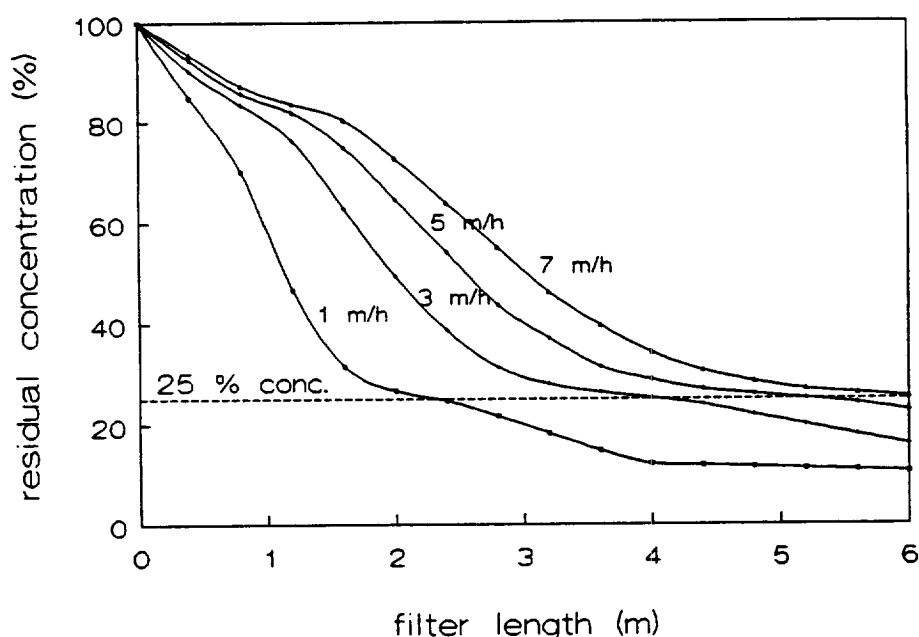


Fig. 1 Residual concentration profiles for different filtration rates as predicted by the model for the first coarse compartment of DHRF. Grain size = 20 mm, initial turbidity = 200 NTU, coagulant dose = 1 mg Al(III)/l.

A similar trend of decreasing turbidity with coagulant dose can be observed in both cases. The jar test results, however, gives only qualitative dose-response relationship; the quality of a DHRF effluent is also influenced by other filtration process conditions like length of filter compartments and their grain sizes, filtration rate, etc. For raw water colour of 80 mg Pt-Co/l, Fig 2b shows the colour of the supernatant in the jar tests at different pH (4.5–7.8) and coagulant dose of 1 and 2 mg Al(III)/l. With the same raw water the effluent colour of three DHRF runs with different combinations of coagulant dose and pH were tested: (i) coagulant 1 mg Al(III)/l and pH 7.8, (ii) coagulant dose 2 mg Al(III)/l and pH 6, and (iii)

coagulant dose 2 mg Al(III)/l and pH 7.8. The DHRF effluent colour is compared in the same graph. Again the colour reduction pattern in jar tests and DHRF showed a similar trend.

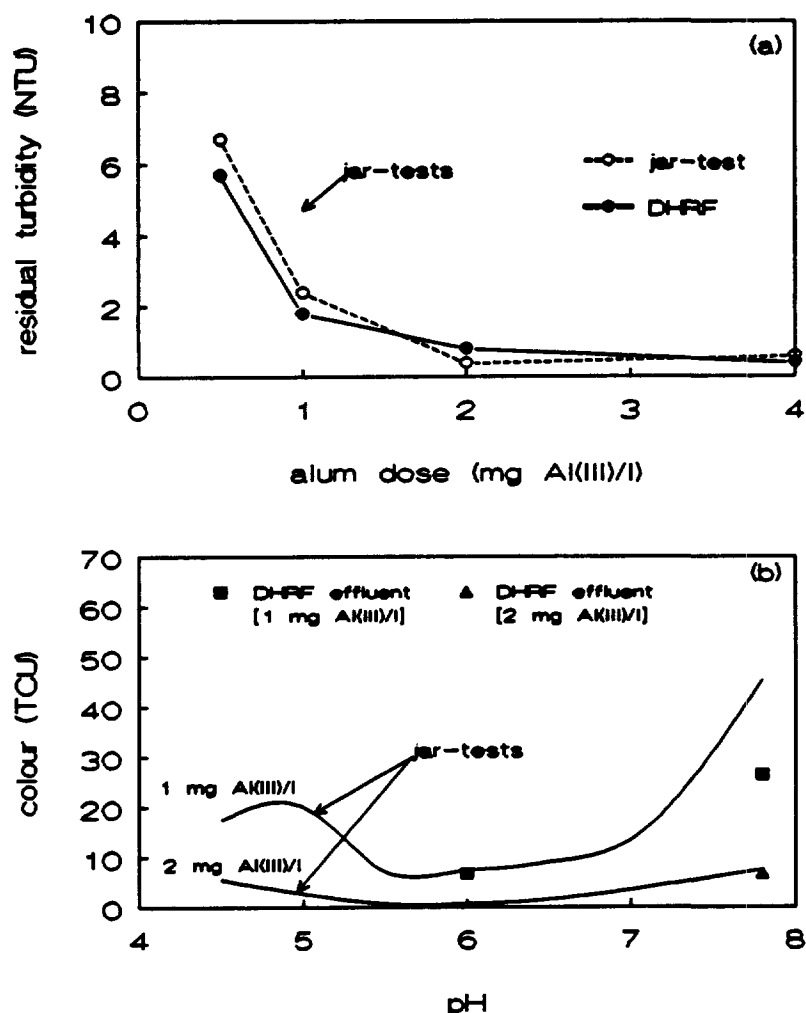


Fig. 2 Comparing the DHRF effluent with jar test results. (a) Residual turbidity; raw water turbidity = 200 NTU. (b) Colour; raw water colour = 80 mg Pt-Co/l and turbidity = 200 NTU.

Jar tests can also be used to determine the deposit characteristics related to filter cleaning (Chapter 7). If organic matter is present in water the deposit strength is decreased causing early breakthrough (Chapter 6). Jar tests may be useful in estimating deposit strength. These results suggest that the jar test can be a useful tool in optimizing the coagulant dose and pH for turbidity and colour removal, and ascertaining deposit characteristics in DHRF relevant for cleaning of filter. The column settling test and jar test results with actual raw water can be compared with those of mineral particles (e.g. Fig. 1 and 2a). The expected deviation

caused by using the design Table can be estimated from the deviations between the column and jar tests with actual suspension and model suspension.

For more accurate prediction and optimal design, a pilot-plant should be constructed. The indispensability of pilot-plants in filter design has since long been noticed (e.g. Camp, 1964; Mintz, 1966; Ives, 1975; Adin et al., 1979). The design tables or the jar and column tests are not intended to replace pilot-plants but would provide a good approximation, reducing the number of experiments required and allowing to use the data in an optimal way.

COMPARING PRETREATMENT ALTERNATIVES

Process Mechanisms

The main characteristics of three pretreatment methods prior to final filtration- HRF, DHRF and the conventional process of flocculation followed by sedimentation - are compared in Table 2. The dominant particle removal process in HRF is claimed to be sedimentation (Wegelin et al., 1987; Boller, 1993), however, in smaller grains other filter transport mechanisms like diffusion can be significant for submicron particles. The first compartment retains the majority of the suspended solids load at the expense of low headloss. The subsequent compartment(s) with smaller grains has(ve) higher particle removal efficiency. Small and colloidal particles are likely to pass the HRF (Boller, 1993). In DHRF coagulant is added in the rapid mixing unit where particles are destabilized. Coagulation is usually achieved in the combined (sweep and adsorption-destabilization) zone (Chapter 3 and 6). The first filter compartment acts as a combined flocculator and settler; the second compartment is characterized predominantly by deep bed filtration (Chapter 2).

Floc breakup in the conventional process can be significant during flocculation and depends on the applied G value and detention time during flocculation; maximum floc size is a function of G (Argaman, 1970; Argaman and Kaufman, 1970). In the DHRF gravel bed flocculator and sedimentation take places simultaneously, the floc breakup is believed to be minimal as the larger flocs (which are more susceptible to breakage) are readily removed from suspension and settles onto the grains. The smaller flocs (which are less susceptible to breakage) and remaining in the suspension will still continue to grow and subsequently removed. Whereas in conventional flocculator no selective (larger) floc withdrawal occur; the Gt values are required to be optimized to minimize the floc breakup.

Table 2 Comparing three pretreatment processes: horizontal-flow roughing filtration (HRF), direct horizontal-flow roughing filtration (DHRF) and conventional flocculation-sedimentation.

characteristics	HRF	DHRF	conventional flocculation-sedimentation
unit operations	multi-stage roughing filtration	rapid mixing -> roughing filtration	rapid mixing -> flocculation -> sedimentation
process mechanisms	<p>particle removal predominantly sedimentation</p> <p>progressively higher removal efficiency in the subsequent compartments with smaller grains</p> <p>in smaller grains other filtration mechanism like diffusion can also be effective</p>	<p>rapid mixing: particle destabilization and initialization of floc formation. Coagulation generally in zone of combination of sweep and adsorption-destabilization.</p> <p>roughing filtration: 1st compt - simultaneously flocculation and sedimentation.</p> <p>2nd compt - predominantly deep bed filtration.</p>	<p>rapid mixing: particle destabilization and initiation of floc formation. Coagulation generally in sweep coagulation zone.</p> <p>flocculation: floc growth and simultaneous floc breakup. Maximum floc size proportional to G.</p> <p>sedimentation: particle removal by flocculant settling.</p>
expected effluent turbidity (influent 100-400 NTU)	5-20 NTU	2-5 NTU	5-10 NTU

characteristics	HRF	DHRF	conventional flocculation-sedimentation
layout (5000 m ³ /day capacity)	44 x filter each (9x4x1.5)	1 x rapid mixing unit (V = 4 m ³) 7 x filter each (8x4x2)	1 x rapid mixing unit (V = 2 m ³) 2 x flocculator (mechanical) each (7x3x2) 4 x sedimentation tank each (15x4x2)
- total surface area - total volume	1,400 m ² 2,100 m ³	224 m ² 336 m ³	282 m ² 562 m ³
construction cost	US\$ 339,000	US\$ 48,000	US\$ 63,000
ditto, per m ³ water produced	US\$ 68 per m ³ /day	US\$ 9.5 per m ³ /day	US\$ 12.5 per m ³ /day
operating cost	US\$ 380 per month	US\$ 1039 per month	US\$ 1933 per month
operating personnel	less skilled	semi-skilled	skilled
sludge withdrawal	manual	manual/semi-mechanized	mechanized
application - pretreatment before	slow sand filter	rapid/slow sand filter	rapid sand filter
- water type	predominantly mineral particles	can be optimized for colour and colloid removal	can be optimized for colour and colloid removal
- socio-economic environment	rural areas	towns and small cities	large cities

In-pore flocculation in DHRF was observed to continue till the end of a 8 m filter (Chapter 2). Due to a supposedly moderate floc breakup the flocculation energy applied could be thought to be more efficiently utilized to promote the floc growth. Although the floc volume concentration decreases along the filter reducing the flocculation efficiency (Harris et al., 1966), the filter grains offer substantially more contact surface for flocculation than a conventional flocculator. Coagulation in the conventional process is generally achieved in the sweep coagulation zone. Coagulation, flocculation and sedimentation occur in three different units.

Design and operation

Suppose a treatment production capacity of 5,000 m³/day is required for serving 50,000-75,000 people in an urbanized area in a developing country. The raw water turbidity is 200 NTU and consist of mainly mineral particles. The three alternative pretreatment designs are considered. HRF is designed based on the HRF manual (Wegelin, 1986), conventional flocculation-sedimentation process on the design guideline applicable for developing countries (Schulz and Okun, 1984; Degremont, 1991) and DHRF on Table 1. The construction and operating costs vary from country to country depending on type of construction material used, availability of materials, supplies and manpower. Cost estimates for the three alternatives designs were prepared based on the 1995 Bangladesh prices applicable for urban areas (Department of Public Health Engineering, 1995) Breakdown of cost components are given in Annexure I. Concrete structure and common civil engineering practice and only marginal costs for the pretreatment components are considered in the estimates.

HRF would require 44 filter units compared to only 7 for DHRF. The total surface area required for DHRF and for the conventional process is in the same order of magnitude, but HRF requires a 7 times larger area. The construction cost per m³ water treated for HRF is US\$ 68 which is comparable to HRF cost estimates by others. Wegelin (1986) estimated the HRF cost to be US\$ 40-60 per m³/day in developing countries, Pardon (1992) reported US\$ 25-50 for Peru. The construction cost of HRF is 5-7 times higher than that of conventional pretreatment or DHRF. However, operating costs are only one third and one fifth of DHRF and conventional pretreatment, respectively. No chemicals are used and power is necessary only to pump against the headloss. The manpower cost component in HRF (which is about 90% of the HRF operating cost) is approximately 3 times higher than for DHRF because of the long working time required for filter cleaning. The filter downtimes for hydraulic flushing and surface washing for HRF and DHRF are estimated to be 1% and 4% respectively.

Compared to conventional pretreatment the construction cost of DHRF is 25% lower. Using low cost structures (e.g. brick masonry in the superstructure) can further lower the cost by

10%. The operating and maintenance cost of DHRF is about half mainly because of lower chemical (coagulant) requirement. The sludge production is expected to be about half than of conventional treatment (Chapter 5). The manpower requirement in DHRF is, however, higher as mechanized filter cleaning process is not considered.

The availability and cost of construction materials and equipment vary from country to country. However, the *relative* cost are applicable for other countries (Schulz and Okun, 1984). Generally, HRF and DHRF can be constructed with locally available materials; the conventional treatment may require some imported items especially for operation and control (e.g. for application in Bangladesh).

Application

HRF will not be feasible in urbanized areas because of its high construction cost and large land requirement. On the other hand, HRF can be feasible for small community water supply in rural areas, usually prior to slow sand filters. The cost of land and labour are there comparatively lower (about 50% and 25%, respectively of those in urbanized areas). Using low cost structures (e.g. brick masonry) the construction cost can be further reduced by 20-25%. More importantly, the level of HRF technology is comparable to that of slow sand filters; operation and maintenance can be performed with less skilled manpower. The HRF option will be favoured because of its simplicity. Other options will require regular supply of chemicals and stricter process control which would be difficult to achieve in a rural situation.

DHRF can be considered feasible for smaller urbanized settlements drawing raw water from heavily turbid water: towns and small cities in developing countries generally have limited finance available compared to large cities but they have more than the rural areas. Coagulant is generally available in urbanized areas (e.g. in Bangladesh Perojpur, Sunamganj and Chandpur towns with population of 16, 26 and 74 thousands, respectively would meet the requirement, whilst rural villages would not have access to regular coagulant supply and skilled labour).

Less operational control is required in DHRF compared to the conventional process. An operating mode (with respect to coagulant dose, mixing condition, etc.) is applicable over a wide range of water quality (turbidity, pH) and thus frequent adjustment is not required (Chapter 6). The operating cost for power and chemicals is also substantially lower than for the conventional process, however higher than for HRF where no chemical is needed. DHRF can also be optimized for removal of organic matter in contrast to HRF. Manual filter cleaning can be feasible for towns. A travelling bridge mounted surface washing system may be developed for more urbanized areas. Thus, DHRF can be operated with semi-skilled

operators which is available in towns and small cities.

Large cities and metropolises where a high capacity is required may still favour the conventional process because it is an established pretreatment technology prior to rapid sand filters. There is also a economy of scale for larger plants (Schulz and Okun, 1984) whereas HRF or DHRF costs would be rather indifferent to scale. Operations can be fully mechanized and even automated; this is desirable for the large capacity plants. The process has to be adjusted for variations in raw water quality. The conventional process asks for a higher degree of operational control for optimal performance and such skilled manpower and facilities are generally available in larger cities.

As a process DHRF is more efficient than the conventional alternative. Filtration in horizontal direction is more efficient than vertical (Chapter 2). Flocculation and particle separation can be combined in a single unit, implying 40% less reactor volume is required (Table 2). Chemical and power requirements are about half. Smaller grains (3-5 mm) or higher coagulant dose (2 mg Al(III)/l) in DHRF have the capability in certain cases to produce effluent turbidity comparable to that after a rapid sand filter (Chapter 5). Appropriate filter cleaning procedures for smaller grains compartments and for other operational conditions are to be further developed so that a complete particle separation can be achieved in a single process line. Further research and development in this direction is recommended. In the present stage DHRF technology appears ready for field validation.

CONCLUSIONS

A tentative design and operation guideline for DHRF has been proposed (Table 1). It is expected to be valid for wide a range of raw water quality. The optimum filter design with respect to its dimensions and grain sizes is mainly limited by the maximum cleaning interval and minimum grain size. The design tables can be used to determine alternative optimum process conditions. Depending on the local costs and other conditions, a least cost solution can be identified.

Column settling test and jar test are useful tools in predicting the DHRF performance for varying raw water quality and process conditions deviating from those used for the pilot-scale experiments.

HRF requires 6-7 times more surface area and construction volume compared to DHRF or conventional flocculation-sedimentation process. The construction cost is 5-7 times higher and thus not feasible for application in urbanized areas where construction, labour and land cost are high. The operating cost is, however, only one third to one fifth than those processes.

As no chemicals are needed and the level of operation is comparatively simple, HRF can be appropriate for rural water supply in small communities for developing countries.

DHRF is a technically, socially and economically appropriate technology for towns and small cities in developing countries. The construction cost is typically 25-35% lower than for the conventional process. The operation is simpler and its cost is about half as coagulant requirement is half that of the conventional alternative. Generally, construction can be done with locally available materials and manpower, goods and supplies for operation and maintenance (e.g. coagulants) are available in the small and medium towns.

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Annexure I

Typical cost components for the three alternative pretreatment designs. Plant capacity 5000 m³/day. Costs are based on 1995 Bangladesh unit prices (Department of Public health Engineering) applicable for urban areas. Conversion rate is 1 US \$ = 40 Bangladesh taka.

A. Construction Cost

description	amount in thousand US \$
HRF	
1. concrete structure	155 (46%)
2. gravel	106 (31%)
3. electrical/mechanical equipment	29 (9%)
4. others (land and site development, utilities, etc.)	49 (14%)
	339 (100%)
DHRF	
1. concrete structure	21 (44%)
2. gravel	14 (29%)
3. electrical/mechanical equipment	5 (10%)
4. others (land cost and site development, utilities, etc.)	8 (17%)
	48 (100%)
Conventional Process	
1. concrete structure	40 (64%)
2. electrical/mechanical equipment	16 (25%)
3. others (land cost and site development, utilities, etc.)	7 (11%)
	63 (100%)

B. Operating Cost (per month)

description		quantity	amount in US \$ per month
HRF			
1.	electric power	455 kWh	40 (11%)
2.	man-power		340 (89%)
	- hydraulic flushing	700 man-hour	
	- manual cleaning (every 2 years)	660 man-hour	
			380 (100%)
DHRF			
1.	chemicals	3.3 MT	858 (83%)
2.	electric power	504 kWh	55 (5%)
3.	man-power		126 (12%)
	- surface washing	280 man-hour	
	- others	140 man-hour	
	(dosing, manual cleaning every 4 years, etc.)		
			1039 (100%)
Conventional Process			
1.	chemicals	6.6 MT	1716 (89%)
2.	electric power	1152 kWh	127 (7%)
3.	man-power	260 man-hour	90 (5%)
			1933 (100%)

CONCLUSIONS

ABSTRACT - The addition of coagulants prior to horizontal-flow roughing filtration appears to be an effective process modification and appropriate technology for towns and small cities in many developing countries.

INTRODUCTION

The demand for safe water is increasing drastically in the urban areas of the developing countries. Many tropical rivers show wide fluctuation in suspended solid content and other water quality; water treatment plants drawing raw water from such rivers are facing growing problem in providing desired treated water quality as well as quantity. This problem for the towns and small cities is compounded due to their limited financial resources. In order to develop an appropriate water treatment (pretreatment) technology the prospect of combining direct filtration with horizontal-flow roughing filtration (HRF) is explored. This study was conducted on lab-scale pilot plant with synthetic raw water, and the following conclusions are made.

CONCLUSIONS

The HRF process was improved to direct horizontal-flow roughing filtration (DHRF): a small and constant amount of coagulant, typically 1 mg Al(III)/l, is rapidly mixed prior to HRF. Some of the limitations of HRF for urban application like low filtration rate could be overcome.

DHRF (8 m long consisting of 4 m first compartment with 20 mm drains and 4 m second compartment with 8 mm grains) systematically yielded good effluent quality (2-5 NTU) with raw water turbidity of 100-400 NTU. It can be applied at a higher filtration rate (3-7 m/h) than HRF (0.5-1 m/h). DHRF run time was 3-15 days depending on raw water turbidity, grain sizes and filtration rate.

The optimum coagulant dose for turbidity removal in DHRF was about 1 mg Al(III)/l. The optimum G value in the rapid mixing unit was 200-300 s⁻¹ with 1 min rapid mixing time. These optimum values were effective over a wide range of turbidity (100-400 NTU) and pH (6.5-9.0). For colour removal in DHRF (~90%) the coagulation condition can be optimized (increasing coagulant dose, lowering pH or both). When colour is present in raw water the filter run time was reduced to about half as compared to the case without colour because of the lower shear strength of deposits.

The use of coagulant increased the filter coefficient λ to about 4 fold in 2 compartment roughing filters as well as in rapid sand filters.

During coagulation of highly turbid water (200 NTU) the zones of coagulation mechanisms (e.g. adsorption-destabilization, sweep coagulation, combination sweep and adsorption-destabilization, etc.) in general are shifted towards higher coagulant dosage as compared to the typical zones for low turbidity water (20 NTU). Coagulation in DHRF is usually achieved in the combination sweep and adsorption-destabilization zone.

The first DHRF compartment with coarse grains acts as a combined flocculator and settler; the second compartment exhibits rather the characteristics of deep bed filtration.

A model to predict the particle separation in gravel bed (> 15 mm) has been developed in analogy to a multiple-plate settler and where both the flocculation and sedimentation processes are incorporated. A procedure was also formulated to use the experimental data from column settling tests with actual suspension in the model for the design and operation of DHRF.

Surface washing was an appropriate roughing filter cleaning procedure. Use of coagulants produced deposit which are easier to clean and require 2-3 times less washwater velocity than when coagulants are not used. Thus in DHRF smaller grain sizes can be afforded; the minimum grain size was 3-5 mm. Long deposit retention time, both with and without coagulants, made deposits increasingly difficult to clean: DHRF should be cleaned at least one a week. The present practice of several weeks of operating time in HRF should be avoided.

A DHRF design guideline has been proposed for standard highly turbid raw water (100-400 NTU) consisting mostly of mineral particles, and under optimum coagulation condition. Depending on the local circumstances a cost optimum design can be made. Jar tests and column settling tests are useful tools in predicting DHRF performance with different raw water quality and under other process conditions.

HRF requires 6-7 times more surface area and construction volume compared to DHRF or the conventional pretreatment consisting of rapid mixing, flocculation followed by sedimentation. The construction cost is 5-7 times higher and thus not feasible for application in urbanized areas. The operating cost is, however, only one third to one fifth of those processes. As no chemicals are needed and the level of operation is very simple, HRF can be appropriate for rural water supply in small communities for developing countries.

DHRF appears to be an appropriate technology for towns and small cities in developing countries. The construction cost is 25-35% lower than for the conventional process. The operating cost is about half as coagulant requirement is typically half. DHRF operation is simpler as frequent adjustment of coagulant dose or pH for changing water quality is not required, and this can be performed by semi-skilled operators. Generally construction can be executed with locally available materials and manpower; goods and supplies for operation and maintenance (e.g. coagulants) are available in the urbanized areas. The required technology level is not as simple as asked for a rural situation but neither as sophisticated as required for large cities. Operation (e.g. filter cleaning) is labour intensive, again not as appropriate as for a rural situation (low labour cost) and neither as mechanized as for a large city (high labour cost).



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