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Iowa State University PH.D. 1981

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Direct filtration using coarse media and dual media filters in series

by

Merrill A. Peterson

A Dissertation Submitted to the
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LIST OF ABBREVIATIONS

Å	Angstrom
cfm/ft^2	Cubic feet per minute per square foot
cm	Centimeter
E.S.	Effective Size
Floc Speed	Flocculation Speed
FTU	Formazin Turbidity Units
g	gram
gpm/ft^2	gallons per minute per square foot
hr	hour
JTU	Jackson Turbidity Units
ℓ	liter
ℓ/hr	liter/hr
m	meter
mm	millimeter
m^2	square meter
m^3	cubic meter
meq	milliequivalent
MGD	Million Gallons per Day
mg/ℓ	milligram per liter
m/hr	meter per hour
min	minute
ml	millliliter
mV	millivolt

n	number of observations
NTU	Nephelometric Turbidity Unit
ppb	parts per billion
psig	pounds per square inch gage
R.M.	rapid mix
rpm	revolutions per minute
R^2	coefficient of determination
TU	Turbidity Unit
U.C.	Uniformity Coefficient
µm	micrometer

INTRODUCTION

During the last decade a great deal of effort has been expended to study and improve water treatment technology. Much of this effort has focused on water filtration. The emphasis on filtration results from the fact that filters are costly to build and operate and from the fact that filtration is the final solids-liquid separation step in water treatment.

There are two basic types of water filtration systems. The first type is termed complete conventional treatment and involves the processes of screening, coagulant addition, mixing, flocculation, sedimentation and filtration. The second type is termed direct filtration and involves screening, coagulant addition, mixing, flocculation and filtration. Conventional treatment has been used to treat all types of raw waters while direct filtration plants have been used to treat raw water supplies that consistently have a low turbidity. This typically has limited direct filtration to the treatment of groundwater supplies and lake or impoundment waters. Most other surface water supplies cannot be treated year around by direct filtration since seasonal rainfall and the resulting runoff create turbidities that are too high to be effectively and or economically removed in the filters.

The problem with high turbidity direct filtration can be related to the solids holding capacity of the filters. The solids holding capacity is affected by the media size, the size of particles to be removed and the interactions between the particles and coagulant and filter media. If the media size is too small, particles will be removed

by straining which will create high headlosses in the upper portion of the filter. This will result in short filter runs and ineffective use of the whole filter. If the filter media are too large or the particles are too small, the particles will escape through the filter producing an unacceptable final turbidity. The selection of an appropriate coagulant is important since this influences the size of particles that can be removed and enhances electrostatic and chemical interactions between the particles and the filter media.

The trend in design of direct filtration systems has been to use dual or multi media filters to increase the solids storage capacity of the filters. The coagulants used are generally alum, cationic polymers or a combination of alum and a cationic polymer or alum and a non-ionic polymer.

Even with advances in media selection and optimization of coagulant usage,most direct filtration plants cannot effectively produce a high quality water under sustained high turbidity loadings. There exists a need to study the direct filtration of high turbidity water. This study was carried out to evaluate the use of coarse media filters followed by standard dual media filters to treat high turbidity waters. The study stressed the use of alum and nonionic polymer as the destabilizing chemicals.

OBJECTIVES

The objectives of the research reported herein are as follows:

1. To compare the optimum chemical dosages as determined by the jar test to the optimum chemical dosages determined in pilot plant studies using the filtration system proposed in this thesis.
2. To evaluate coarse media filters in series with standard, dual media filters as a possible treatment alternative for direct filtration of high turbidity water.
3. To test the optimum combination of chemical treatment and filter operation found with a synthetic suspension in the laboratory with a naturally turbid water.
4. To design a direct filtration system for a raw water source that exhibits seasonal variations in quality.

All but the third objective of this study were met in full. Unfortunately, the third objective was strongly influenced by the weather. An unseasonally mild winter with little snow and intermittent warm periods resulted in essentially no spring runoff in central Iowa. The mild winter with below normal spring precipitation, left little opportunity for the normal source, amount and duration of high turbidity surface water. A field investigation however was undertaken in spite of the relatively low turbidities. Instead of the usual clays and silts comprising surface water turbidity in Iowa spring runoff, the turbidity consisted mainly of diatoms. Even at a low turbidity the diatoms proved difficult to remove.

REVIEW OF THE LITERATURE

Colloid Stability

Particles in natural water systems contributing to water turbidity originate from a variety of sources. These sources include inorganic clays and silts and organic plant and animal organisms. Because of the diverse origins of the particles, the stability of these particles in water is not the same and therefore destabilization may not be caused by the same means. Adin, Baumann and Cleasby (1) point out that not all particles will interact with the same polymer under the same water ionic content and pH conditions.

Particles in water can be generally classified as either hydrophobic or hydrophilic depending upon the affinity of the colloid particle for water. A particle which is preferentially wet by water in competition with oil is termed hydrophilic and a particle which is preferentially wet by oil in competition with water is termed hydrophobic. Hydrophobic particles tend not to dissolve in water whereas hydrophilic particles tend to dissolve in water. The distinction between the particle types is somewhat vague since some particles may exist with various degrees of hydration. The degree of hydration may range from fully covered particles to particles whose surface is almost entirely void of bound solvent molecules (61). Usually hydrophobic soils will adsorb one or two monolayers of water on the particle surface (97).

Clay particles, which comprise a large portion of surface water turbidity, are hydrophobic. Organic particles in water such as plant and animal cells, sewage and other organic pollutants may be classified

as hydrophilic colloids due to their affinity for water. Van Olphen (97) classifies a hydrophilic soil as one that is prepared from organic macromolecular substances such as natural or synthetic gums. These types of colloidal systems should be considered true solutions of macromolecules or macro-ions and consequently hydrophilic colloids are called macromolecular colloids or polyelectrolyte solutions. The "colloidal" properties of the hydrophilic colloids results from the large size of the dispersed molecules with respect to the size of water molecules (97). In a colloidal system containing both hydrophobic and hydrophilic colloids, adsorption of the hydrophilic colloids on the surface of the hydrophobic colloids can take place. The hydrophobic particles may then take on the characteristics of surface charge and electrophoretic mobility of the hydrophilic particles (75). Hydrophobic particles are generally more easily coagulated by electrolytes than hydrophilic particles. This is due in part to the strong adsorptive capacity of hydrophobic particles to adsorb a variety of complex counterions (61).

The stability of hydrophobic colloids can be attributed mainly to electrostatic repulsion and the stability of hydrophilic colloids can be attributed mainly to their affinity for water (97). In addition, for hydrophilic colloids, some degree of stability may be attributed to electrostatic repulsion. For hydrophobic colloids the stability due to hydration is usually disregarded since it is only effective over short distances, usually less than 10 Å. The repulsive energy due to hydration is the energy to desorb the water layers surrounding the particles and is partly responsible for the steep rise in repulsive energy near the particle surface. However, in hydrophobic colloidal systems the

hydration energy involved is not large enough to affect the balance of the double layer and van der Waals interaction energies since particle interactions occur at relatively larger distances from the particle surface. The interaction between particles at large distances from the particle surfaces controls the fate of colliding particles (97).

Electric double layer repulsion is a major stabilization mechanism for hydrophobic and hydrophilic polyelectrolyte particles. The electric double layer consists of a particle surface charge and a compensating counter-ion charge in the liquid near the particle surface. The surface charge on hydrophobic particles originates from imperfections in the crystal structure or by preferential adsorption of certain specific ions. When the charge originates by the adsorption of ions, the ions may be either peptizing ions adsorbed by chemical bonds or potential determining ions identical to those composing the particle. Aluminum hydroxide sols are an example of stabilization by potential determining ions. In simplified terms, either a positive or negative sol is created depending on the availability of Al^{3+} or OH^-. At acid pH values, Al^{3+} is available and positive sols exist whereas at alkaline pH values the OH^- availability is responsible for a negative sol (97).

The surface charge of hydrophilic colloids is attributed to the functional groups on the colloid surface such as carboxyl groups or phenolic groups. Dissociation may occur with these functional groups depending on the pH.

Once the surface charge of a particle is established, a layer of oppositely charged ions will accumulate at the surface. Even though these counterions are attracted by the surface they have a tendency to

diffuse away from the surface to areas where their concentration is lower. Ions of the same charge as the particle surface also exist in the diffuse layer. Near the particle surface a deficiency of these ions exists due to the electrostatic repulsion caused by the surface charge (97). Figure 1 illustrates the electric double layer surrounding a particle.

As one particle nears another, their diffuse layers begin to interfere, which leads to a change in the distribution of ions in each particle's double layer. This collision of double layers results in a repulsion between the particles. The magnitude of the repulsive energy decreases approximately exponentially with increased separation of the particles.

Attractive forces also exist between particles. These attractive forces are termed van der Waals forces and are attributed to the interaction of electric fields brought about by orbiting electrons. The van der Waal's attractive force between atom pairs is additive. The sum of these dipolar interactions between every atom of one particle and every atom of another particle results in a larger total attractive force that decays less rapidly with increasing distance between particles than would be expected. The attractive force for atoms is inversely proportional to the seventh power of distance between the particles and the attractive energy is inversely proportional to the sixth power. For larger particles, the attractive force is inversely proportional to the third power of the distance between the particles and the attractive energy is inversely proportional to the second power. The net interaction energy between particles is illustrated in Figure 2 when electrolyte concentration is low and in Figure 3 when electrolyte concentration

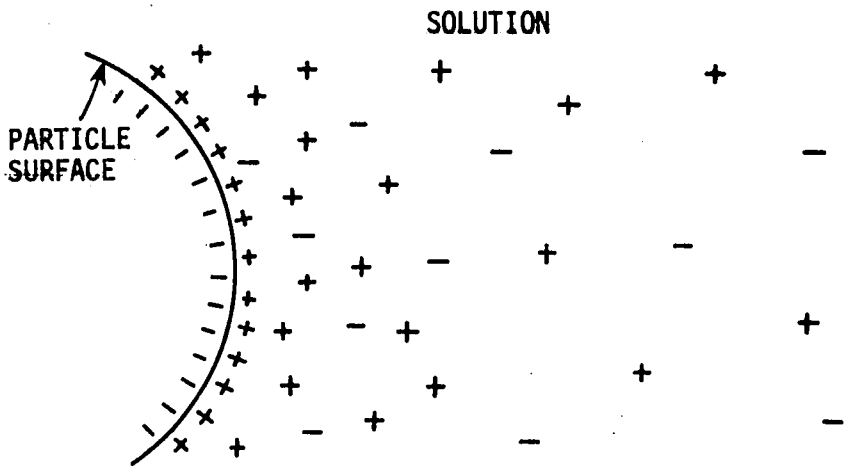


Figure 1. Electric double layer surrounding a particle (97)

is high. The difference between these two figures shows the results of compression of the double by the addition of electrolytes (97).

Destabilization With Aluminum Sulfate

Alum (aluminum sulfate) is probably the most common chemical used for particle destabilization in water treatment processes. Due to its widespread use it is important to understand the chemistry involved in particle destabilization with alum. The aqueous chemistry of aluminum is complex and diverse as a result of the various hydrolysis intermediates formed when alum is added to water. The net product of the hydrolysis reaction depends upon the pH of the receiving water and can be affected by other ionic species present in the water. For example, at a higher pH where the aluminum ions are predominantly hydrolyzed, complex basic aluminum sulfates may be formed (62).

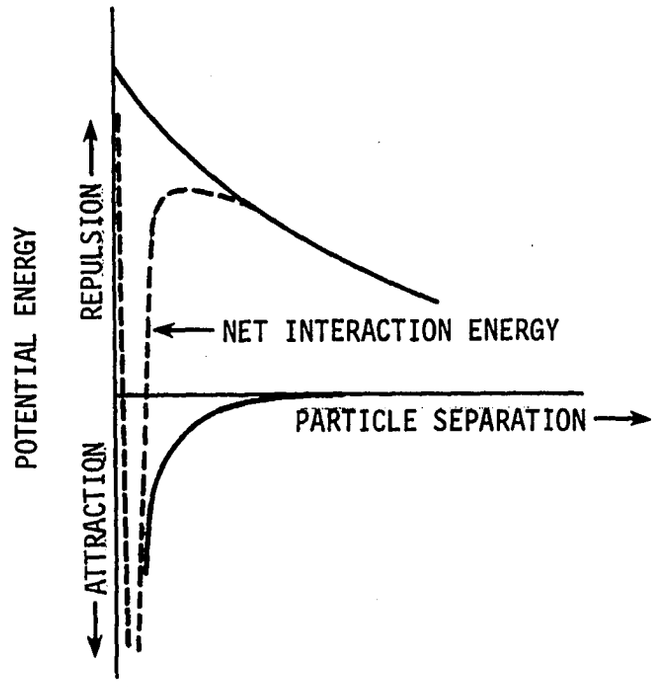


Figure 2. Net interaction energy at low electrolyte concentrations (97)

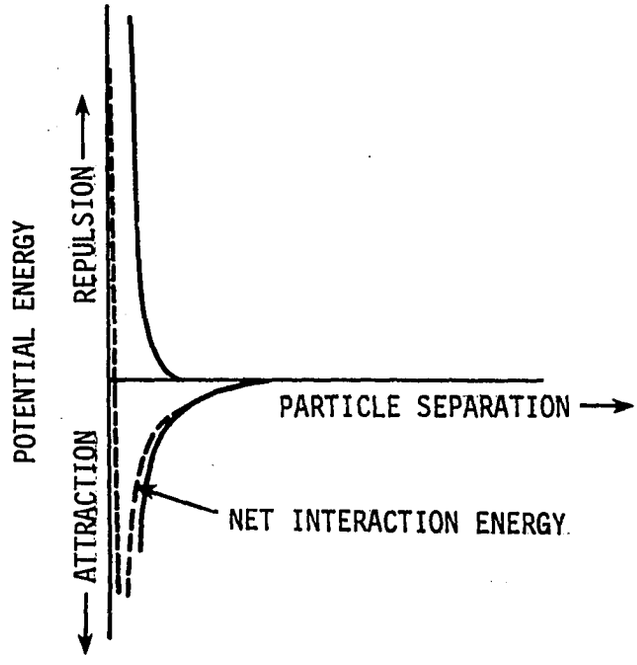


Figure 3. Net interaction energy at high electrolyte concentrations (97)

When alum is added to water the aluminum ion, Al^{3+}, hydrates with six water molecules and forms the aquometal ion $Al(H_2O)_6^{3+}$. The aquometal ion can then react to form several hydrolysis species. The first hydrolysis reaction can be represented by the following reaction:

$$Al(H_2O)_6^{3+} + H_2O \rightleftarrows Al(H_2O)_5(OH)^{2+} + H_3O^{+} \quad (1)$$

As shown in the above reaction, the hydrolysis reaction releases hydrogen ions. This results in a lowering of the water pH as hydrolysis proceeds. Further, hydrolysis may result in the formation of either positively charged, negatively charged or neutral hydrolysis products (3). The products of further hydrolysis can be represented as follows:

$$Al(H_2O)_5(OH)^{2+} \rightarrow Al(H_2O)_4(OH)_2^{+} \rightarrow Al(H_2O)_3(OH)_3 \quad (2)$$

$$Al(H_2O)_2(OH)_4^{-} \rightarrow Al(H_2O)(OH)_5^{2-} \rightarrow Al(OH)_6^{3-} \quad (3)$$

Furthermore, from the above hydrolysis products it is possible to build polymers such as $Al_{13}(OH)_{34}^{5+}$, $Al_7(OH)_{17}^{4+}$, $Al_8(OH)_{20}^{4+}$ and $Al_6(OH)_{14}^{4+}$. These aluminum complexes may be bridged initially by hydroxide ions and upon ageing the hydrogen may split leaving an oxygen bond. This accounts for the reduction in pH of aged aluminum solutions (39). Studies of aluminum hydrolysis indicate that the primary hydrolysis product at alkaline pH is the aluminate ion, $Al(OH)_4^{-}$ (39,63). The aluminate ion is in equilibrium with the aluminum hydroxide precipitate according to the following equation:

$$Al(OH)_3(S) + H_2O \rightarrow Al(OH)_4^- + H^+ \quad (4)$$

The aluminum hydroxide precipitate formed when the solubility limit is exceeded is weakly negative above pH 8 and strongly positive at pH values less than 7 (3). Differing pH values for zero point of charge of aluminum hydroxide floc have been reported (12). The range of pH values reported is 7.7 to 8.4. The zero point of charge of a sol particle is usually not obtained at equal concentrations of the positive and negative peptizing ions but at a slight excess of one of the ions. For an aluminum hydroxide sol, the zero point of charge is not necessarily at a neutral pH where there is an equal concentration of H^+ and OH^- ions (97).

The destabilization of colloids with alum does not conform well with the theory of electrical double-layer interactions as described by the Guoy Chapman model for diffuse layers (87). In order to satisfy the conditions of this model, destabilization with alum would conform to the Shulze Hardy rule and be independent of the colloid concentration. Also, excess amounts of the electrolyte would not produce restabilization of the colloid. For flocculation, the Shulze-Hardy rule is applicable to a variety of sols and indifferent electrolytes if the electrolytes do not engage in any sort of specific reaction with the sol. The electrolytes should not contain potential determining ions for the sol or other ions which adsorb on the sol particles. In addition, the electrolytes should not react chemically with the ions which comprise the electric double layer on the particle surface (97).

Ishibashi (45) studied coagulation with alum using a transmission electron microscope to determine the extent of aggregation of particles.

He found that coagulation of particles occurred even at high zeta potentials which would indicate that adsorption of potential determining ions and bridging better explain the destabilization mechanism than the double layer theory.

The principal mechanisms of particle destabilization using alum are adsorption and enmeshment. Adsorption to produce charge neutralization of negative sols is predominant at pH values below the zero point of charge of the insoluble metal hydroxide, and also dosages and pH values where positively charged soluble hydrolysis species are formed. Above the isoelectric point destabilization by adsorption is accomplished by bridge formation (67). Enmeshment requires high alum dosages for production of a large volume of insoluble aluminum hydroxide floc. Under certain conditions both these mechanisms may work simultaneously. Amirtharajah and Mills (3) summarized the conditions necessary for destabilization by each mechanism. Figure 4 is a culmination of their results and the work of other investigators in this area.

According to Matijevic and Allen (61) metal ions in their unhydrolyzed state will strictly obey the Schulze-Hardy rule but the hydrolyzed species may or may not. The hydrolyzed species will adsorb onto hydrophobic particles much stronger than the hydrated unhydrolyzed metal ion (55, 61, 72). This is especially true at low electrolyte concentrations. Hall (38) found a limiting pH of about 5 which would produce a change in the electrokinetic potential of a kaolinite suspension in the presence of aluminum sulfate. Hydroxyl groups must be present for adsorption but the actual ionic charge has little or no effect. Because of the strong adsorption of the hydrolyzed species, coagulation can occur at electrolyte concentrations less than those predicted by the

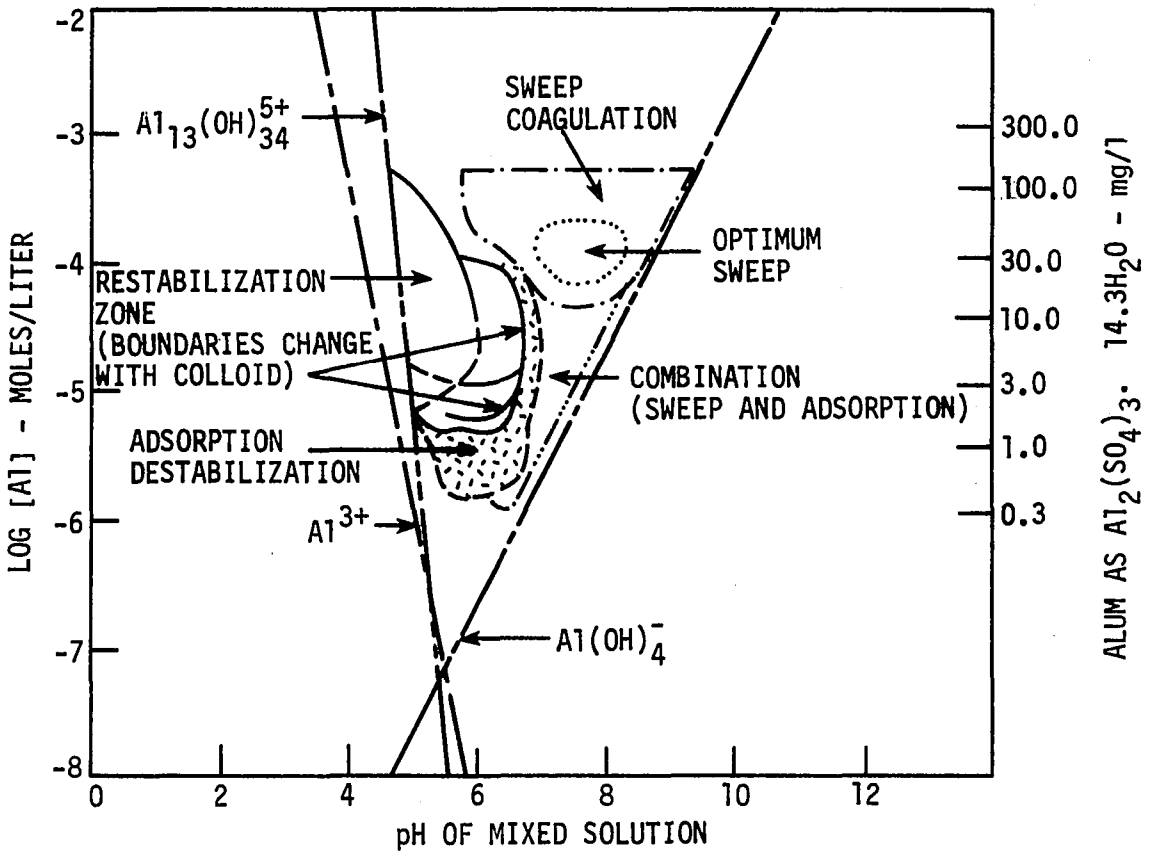


Figure 4. Design and operation diagram for alum coagulation (3)

Schulze-Hardy rule.

Stumm and O'Melia (87) concluded that the double layer model is restricted to lyophobic colloids and simple electrolytes since it neglects the dominating role that chemical forces may have in causing adsorption. These chemical interactions between the particle surface and adsorbate may include hydrogen bonding, coordination reactions, covalent bonding and ion exchange reactions.

Further evidence of the inadequacy of the double layer theory was reported by Rubin and Blocksidge (78). According to the theory, ions of the same sign as the colloid should have essentially no effect on destabilization, but significant effects have been shown to occur. When using aluminum or iron salts as coagulants, the presence of sulphate widens the pH range of effective clarification to the acid side while phosphates lower the entire range to the acid side without significantly broadening it. It should be noted, however, that phosphate is a potential determining ion for some clays, especially montmorillonite.

Enmeshment is an important destabilization mechanism when alum is used. This mechanism involves high concentrations of alum to form a settleable precipitate which combines with the particles in the water. It should be noted that formation of a precipitate does not in all cases lead to settleability of the precipitate. Precipitation with alum occurs upon an increase in the solution pH. When the hydroxide to aluminum ratio is less than 2.75, the precipitate is X-ray amorphous. At a ratio of 3.0 the precipitate is crystalline $Al(OH)_3$ in various forms. Examples are bayerite and gibbsite. At ratios above 4.0 the precipitate will dissolve forming the aluminate ion (78).

The settleability of a precipitate is dependent upon its particle size, structure and surface properties (78). The precipitate has to be large enough to settle or a stable sol may be formed that is subject to the stability forces of a colloidal system. Amorphous precipitates have no orderly geometric structure and are very flocculent and fluffy. The crystalline precipitates are geometrically structured and will settle as a dense compact mass if they are of a size large enough so as not to be colloidally dispersed. The stability of the precipitate is due mostly to its charged surface. The surface charge arises from preferential adsorption of hydrogen or hydroxide ions, adsorption of common ions or ionization of surface groups. Precipitates, therefore, can be coagulated with electrolytes in accordance with the Shulze-Hardy rule with best settleability occurring around the pH of its isoelectric point.

There are several factors which may determine the ability of alum to destabilize particles. The solution pH and alum concentration which are the most important factors have been discussed previously (3, 78). Two other important factors are the surface area and type of particles to be stabilized. The effect of particle concentration has been discussed extensively by Stumm and O'Melia (87) and by others (3, 67). Restabilization boundaries shown in Figure 4 will vary slightly depending on the surface area of the colloid. Low colloid surface area may require more alum than when the surface area is higher. With medium particle surface areas, a stoichiometric relationship for coagulation of hydrolyzed metals is often reported. High surface area concentrations may require high concentrations of alum above that required for sweep floc. At these concentrations, both sweep floc and adsorption are important.

The type of particle and the resulting destabilization with alum can be considered in two parts. Hydrophobic particles are more easily coagulated than hydrophilic particles. Disagreement on the amount of alum required to destabilize inorganic particles often occurs in the literature. This disagreement is often times attributed to the cation exchange capacity of the particle (52, 70). The lower the cation exchange capacity, the less alum required. Evidence contrary to this is demonstrated by Amirtharajah and Mills (3) and supported additionally by the work of others (12, 71).

Destabilization with Polyelectrolytes

Organic polymers are long chain molecules comprised of many monomeric units which may or may not possess an ionic charge in solution. Organic polymers may be either natural substances or synthetic. Some of the natural polymers are starch and starch derivatives, cellulose compounds, polysaccharide gums and proteinaceous materials. Functional groups on organic polymers may include quaternary ammonium groups, polyacrylamides, polyamines and dimethyldiallyl ammonium groups. The ionization of the functional groups controls the charge on the polymer in solution. Polymers may be positively charged (cationic), negatively charged (anionic) or neutrally charged (nonionic). Synthetic polymers can be manufactured with various molecular weights. The molecular weight will affect the required polymer dosage since the molecular weight is generally related to the number of functional groups and the physical size of the polymer molecule. The size of the polymer molecule is important in regard to the probability of interaction of the polymer and particles in suspension.

Two types of adsorption mechanisms are used to describe particle destabilization with polymers. One mechanism is adsorption resulting in charge neutralization. The other mechanism is adsorption resulting in a polymer bridge between particles or between particles and filter media.

Destabilization by charge neutralization refers to adsorption of the polymer at the solids-liquid interface resulting in a decrease in the particle charge. Excessive amounts of polymer can result in particle restabilization by reversal of charge (9, 11, 37, 75, 100). This mechanism is most often used to describe destabilization with cationic polymers, particularly low molecular weight cationic polymers (31, 35). High molecular weight cationic polymers may also destabilize by bridging in combination with charge neutralization (11, 53, 56). O'Melia (66) found that the optimum polymer dose for destabilization of negative latex particles occurred when the particle zeta potential was negative. Restabilization occurred before the charge on the latex particles had been reversed. This evidence indicated that charge neutralization could not be the sole destabilization mechanism.

Gregory (35) described an electrostatic patch model to explain destabilization with cationic polymers. This model explains the effectiveness of cationic polymers differing in molecular weight but having the same optimum concentration for flocculation. Some limitations of this model are that it may not apply to very high molecular weight polymers and to high suspension concentrations. The polymer is adsorbed on the particle surface in uneven patches so only a portion of the particle is coated with polymer. This produces a particle having both positive and negative charges. Support for the patch model was reported by Lindquist and

and Stratton (56).

The polymer bridging mechanism was first proposed by Ruehrwein and Ward (79) for destabilization of clay suspensions. Particle destabilization by polymer bridging occurs when one end of an extended polymer chain becomes attached to a particle. The unattached segment extends out into solution making possible further interaction with another particle or with filter media. The growth of the particle chain is limited only by the shear gradient imposed by agitation in the system and by the amount of polymer initially adsorbed on the particle surface (11).

LeMer and Healy (53) developed a mathematical model for the polymer bridging mechanism. In their work with uranium ore slimes, they found that the optimum polymer dosage for the most efficient filtration was one that would provide about 50 percent coverage of the suspended particles. Excessive amounts of polymer resulted in a more dispersed system that was less filterable. According to the bridging model the requirements for flocculation are:

1. extended segments must be available,
2. the segments must be of sufficient length, and number, and
3. free surface for bridge sites must be available.

A flocculated system passes to a dispersed one if: (1) the surface becomes so covered that there is insufficient free surface for bridging, (2) the extended polymer molecule is so long as to be unable to overcome the effect of Brownian motion on individual floc particles, and (3) the extended segments physically interfere with one another (53).

It is generally accepted that destabilization of negative colloids with anionic and nonionic polymers functions through the bridging

mechanism. A minimum molecular weight on the order of one million appears to be necessary when nonionic or anionic polymers are used to bridge two negative colloid particles (67). The minimum size will depend upon the number of charged groups, degree of polymer branching, colloid charge and the ionic strength of the solution.

Birkner and Edzwald (9) have theorized that destabilization of clay with nonionic polymers occurs by hydrogen bonding. At low pH values negative aluminum and silicon oxide groups may provide additional adsorption sites. LaMer and Healy (53) point out that even though the strength of an individual hydrogen bond is relatively weak, a polymer attached by many hydrogen bonds would be strongly attached.

Black, Birkner and Morgan (11) demonstrated that destabilization of clay with anionic polymers is influenced by coulombic forces. High concentrations of calcium ions were necessary to flocculate a clay with an anionic polymer. They proposed three ways in which calcium ions may affect anionic polymer and clay systems:

1. the divalent ions compress the diffuse layers surrounding the clay particles and thereby reduce the repulsive forces,
2. the ions reduce repulsive forces between the anionic polymer and the particles and thereby enhance polymer-clay attachment, and
3. the ions reduce repulsive forces between attached polymer segments thereby facilitating bridge formation.

Metal ions may also reduce the interaction between polymer molecules

and thereby enhance increased polymer adsorption onto a particle (67).

Stumm and O'Melia (87) emphasize that coagulation and restabilization can be significantly influenced by adsorption and that chemical interactions can outweigh electrostatic effects. At close ranges the chemical forces often overwhelm the electrostatic forces of attraction and repulsion. This explains why negatively charged particles are often found to be best aggregated by negatively charged polymers.

Filtration

The removal of colloidal and suspended particles from a water by granular media filtration involves three principal mechanisms. These mechanisms are transport, attachment and detachment. The transport step involves the movement of the particles into and through a filter pore so that the particles come into close proximity with the filter grains and/or previously removed particles. Particle attachment involves contact between a particle and either a grain of the filter media or particles previously removed which results in the particle being retained in the filter bed. The detachment mechanism involves the effects of the hydrodynamic shear forces of the flow which cause previously attached particles, less strongly linked to the others, to detach and move deeper into or through the filter (5).

Transport mechanisms

Important transport mechanisms may include gravitational settling, diffusion, interception, inertial forces and hydrodynamics (17, 40, 41).

Physical factors which can influence particle transport are media size, filtration rate, fluid temperature and the density and size of the particles to be removed (69). The size of the particles may have the most effect on particle transport in water filtration (5, 67, 69). Yao, et al. (100), found that there exists a critical size at which the suspended particles have a minimum contact opportunity and therefore a minimum removal efficiency. The critical suspended particle size is about 1 μm. With suspended particle sizes above 1 μm, removal efficiency increases rapidly with particle size. This is due to the enhancement of the sedimentation and/or interception transport mechanisms. Sedimentation removal increases with the square of the particle size (68). As for interception, a larger particle will have a greater chance of contact than a smaller particle. At a suspended particle size below 1 μm, removal efficiency increases inversely with particle size due to the enhancement of removal by diffusion (100).

There is a similarity between particle transport in coagulation processes and in filtration processes. In both processes particle transport can be considered separately in two distinct regions (e.g., orthokinetic and perikinetic flocculation). Diffusion controls the transport rate in both processes when the particles are smaller than about 1 μm, (67).

Attachment mechanisms

The attachment step is affected by both physical and chemical factors which may involve electrostatic interactions, chemical bridging or specific adsorption. All of these modes of attachment are affected by

the coagulants applied in the pretreatment and by the chemical characteristics of the water and the filter media (18). In order for attachment to occur, the electrostatic repulsion between the particles and filter media must be reduced or altered. One way to lower the electrostatic repulsive forces is the addition to the water of metal salts and/or synthetic or natural polymers which change the electrostatic forces around the particle. A slightly negatively charged or zero charged particle can then come into contact with the filter media and an interaction (electrostatic, chemical, or van der Waals) can take place. In some cases, a chemical bridge may be formed when the polymer is adsorbed at one end onto a particle and at the other end onto the filter media. Since the usual filter media have a net negative charge at pH values common to water treatment processes, it is possible to change the charge on the particles to be removed to positive values so that the positive particles will be attracted to the negative media. In some cases, it is possible to have successful particle removal when the treatment involves adding a negative polymer which provides more readily accessible negative sites on the particle. The particle can then attach to the filter media by chemical links formed with specific ions such as calcium and magnesium (66).

Detachment

The role of detachment as a mechanism in filtration is a controversial subject. Mints believed that particles deposited on filter grains are subject to scouring away by the flow back into suspension at a rate

proportional to the quantity of deposit on the grains (98). According to Mints the rate of particle deposition, in terms of efficiency, is constant throughout the filtration process. Local increases in fluid velocity as filter pores become clogged affects the scour rate. The increase in fluid shear causes previously attached particles to be re-entrained in the flow. At equilibrium the rate of scour is equal to the rate of deposition and consequently the concentration of suspension entering and leaving any filter layer does not change. Mints supported his theory with experimental observations based on the success of polymers in improving filtration efficiency by increasing the strength of the deposit and that larger particles were leaving a filter than were entering a filter. The larger particle size was attributed to shearing off of groups of particles.

There is little doubt that detachment of particles in a filter does occur when the overall flowrate increases, especially if this increase is sudden. This type of detachment phenomenon was reported by Cleasby et al. (19), and by Tuepker and Buescher (96).

Ives (47) and Mackrle and Mackrle (60) opposed Mints' detachment theory. Ives attributed the variable efficiency of filtration to geometric changes in the filter pores caused by particle deposits. The deposits change the amount of surface available for deposition which changes the flow pattern and in turn causes locally increased pore velocity. At equilibrium there is no particle retention because of the reduced surface area and the high velocities sweep the particles through

the pores before they can be attached. Mackrle and Mackrle (60) supported this theory. These investigators found that a filter would produce a clear water with no change in headloss distribution after a period of particle removal if the influent particle suspension was changed to clear water without a change in flowrate. Stanley (85b) found that no change in deposition within a filter occurred when radioactively labelled particles were substituted with the same particles not radioactively labelled.

Ironically, Ives and his students observed particle detachment in a filter with scaled up media. Using a suspension of kaolin and kaolin flocculated with ferric sulphate, it was observed that detachment with kaolin alone was insignificant but with the flocculated kaolin detachment was much more frequent, particularly when suspended particles collided with unstable high domes of deposit (76).

Adin et al. (1) stated that the transport step may be relatively insignificant in filter design when compared to the attachment step. Transport mechanisms will exist in all filters and a minimum transport efficiency does not imply that it is necessary to increase the efficiency to meet filtration goals. O'Melia (66) has concluded that transport mechanisms are sufficient to remove particles that would be expected to have minimum transport efficiency. With the right chemical treatment, particles ranging in size from 0.01 μm to 100 μm can be effectively removed in filtration. Adin et al. (1) emphasized that attachment should be viewed as the major factor influencing filtration processes and that improvement in filtration be directed in this area.

Coarse Media Filters

The definition of a coarse media filter may encompass a wide range of media sizes and depths. However, most coarse media filters can be characterized as having a media size greater than 1 mm and a depth greater than about one meter. In the United States coarse media filtration is rarely used. European countries on the other hand have used coarse media filters for many years for both water and wastewater treatment (49).

The most common coarse media filter for potable water treatment in Europe utilizes a uniform layer of media with an effective size of 0.9 to 1.35 mm. Sometimes a coarser media, 1.35 to 2.5 mm, is used for coarse contaminant removal from industrial wastes or for tertiary effluent treatment. The media depth is approximately 1.5 m. Pretreatment ranges from no chemical addition to the use of metal salts in combination with polymers. Cationic polymers are sometimes used as the sole coagulants. The usual range of alum dosage for potable water treatment is 2 to 10 mg/l with a maximum of 15 mg/l. Both influent turbidity and color must be fairly low and uniform to use with these types of filters when a high quality effluent is desired. Coarse media filtration of water for use by industry with coagulation in the filter is sometimes used to treat river water with variable suspended solids levels (28).

The European style coarse media filters are backwashed with a combination of air and water. Media expansion does not occur during backwash so that the media will remain homogeneous. The first stage of backwashing

uses a high rate of air flow in combination with a low rate of water flow. The second stage uses a medium water flow rate as a rinse. The filters are operated at filtration rates between 7 to 10 m/hr for high quality applications. Usually, the terminal headloss is limited to less than 2 m to avoid excessive filter clogging. The filters are usually of the downflow gravity type (28).

Calise and Homer (15) reported on the development and testing of coarse media filters in Russia in 1953-54. Tests indicated that for certain waters the Russian "contact clarifier" produced an effluent equivalent in quality to conventional sedimentation and filtration. The Russian coarse media filter was an up-flow type. The media depth was about 2.6 m and the flow rate was approximately 17 m/hr. No detailed operating information was available and no media sizes were given. However, it was stated that water supplies with turbidities greater than 150 "mg/1" were not suited for coarse filter applications.

Jung and Savage (49) reported on the operation of a pressure type coarse media filtration system treating Ohio River water for industrial use. The media depth was about 2.4 m and its effective size was 1.2 mm with a uniformity coefficient of 1.3. The only treatment prior to filtration was coarse screening and the addition of 1.0 mg/1 of a cationic polyelectrolyte prior to the filter. The filtration rate during the reporting period was about 13.5 m/hr. The effluent from the system averaged less than 1 JTU. Operating data were given for the months of August, September and October of 1970. Ninety-two values of influent turbidity were given and of these 28 values exceeded 15 JTU. During a

nine-day period in September, the turbidity ranged from 33-130 JTU whereas the effluent turbidity exceeded a value of 1.0 JTU on only one day. The highest effluent turbidity reported during the three month period was 4 JTU when the influent turbidity was only 12 JTU.

In the United States, the most serious consideration given to the use of coarse media filtration has been in the treatment of wastewater. O'Melia (66) conducted an extensive investigation to evaluate chemical requirements and media sizes necessary for direct filtration of a trickling filter effluent. O'Melia determined that the optimum chemical pretreatment was a combination of alum and an anionic polymer. Alum dosage was in the range of 150 mg/l (for increased phosphorus removal) and polymer dosage was about 1 mg/l. The results of this study showed that a media size of 5 mm produced optimum results. Breakthrough occurred at about the same time as terminal headloss was reached. Media depth was 0.84 m and the flowrate was 4.88 m/hr.

From this study, O'Melia made three important and at times controversial conclusions that apply not only to coarse media filters but to almost all types of granular media filtration systems:

1. There exists a ripening period where the filter effluent is poor immediately after the filter run commences. In a properly designed filter, this ripening period will be short and an acceptable effluent will soon be established. The best quality of effluent may however be produced just prior to run termination. O'Melia recommends that this ripening period may be shortened by precoating the filter media with polymer prior to beginning a

filter run. This can be done by adding polymer to the backwash water.

2. The optimum dose of coagulant for filtration is the same as the optimum dose in a jar test. Often times it is found that optimum conditions in jar tests require more coagulant than for filtration. O'Melia feels this discrepancy is due to adsorption of coagulant on the glassware.
3. Cationic polyelectrolytes can be used as the sole coagulant and produce an acceptable filter effluent. The major drawback to this practice is the high dosages needed and the high rate of headloss that results from these high dosages.

A study performed at Iowa State University by Biskner (10) compared the use of a coarse media filter in series with a standard dual media filter for treatment of trickling filter effluent. The coarse media filter consisted of 0.76 m of either 1.9 mm or 2.45 mm anthracite coal. The dual media filters used 0.45 m of coal and 0.30 m of sand. The effective size of the coals used were 1.45 mm and 0.98 mm. The effective sizes of the sand used were either 0.38 mm or 0.22 mm. Flow rates evaluated during the study were 4.88, 7.32 and 9.76 m/hr for the dual media filters. The coarse media filter was operated at twice the dual media flow rate. The influent wastewater turbidity was in the range of 30-50 FTU. No chemical pretreatment was used.

The results from this study indicated very little difference in the quality of effluent produced from the two stage system compared to the single stage system. The most obvious difference in the two systems was

in the rate of headloss development. Headloss was distributed more evenly throughout the dual media filters that were preceded by the coarse media filter. Also the terminal headloss in the single stage dual media filter was 35% to 65% higher than that in the two stage filters.

A study similar to Biskner's was performed by Dahab (27). Dahab concluded that an unstratified media size of 1.34 to 2.31 mm produced essentially the same effluent quality as conventional dual media filters. Filter run lengths were 3 to 4 times longer with the coarse media filters than with the dual media filters, yet the quality of the effluents were comparable.

The literature contains limited information concerning the application of coarse media filtration to water treatment in the United States. A study by Cleasby and Saleh (18) at the Iowa City, Iowa, water treatment plant utilized a Culligan type water filtration system proposed by Baumann in which coarse media filters are used in series with dual and multimedia filters. The water source was the Iowa River. The coarse media was sand with an effective size of 2.17 mm and a uniformity coefficient of 1.28. The media depth was 1.37 m. The water was pumped from the river and passed through a centrifugal separator to remove sand and other large particles before proceeding to the filter system. Chemical addition points were in the feed line to the filters. Some flocculation time was provided in the feed line to the filters, however no flocs were visible entering the first stage filters. Chemicals could also be added after the first stage filter. The chemicals used during the study were either alum or a cationic polymer, Culligan F-86. Alum

dosages varied between 5 and 30 mg/l and polymer dosage varied between 2 and 30 mg/l. The flow rate was either 9.27, 15.4 or 20.5 m/hr.

Cleasby and Saleh (18) were able to produce a water with a final turbidity less than 1.0 NTU when the influent turbidity averaged as high as 123 NTU. At times, however, a final turbidity of less than 1.0 NTU was difficult to obtain at high flowrates when the influent turbidity averaged as low as 17 NTU. These investigators reported a long ripening period for the two stage system. Evidence from a previous phase of this study where the influent was taken from the Iowa City treatment plant clarifier where alum had been added indicated that a combination of alum and polymer may shorten the ripening period. Overall, the use of the cationic polymer alone produced a better effluent quality than alum alone.

Trussell et al. (95) reported on tests conducted by the Los Angeles DWP for treatment of Owens River water using two 1.8 m deep beds of 1.66 mm sand. One filter media was rounded and almost spherical while the other was angular. At filtration rates of 22 and 29 m/hr the angular media gave better results in terms of both run length and effluent quality. For all cases, the effluent turbidity was less than 0.4 TU and the influent was about 10 TU. Two milligrams per liter of cationic polymer was used as the coagulant. The Los Angeles plant is being designed using deep bed, single media coarse filters to be operated at a rate of 22 m/hr.

Some investigators have used uniform coarse media filters to study the behavior of deep bed filtration. Hsiung and Cleasby (41) used a uniform sand media to develop empirical relationships for prediction of filter performance for design of rapid sand filters. Tchobanoglous and

Eliassen (91) used uniform sand filters to study the filtration of sewage effluents. They concluded that sand size has a more pronounced effect on filtration performance when compared to filtration rate. Decreases in filtration rate resulted in more removal in the upper portion of the filter. For their application of filtration, straining was a dominant removal mechanism. Deep bed filtration removal mechanisms were studied by Ison and Ives (46) using uniform media pilot filters. Their results indicated that gravity settling within the filter was an important removal mechanism. They concluded that, for their system, transition from removal by sedimentation to removal by diffusion probably occurred at a suspension particle size between 0.5 to 2.0 μm. Ling (57) also studied filtration with uniform sand media. He concluded that the filtering mechanisms in a uniform sand filter were the same as those in a graded sand filter. The upper portion of a filter removes the most turbidity during the first stages of a filter run. As the run continues, the removal is shifted downward into the lower portion of the filter. In a uniform sand filter the filter run length varies directly with the sand size and inversely with the filtration rate. The sand size has the greatest overall effect.

Dual Media Filters

Rapid sand filters came into widespread use during the early 1900s. These filters were typified by an average media effective size of 0.5 mm and a filtration rate of 4.9 m/hr (2 gpm/ft^2). The single media in these filters was graded hydraulically during backwashing. The finest media

accumulated at the top and the coarsest at the bottom. This hydraulic grading resulted in most of the applied particles being removed in the top few centimeters of the filter (5).

One of the first investigators of dual media filters was John Baylis at Chicago's South District filter plant (6, 7). In 1935, Baylis experimented with a dual media filter where a two inch layer of anthracite had been placed over the conventional sand filter. The experiment was abandoned due to excessive anthracite loss during backwash. Baylis attempted a second study in 1939 in which he used 4 inches of 1.5 mm anthracite over 20 inches of 0.5 mm sand. With a flow rate of 3.5 gpm/ft^2 he was able to increase filter run length by a factor of 4 times over similar runs using sand alone.

More recent advances leading to present designs using dual media filtration were made by Conley (22). In his work at Hanford, he was given the assignment of designing a filter operating at 6 gpm/ft^2 that would produce an effluent water with a turbidity less than 0.5 JTU with a run length of at least six hours. The initial investigations used single media sand and anthracite filters but both proved to be unsatisfactory. Dual media anthracite-sand filters were then evaluated and proved to be satisfactory for producing the desired water quality. With the dual media filter, the media were hydraulically graded coarse to fine during backwash. The coarse anthracite removed the larger particles that would usually clog the finer sand layers.

Conley (22) also was one of the first to use synthetic polymers in dual media filtration. The polymers were needed to control the

occurrence of solids breakthrough due to the high filtration rates and the increased water viscosity during cold weather periods.

One of the major areas of emphasis in filtration research since the 1960s has been in filter media design to increase the solids holding capacity of filters. This research has lead to present filter designs using dual and multimedia filters. These filters have almost completely replaced the single media sand filters used in potable water treatment plants since the early 1900s (4).

With dual or multimedia filters, run lengths can be extended up to five times or more compared to sand filters operated at the same filtration rate. Also the dual or multimedia filter can provide run lengths of 24-36 hours at normal terminal head loss at flow rates three or more times the rate used in sand filters (4).

Dual media filters commonly use anthracite coal on top of silica sand. Multimedia filters, in addition to the coal and sand, may use garnet sand and plastic media. Table 1 shows typical media characteristics of filter media in present day use. Additional information concerning the design and operation of dual and multimedia filters can be obtained from excellent reviews by Baumann and Cleasby (5, 17).

Direct Filtration and Complete Conventional Treatment

A successful water treatment process provides a potable water that is both esthetically pleasing and possess minimum health hazard to the consumers. A major raw water source for potable water in the United States is surface water. Surface water includes water in rivers, streams,

Table 1. Typical dual media and multimedia characteristics for potable water filtration (4)

Media Characteristics	Dual-Media Filter	3-Media Filter	4-Media Filter
1st Upper Media			
Type	Anthracite Coal	Anthracite Coal	Plastic
Size,[a] mm	1.0-1.5	1.0-1.5	2.0-4.0
Depth, in.	15-18	15-18	3-6
Density	1.35-1.75	1.35-1.75	1.1-1.2
2nd Media			
Type	Silica Sand	Silica Sand	Anthracite Coal
Size,[a] mm	0.5-0.6	0.5-0.6	1.0-1.5
Depth, in.	12-15	12-15	15-18
Density	2.65	2.65	1.35-1.75
3rd Media			
Type	---	Garnet Sand	Silica Sand
Size,[a] mm	---	0.2-0.3	0.5-0.6
Depth, in.	---	3-5	12-15
Density	---	4.0-4.2	2.65
Lower Media			
Type	---	---	Garnet Sand or Illmenite
Size,[a] m	---	---	0.2-0.3
Depth, in.	---	---	3-5
Density	---	---	4.0-4.2

[a]Effective size.

lakes and reservoirs. These raw waters must be treated to remove undesirable quantities of color, turbidity and taste and odors. Standards for potable water quality have been in effect for many years. Prior to 1962, the turbidity of a finished water could not exceed 10 turbidity units. In 1962, under the recommendation of the United States Public Health Service, the turbidity limit was lowered to 5 turbidity units and color level was limited to no more than 15 units. In June, 1977, the National Interim Primary Drinking Water Regulations made it mandatory that water utilities using surface water supplies, treat the water to a finished water turbidity not exceeding 1.0 turbidity unit. At least one state, California, requires that potable water turbidity not exceed 0.5 turbidity units. In addition, in 1968 the American Water Works Association established a goal of 0.1 turbidity unit or less for finished water turbidity. The implementation of the 1977 standard had a significant effect on the water treatment practices in many large metropolitan areas which obtain their water from impoundments. Prior to 1977 many installations were only disinfecting the raw water since its turbidity was less than 5 turbidity units. With the implementation of the 1977 standards, these metropolitan areas must provide a treatment system which includes filtration to meet the standard.

Two major treatment systems are presently utilized for surface water treatment. One system can be termed complete conventional treatment. This type of treatment includes screening, mixing, flocculation, sedimentation and filtration. Complete conventional treatment can produce a high

quality finished water insofar as color and turbidity are concerned regardless of the raw water quality. The objective of coagulation and flocculation in this system is to produce large dense floc particles which will be primarily removed in the sedimentation step. The filters are used to capture particles not removed by sedimentation.

The other treatment system is termed direct filtration. The unit processes used in this system include screening, mixing, flocculation and filtration. Some systems may use little or no flocculation except that which takes place in the filter media itself and are often termed contact flocculation systems. The purpose of coagulation and flocculation in direct filtration systems is to produce a small dense floc amenable to depth filtration. The filters are solely responsible for particle removal in this system. As such, a relatively good raw water quality is necessary when direct filtration is to be considered as a treatment alternative. Direct filtration systems have come into increasing use since the early 1970s. This is due to two basic reasons. The first reason is the implementation of the 1977 turbidity standard. Many utilities which had been meeting the 5 turbidity unit standard could use direct filtration to meet the lower level imposed in 1977 more economically. Direct filtration offers the advantage of lower capital cost due to elimination of sedimentation and in most cases a reduction in operating costs through lower chemical and sludge disposal requirements. The second reason can be attributed to increased technology which accounted for a better understanding of the nature and affects of

turbidity causing particles. In some instances low turbidities are not synonymous with acceptable levels of certain contaminants. A good example of such a case is the problem with asbestos fibers at Duluth. Asbestos is a suspected carcinogen. During an 18-month study at Duluth, the raw water turbidity was 1 turbidity unit or less for most measurements (58). At the same time asbestos fiber (amphibole) counts in the drinking water were most often in the range of 10^7 to 10^8 fibers per liter. A complete conventional plant which could also be operated as a direct filtration plant was constructed to reduce the asbestos fiber level by maintaining a finished water quality of 0.1 turbidity unit or less. Direct filtration is used not only because it provides the desired treatment, but it can do so more inexpensively than when using complete conventional treatment.

Kavanaugh et al. (50) and Kavanaugh et al. (51) found that many undesirable trace pollutants such as trace metals, toxic organic material, and viruses are usually associated with colloidal or coarse colloidal (>0.4 μm) particulates. These particles would not contribute greatly to turbidity.

Raw water turbidity

While complete conventional treatment useage is not limited by raw water turbidity, the use of direct filtration systems generally requires a low turbidity raw water source. Early proponents of the direct filtration process considered the maximum turbidity for direct filtration to be 100 turbidity units (23, 73). Conley (21) reported on a design for a water plant on the Columbia River. The design provided for flash

mixing of chemicals followed by filtration. The design turbidity was 260 units and the maximum color was 30 units. The effluent turbidity was expected to contain less than 5 units of color and 0.5 turbidity units. Many of these early designs were developed without the aid of pilot plant work. Conley and Hsiung (23) reported on a direct filtration plant designed for use on a raw water with a maximum turbidity of 100 units. Even at a raw water turbidity of 25 turbidity units and a filtration rate of 12.2 m/hr, the filter runs were only 12 hours long. Settling facilities were later added to extend filter run lengths. Other early investigators found that direct filtration was a viable alternative when the raw water turbidity was less than 25 turbidity units (77). Culp (26) proposed preliminary guidelines for the selection of direct filtration as a treatment alternative. According to Culp, direct filtration is possible when, 1) the raw water turbidity and color are each less than 25 units; 2) the color is low and the maximum turbidity does not exceed 200 turbidity units; or 3) the turbidity is low and the maximum color does not exceed 100 units. In any case, pilot plant studies should be performed to insure applicability. Baumann (5) indicates that direct filtration may be possible when raw water turbidities are 50 to 60 turbidity units. In Europe, direct filtration is considered a viable alternative up to an influent solids concentration of 50 mg/l (28).

In a study in Canada, Hutchison (43) concluded that direct filtration was a viable treatment alternative under three conditions. These conditions were, 1) the turbidity was low enough to allow alum

dosages of less than 12 mg/l on a continuous basis; 2) the color level was below 25 units when alum was used or when polymer could be used at higher color levels, and 3) diatom levels were less than 1000 asu/ml.

A survey by the AWWA Direct Filtration Subcommittee reflects the worldwide status of direct filtration (20). The survey reports on the operation of approximately 70 direct filtration plants. The committee concluded that problems could be expected when raw water color exceeded 30-40 units or when turbidity exceeded 15 FTU on a continuous basis. On a short term basis, problems may be alleviated with the use of polyelectrolytes either substituting for or in addition to the primary coagulant. Also, changing mixing energies to alter floc size and strength could increase filter particle retaining capabilities. Some plants operated well at turbidities up to 250 FTU and alum dosages up to 80 mg/l. The committee reported a case study on a direct filtration plant for an industrial application where the raw water turbidity was continuously in the range of 75 to 100 FTU. With alum, the plant consistently produced a finished water with turbidity lower than 2 FTU at a filtration rate of 12.2 m/hr. Filter runs were 6 to 8 hours in length. By switching to a cationic polymer, finished water turbidities decreased to less than one unit and filter run lengths were extended to 16 hours or more.

Other investigators have also commented on the limitations of direct filtration applications in terms of required raw water quality (16, 33, 59, 94, 99).

Coagulants

The coagulants used most often in direct filtration applications are alum, cationic polymers, alum plus cationic polymers, and alum plus nonionic polymers. Iron salts are seldom used probably due to increased handling and storage difficulties, but may have other advantages (33),

Alum Pilot plant and plant scale studies have both shown that alum can be used successfully to produce a finished water turbidity of 1 NTU or less in direct filtration systems (9, 14, 18, 32, 33, 43, 44, 83, 84, 93). Required alum dosages depend upon the quality of the raw water but generally range between 2 to 20 mg/1. Excessive alum dosages can enhance the probability of turbidity breakthrough and thereby decrease filter run length. Also, excess alum can cause increased rates of headloss in the filters. Another problem associated with use of alum is the possibility that excessive aluminum residuals may appear in the finished water (43, 44, 64). This problem may be controlled by proper pH control and/or adjustment in flocculation periods preceeding the filters. One investigator reported on the use of ferric chloride instead of alum (33). Ferric chloride was initially tried because of problems experienced with aluminum residuals using alum. It was concluded that ferric chloride produced a stronger floc that would result in a higher net water production.

Breland and Cleasby (14) used 5-10 mg/1 of alum in a contact flocculation filtration study with Iowa river waters. The raw water turbidities during this study ranged from 50 to 350 JTU. With dual media

filters operated at a filtration rate of 4.9 m/hr more than 95 percent turbidity removal was achieved. Except during periods of high algae populations, runs were always terminated due to solids breakthrough.

Cationic polymers

Cationic polymers have been shown to be effective for direct filtration applications either as sole coagulants or in combination with alum. Adin and Rebhun (2) concluded that the use of polyelectrolytes enabled the useage of a larger filter media size which decreased the rate of headloss buildup while maintaining an acceptable effluent quality.

Yeh and Ghosh (101) studied the effect of polymer molecular weight in direct filtration applications. They found that usually linear homopolymers with a molecular weight between 10,000 to 100,000 provided the best results. For their operating conditions, optimal filtration was achieved at a polymer dosage that corresponded to the production of the largest number of particles with a diameter of 20 μm. Stump and Novak (88) concluded that cationic polymers used for direct filtration should have a molecular weight between 10,000 and 200,000. High molecular weight polymers provide good particle removal but the rate of filter headloss is excessive and results in short filter runs.

Foley (33) reported on the use of a cationic polymer as the sole coagulant to replace alum. The cationic polymer produced acceptable results but after two years of use it was discontinued due to development of mudballs in the filter media even when backwashed with the aid of surface agitators.

When using alum and cationic polymer in combination, both coagulants contribute to particle destabilization. The use of alum generally tends to decrease the filter ripening period compared to polyelectrolytes. The polyelectrolytes provide added floc strength which results in increased resistance to shear forces and longer filter runs. Tredgett (93) concluded that during cold weather periods polymers in addition to alum were beneficial in maintaining an acceptable finished water quality. Only alum was normally used. The effect of coagulant type on the duration of the ripening period can be seen in the work of Cleasby and Saleh (18), Adin and Rebhun (2) and Shea et al. (83).

Shea et al. (83) conducted contact flocculation studies using alum and Cat Floc, a cationic polymer. They found that higher polymer dosages were needed as the size of the filter media increases. Comparing runs made with polymer to those made with alum showed higher terminal headlosses with polymer useage. In a filter run initially using alum and polymer, the alum feed was discontinued after five hours. When the alum feed was discontinued the effluent quality decreased rapidly.

Tate et al. (89) reported successful results using 3-6 mg/l of alum and 0.12-0.25 mg/l of a cationic polymer. The raw water turbidity was 2 turbidity units or less while final turbidities were generally less than 0.1 turbidity unit. Tate and Trussell (90) have reported on other experimental investigations for direct filtration of low turbidity raw water using alum and cationic polymer as the coagulants.

Nonionic polymers In direct filtration systems, nonionic polymers can be used in combination with alum. The use of solely nonionic polymers to destabilize particles is not common but under certain conditions it is possible (53). In direct filtration systems the alum is used as the destabilizing agent and the nonionic polymer serves to strengthen the floc and also serves as a bridge between the destabilized particles and the filter medium.

In a comparison of direct filtration and complete conventional treatment, Stephenson (86) used a combination of 10-20 mg/1 of alum and 0.075 mg/1 of a nonionic polymer. The combination of alum and polymer resulted in a high proportion of solids being removed in the coarse coal media in a dual media filter. The results proved equal to or better than those obtained with complete conventional treatment.

Dharmarajah (29) made direct filtration runs using the same apparatus as Stephenson (86). He used a different nonionic polymer in combination with alum to study color removal in direct filtration. At alum dosages of 15 mg/1 and polymer dosages of 1.5 mg/1, direct filtration without flocculation was most effective in producing a quality finished water. Without flocculation, he experienced high rates of head loss development but a greatly increased time to solids breakthrough.

Hutchison and Foley (44) found nonionic polymers superior to activated silica in combination with alum in preventing turbidity breakthrough. Polymer dosages ranged from 25 to 200 ppb depending on the operating conditions such as filter media size and filtration rate. They also

concluded that the polymer should be added just above the filter media for optimum results.

Geise et al. (34) reported on the use of a nonionic polymer added as a filter conditioner in a complete conventional treatment plant. The addition of 0.01 to 0.025 mg/l of the polymer increased water quality in general, and in particular, when flow increases during the backwash of a filter normally decreased water quality.

Tate et al. (89) reported on the use of a nonionic polymer as a filter aid following particle destabilization with alum and a cationic polymer. The results were inconclusive since only limited testing was performed.

In a study of asbestos removal in Lake Superior water by direct filtration, Peterson et al. (74) reported that asbestiform fibers are the last particles to be destabilized and removed by aluminum hydroxide floc formation. A nonionic polymer used as a coagulant and filter aid strengthens the aluminum hydroxide floc and decreases the filter ripening period. Their pilot plant runs indicated that excessive polymer dosages increased the rate of filter headloss while underdosing the polymer resulted in early turbidity breakthrough.

Pretreatment

Pretreatment requirements for direct filtration systems include rapid mixing and can include flocculation facilities. The main purpose of the rapid mix is the instantaneous dispersion of the coagulants into the fluid. Traditionally, rapid mix periods have been on the order of 30-60

seconds. In adsorption destabilization with alum, the reactions are extremely fast and occur within microseconds without formation of hydrolysis polymers and within one second if polymers are formed. Destabilization by sweep coagulation occurs in the range of 1-7 seconds (3). Amirthrajah and Mills (3) conducted an experimental study to determine the optimum type of rapid mix unit for alum destabilization. They concluded that if destabilization occurs by adsorption a high intensity short duration rapid mix produces the best results. Where sweep coagulation dominates there is little difference between high speed short duration and lower speeds for longer durations. Velocity gradients tested were 300 sec^{-1}, 1000 sec^{-1} and 16,000 sec^{-1} for 60, 20 and 1 second respectively.

Letterman et al. (54) found the existence of an optimum rapid mix time for each value of velocity gradient and alum dosage tested, as measured by residual turbidity. This suggests that the early conditions of floc formation are important and the function of rapid mixing is more complex than simple dispersion of the coagulant.

Yeh and Ghosh (101) determined that cationic polymers were most effective in direct filtration when rapid mixing provided a velocity gradient in the range of 300 to 600 sec^{-1} for a period of 3 to 8 minutes. Stump and Novak (88) found that for a rapid mix period of 2 minutes velocity gradients for cationic polymers with a molecular weight less than 50,000 should be about 300 sec^{-1} and for cationic polymers with a molecular weight > 100,000 the velocity gradient should be between 600-1000 sec^{-1}.

Morrow and Rausch (65) studied the effect of velocity gradient and mixing time on coagulation and flocculation when cationic polymers were the primary coagulants in complete conventional treatment. They concluded that cationic polymers could replace inorganic coagulants completely for both low and high turbidity surface waters when velocity gradients greater than 400 sec^{-1} are applied. The optimum coagulation occurred in less than 2 minutes and an increase in the velocity gradient decreased the required retention time and polymer dose necessary for destabilization. They also found, in some instances, that a nonionic polymer filter aid improved filter performance.

There appears to be no well established criteria for the need or duration of mixing in flocculation facilities preceeding direct filtration. Stephenson (86) and Dharmarajah (29) each studied the effect of flocculation time and found that no flocculation before filtration resulted in higher terminal headloss but increased filter run lengths due to elimination of solids breakthrough. In both investigations the finished water turbidity was equivalent from the filters receiving both flocculated and nonflocculated influent.

Yeh and Ghosh (101) concluded that flocculation may not be necessary for most direct filtration operations, especially if the suspended solids concentration is 30 mg/l or greater. O'Melia (66) found that flocculation was not essential in coarse media direct filtration of high turbidity wastewater effluents. In direct filtration studies of Iowa River water, Cleasby and Saleh (18) produced acceptable filtered water turbidity by adding the coagulants into the influent lines to the filters.

Hutchison (43) and Hutchison and Foley (44) conducted direct filtration studies on several Great Lakes waters. These investigators found when using alum that increasing flocculation time above 10 minutes increased the chances of early breakthrough. A velocity gradient greater than 20 sec^{-1} also increased the probability of early breakthrough. Hutchison and Foley (44) speculated that the floc strength was reduced by longer flocculation times. The floc strength reduction was the result of constant breaking down and rebuilding of the floc particles. The reduced strength of floc was more susceptible to shear in the latter stages of a filter run. The growth and breakage of alum floc was studied by Boadway (13). He concluded that as the floc particles increase in size, their surface geometry becomes more complex leading to a progressively poorer fit during formation. The poorer fit leads to a lower floc density.

Letterman et al. (54) studied direct filtration using cationic polymers with a pretreatment velocity gradient ranging from 0 to 700 sec^{-1} for periods varying from 2 to 10 minutes. They concluded that within a normal range of polymer concentrations the effluent turbidity was independent of the pretreatment and filter operating conditions. Trussell et al. (95) reported on several studies with and without flocculation preceding direct filtration. No general trend was established. In one case particle counts were lower in the finished water when flocculation was used, however the turbidity was not much different than when no flocculation was employed. In another case, flocculation improved

the effluent turbidity but did not improve the effluent particle count and resulted in early turbidity breakthrough.

Tate et al. (89) found no difference between 13 and 26 minutes of flocculation time on treated water turbidity or particle counts. They used alum in combination with a cationic polymer for direct filtration of low turbidity raw water. When using alum and a cationic polymer, Treweek (94) found that increasing the flocculation time beyond 7 minutes resulted in the formation of a larger floc. The increased energy expenditure did not improve the quality of the filter effluent.

Monscvitz et al. (64) studied the effect of flocculation time on filtered water quality using alum and a cationic polymer with a low water turbidity. They found that the minimum coagulant use was obtained by 20 to 30 minutes of flocculation. These findings were based on zeta potential adjustment and process performance as measured by effluent turbidity, filter run length and filter headloss. Without flocculation alum carry over was a problem. Stump and Novak (88) concluded that when using cationic polymers flocculation times of 20 minutes or more are necessary.

Rate control

There are two basic methods of filter operation that differ mainly in how the rate of flow through the filter is controlled. These two methods are constant rate filtration using effluent and using influent rate control and variable declining rate filtration (17).

Constant rate filtration provides the same filtration rate (within

minimum limits) throughout the duration of the filter run. Constant rate is maintained by a flow control valve on the effluent line from the filter or by providing equal flow to all filters and having a deep enough depth of filter box to provide an increased water level above the filter media. Because a constant rate is maintained, filtered water quality may decrease during the latter stages of a filter run.

Declining rate filtration splits the flow to the filters in accordance with the length of time the filters have been in service. The dirtiest filter receives the least flow while the cleanest filter receives the most. The flowrate in a filter decreases as the filter run progresses after each of the other filters is backwashed and returned to service. This not only aids in maintaining a good quality effluent from that filter but when the filter is taken out of service for backwashing there is less of a demand put on the other operating filters. Filters containing media with a low total surface area may be more subject to turbidity breakthrough with changing flowrates. Trussell et al. (95) found that in uniform media filters the use of angular media was superior to rounded or smooth media. Performance was based on limiting headloss, time to turbidity breakthrough, average effluent turbidity and average concentration of particles in the effluent. They attributed the shape of the media to its effect on the bulk porosity of the filter bed which affects the headloss resulting from deposits in the filter.

Hutchison and Foley (44) concluded that declining rate filtration produced less shear force on the floc particles deposited within the filter bed when compared to constant rate filtration. They speculated

that as the rate declined floc deposition would take place in areas within the filter that ordinarily would not be occupied during constant rate filtration. The net result would be longer filter runs due to increased particle removal and less chance of run termination due to turbidity breakthrough.

Cleasby (16) points out that in a filter which contains previously deposited solids an increase in the filtration rate can be detrimental to effluent quality. The extent of the disturbance in the filter will depend upon the type of solids deposited in the filter and the magnitude and suddenness of the rate change. As a result all sources of sudden rate change should be avoided in design and operation.

DiBernardo and Cleasby (30) conducted pilot filtration experiments with about the same apparatus used in this study. They compared constant rate to declining rate filtration. The conclusions of the study indicated that declining rate filtration was superior to constant rate filtration in terms of the following:

1) The declining rate filter system produced average effluent turbidities consistently and substantially lower than the constant rate filter system.
2) When operated to the same terminal headloss, the constant rate filter run length was substantially shorter than the run length obtained with declining rate filtration.
3) Less effect on effluent water quality when a filter was taken out of service for backwashing was observed when the filters were operated in the declining rate mode.

4) When operating in the constant rate mode, the filter effluent quality gradually deteriorated during the filter run and eventually turbidity breakthrough occurred. This was not evident when the declining rate mode was used.

The AWWA committee report on direct filtration reported that the majority of direct filtration plants operated at a constant rate rather than in the declining rate mode. Limited pilot plant test results showed no advantage of declining rate filtration over constant rate filtration but the committee speculated that declining rate filtration may be advantageous under the right conditions (20).

Literature Directly Related to This Study

Two other studies carried out at Iowa State University involved the evaluation of contact flocculation for direct filtration of turbid waters. The first was conducted by Breland and Cleasby (14) and involved both laboratory and field investigations using alum alone as the coagulant. The pilot plant apparatus consisted of four filters of various media depths. Three, six and nine inches of anthracite and 9 inches of anthracite over 15 inches of sand were used in the laboratory studies. The sand media had an effective size of 0.77 mm. In the field studies, dual media filtration was evaluated. The anthracite media had an effective size of 0.98 mm. Breland and Cleasby determined that 5 mg/l of alum was the optimum chemical dosage for removal of kaolin, the same type of kaolin as used in this study, by filtration. The kaolin concentration

used by Breland and Cleasby was 20 mg/l. Alum dosages between 0 and 20 mg/l were evaluated in the pilot filter runs. Alum dosages above the optimum resulted in a shorter initial improvement period but produced essentially the same effluent quality. With the higher alum dosages earlier turbidity breakthrough occurred and higher rates of headloss developed compared to when the optimum alum dosage was used. (In this study, the optimum alum dose was 7.5 mg/l for an equivalent kaolin concentration.)

In the field evaluation, the optimum alum dose for direct filtration was determined to be 10 mg/l. Turbidity breakthrough was defined as an abrupt increase in the effluent turbidity from the 24 inch dual media filter and was determined by an increase in effluent turbidity of 1 JTU over the minimum previously observed effluent turbidity value. When algae counts were high in the raw water, poorer effluent turbidities and higher rates of headloss in the filters resulted, compared to when the turbidity consisted mostly of silts and clays. When the influent turbidity was between 50 and 80 JTU and was due primarily to the presence of clay and silt, the filtered water turbidity was half as high as when the same level of influent turbidity was due primarily to high algae populations. During high algae blooms, however, the time to turbidity breakthrough in the 24 inch filter was almost 40 hours compared to 10 hours when the equivalent turbidity was caused by clay and silt. Terminal headloss in the 24-inch filter was about 10 feet during algae blooms and about 1.5 - 2.0 feet when the equivalent influent turbidity was caused by clays and silts. If one was to consider Breland's data in terms of producing a

final effluent turbidity of 1 JTU or less for a minimum run length of 10 hours, the maximum influent turbidity that could be accommodated would be about 80 JTU when using 10 mg/l of alum at a flowrate of 2 gpm/ft^2 through the 24 inch dual media filter. Filtration rates of 4 and 6 gpm/ft^2 were evaluated in the field investigation. The use of the higher rates resulted in severe reductions in water production due to turbidity breakthrough shortening the filter run lengths.

Additionally, two important observations were made in Breland and Cleasby's study. It was determined that the time between chemical addition, namely alum, and entrance to the filter had essentially no bearing on the filtered water results when this time period was varied between 6.5 seconds and 3.4 minutes. Secondly, for the filter runs during the field investigation, all runs except those during algae blooms were terminated due to turbidity breakthrough when 10 mg/l of alum was used with a filtration rate of 2 gpm/ft^2.

The results obtained in the unpublished laboratory and field studies by Breland and Cleasby led Baumann[1] to conclude that use of direct filtration of high turbidity waters was technically feasible if the solids holding capacity of the filter could be increased. He suggested (recommended) that a series filtration system be evaluated using a coarse media filter for increasing contact-flocculation and solids holding capacity ahead of a standard dual media filter. The Breland study indicated that small alum dosages would permit production of low turbidity

[1]Personal communication, Dr. E. R. Baumann, Department of Civil Engineering, Iowa State University, Ames, Iowa.

filtered water. The system evaluated by Breland was impractical, however. since the solids holding capacity at reasonable filtration rates (5 gpm/ft^2) was so low.

Cleasby and Saleh (18) conducted pilot filtration runs using a coarse media-multi media series filtration system similar to that used in this study. Both systems were designed based on recommendations of Dr. E. R. Baumann (personal communication, Dept. of Civil Engineering, Iowa State University). The coarse media filter contained 1.37 m of coarse sand with an effective size of 2.17 mm and a uniformity coefficient of 1.28. When operated as the Culligan USA system, recommended by Baumann, the coarse media filter was followed in series by a filter containing a multi media arrangement. The multi media filter contained 46 cm of anthracite coal with an effective size of 23 cm of calcined aluminum silicate and 7.6 cm of garnet with an effective size of 0.37 mm and a u.c. of 1.56. When operated as a modified Culligan USA system, the coarse media filter was followed in series by a dual media filter. The dual media filter contained 53.3 cm of anthracite coal with an effective size of 0.94 mm and a u.c. of 1.43, and 30.5 cm sand with an effective size of 0.49 mm and a u.c. of 1.37.

Although Cleasby and Saleh (18) evaluated the filter system on various waters, only the portion of their study relating to direct filtration of Iowa River water at Iowa City will be discussed herein. Raw water was pumped directly from the Iowa River to the filter system. Treatment chemicals were injected into the influent line where the only flocculation that occurred prior to filtration took place. The treatment chemicals used were either alum or a cationic polymer. Headloss

measurements were made across each filter in the series but only final effluent turbidity was measured. Turbidity breakthrough was generally defined by a filter water turbidity greater than 1 TU when the final turbidity during a run was less than 1 TU. When a final turbidity of less than 1 TU was not obtained, breakthrough was defined as a sudden and sustained increase in the filter water turbidity. The filter run conditions and results are summarized in Table 2.

At a filtration rate of 3.8 gpm/ft^2 using a cationic polymer in the pretreatment, run lengths in excess of 10 hours were possible while maintaining an average effluent turbidity of less than 1.0 TU. No attempts were made to measure the coarse media filter effluent turbidity or to backwash this filter separately from the multimedia filter. In the filter runs made at a rate of 3.8 gpm/ft^2, the largest headloss development in the coarse media filter was 3.4 m after a 21 hour filter run. This filter run with an average influent turbidity of 73 NTU, was terminated due to final turbidity breakthrough rather than due to excessive headloss development.

Increasing the filtration rate to 6.3 gpm/ft^2 and using a cationic polymer in the pretreatment resulted in poorer effluent quality and a higher rate of headloss development, especially in the multimedia filter, Table 2. Net water production per unit of headloss development was lower at a filtration rate of 6.3 gpm/ft^2 (avg. = 550 gal/ft^2/ft) than at a filtration rate of 3.8 gpm/ft^2 (avg. = 1100-1500 gal/ft^2/ft). In Runs 17-1 and 18-1, where only alum was used as the coagulant, the run lengths were reduced by almost one half compared to when a cationic

Table 2. Summary of field evaluation of Culligan filters for water treatment (18)

Run #	Rate gpm/ft^2	Alum Dose mg/l	Polymer Dose mg/l	Influent Turbidity TU ± (S)[a]	Effluent Turbidity TU ± (S)[a]	Headloss Increase Coarse Media	Headloss Increase Media ft.	Run Length hrs.[b]
8-1	3.8		10	63(1.3)	0.98(0.49)	1.5	0.6	10
9b-1	3.8		15	53(5.0)	0.36(0.19)	2.9	4.6	35
10-1	3.8		15	109(99)	0.53(0.35)	1.1	0.6	22
11-1	3.8		15	123(51)	0.62(0.45)	1.7	3.1	19
12-1	3.8		15	73(10)	0.71(0.23)	3.4	4.7	21
13-1	6.3		15	61(3.0)	2.85(0.46)	1.7	2.2	9.5
15-1	6.3		30	62(4.1)	1.03(0.74)	4.2	8.3	13.5
16-1	6.3		15	62(3.8)	1.64(0.75)	2.1	8.3	11.5
17-1	6.3	20		53(1.7)	1.52(1.07)	0.2	4.6	5
18-1	6.3	10		49(1.7)	1.18(0.46)	0.2	5.5	7
19-1	6.3		30	39(4.3)	0.60(0.29)	4.9	11.8	26
20-1	6.3		20	29(3.8)	0.94(0.29)	4.4	13.6	24
21(0-25 hr)	6.3		15-25	24(3.3)	1.14(0.53)[c] 1.53(0.71)[d]	5.3[c] 3.2[d]	8.8[c] 3.7[d]	35
21(25 hr-end)	6.3			21(5.4)	0.67(0.05) 0.79(0.08)	9.5 4.6	20.6 8.8	
22	814		25	17(2.8)	1.57(0.48)	5.1 3.9	10.2 5.5	20

[a](S) = standard deviation.

[b]Both filters backwashed together.

[c]USA system.

[d]Modified USA system.

polymer was used. The headloss data indicated that most of the removal of solids occurred in the multi media filter and that filtered water turbidities averaged greater than 1 TU.

In Runs 21 and 22, the performance of the Culligan USA system was compared to that of a modified USA system. The average effluent turbidity from the Culligan USA system using a four media filter was lower than that from the modified USA system using a dual media filter. The headloss development in the Culligan USA system was significantly higher than in the modified USA system. Both the turbidity and headloss data may have been influenced by the fact that the dual media in the modified USA system was clean media and placed in the filters just prior to Runs 21 and 22. The four media in the Culligan USA system had been used in previous runs. Also, a lower headloss was expected in the modified USA system since the coal used in the second filter was coarser than the coal used in the second filter of the Culligan USA system.

Cleasby and Saleh made several conclusions and observations from the results of their studies which are relevant to the basic design and operation of direct filtration system employing the use of coarse media filters in series with dual or multi media filters. The most important observations and conclusions are as follows:

1. In the early part of a filter run, the depth clarifier or the coarse media filter produces a reasonably low effluent turbidity but as the run progressed the filter gets loaded with solids and more solids begin to pass on to the second filter. Near the end of a filter run practically all the solids in the form of large floccules, pass through

the depth clarifier.

2. The second filter in the series should have a top media large enough to prevent surface cake formation. The extra layer of plastic chips in the Culligan four-layer filter helped accommodate the heavy load of flocculated solids that pass through the coarse media filter or depth clarifier as the filter run progresses.

3. When only cationic polymer was used for chemical pretreatment, long ripening periods were observed. The prolonged period of poorer effluent quality was considered a principal weakness of using a single coagulant. The use of alum plus polymer was recommended.

4. When using cationic polymer alone for pretreatment, the final effluent turbidity could not be maintained below 1 TU at a filtration rate of 6.3 gpm/ ft^2 over a practical run length. Therefore, they recommended that when using the Culligan cationic polymer alone, that a maximum filtration rate of 5 gpm/ft^2 be used to "achieve consistently a filtrate turbidity of less than 1 TU with widely varying raw water quality."

The first and second observations relate both to the design and operation of a filtration system using the depth clarifier or coarse media filter concept. In the Culligan study, Cleasby and Saleh operated both the coarse media and multimedia filters for the same run length. When the final effluent turbidity exceeded the turbidity goal, the filter run was terminated and the filters backwashed. This type of operation places a heavy solids removal burden on the second filter near the end of the filter run, especially when the raw water turbidity is high

($\approx \geq$ 50 NTU). An alternative to this type of operation would be to backwash the coarse media filter alone when the turbidity level in its effluent exceeds a value of say 1/3 of the raw water turbidity. The result of this operational mode would be to reduce the run length of the coarse media filter and to increase the run length of the dual or multimedia filter. The chances of early solids breakthrough in the final effluent would be lessened by this type of operation.

Depending on the operational mode, the size of the top layers of media in the second filter is an important design consideration. The media must be large enough to allow solids penetration into the filter and thus avoid excessive headloss development near the filter surface. When both filters in the system are operated for the same run duration, it is essential to have a large media at the surface of the dual or multimedia filter to prevent excessive headloss in the latter stage of the run when the second filter receives a greater solids load of flocculated particles. When the filters are backwashed independently, the need for a large surface media is reduced but not avoided. Since the purpose of the entire system is to directly filter high turbidity waters, it is likely that the terminal turbidity of the coarse media filter will be at or above the turbidity levels commonly considered as the maximum levels for standard direct filtration systems ($\approx$ 30 NTU). The importance of the plastic top media in Culligans four-media filter was shown in Cleasby and Saleh's Runs 21 and 22. Although the Culligan USA system had a thin layer of plastic media at the surface, the coal size in the four-media filter was much smaller than the coal media in the dual media

filter in the modified USA system. Even with the coarse plastic chips on top, the four-media filter developed more headloss than the dual media filter in the modified USA system.

The third and fourth observations relate to chemical pretreatment. All the filter runs made by Cleasby and Saleh were conducted without conventional flocculation and with either alum or cationic polymer used as the coagulant. The filter ripening period should be reduced by using a combination of both alum and polymer. The scope of the Culligan study was limited to the use of only one cationic polymer. It is possible that other polymers could have produced more desirable results. It is also possible that filtration performance could have been enhanced by the use of conventional flocculation facilities prior to filtration.

The results of the direct filtration studies conducted by Breland and Cleasby (14) and Cleasby and Saleh (18) had a direct influence on the design of the study reported herein. The major areas of influence were:

1) Breland and Cleasby (14) found that a low alum dosage (10 mg/l) was required in direct filtration of waters representing a broad range of high influent turbidities. Their filter runs were short due to early solids breakthrough at low terminal headloss. This study suggests that using a coarser media filter ahead of and in series with a dual media filter to reduce the influent solids concentration to the dual media filter could produce acceptable filtrate in the direct filtration of high turbidity waters using a relatively narrow range of alum dosage in the chemical pretreatment.

2) Cleasby and Saleh (18) used a coarse media filter in series with a four-media or dual media filter. They found alum alone ineffective in producing satisfactory filtration results both in terms of final effluent turbidity and filter run length. The use of a cationic polymer as the pretreatment chemical resulted in the production of a filtered water with a turbidity below 1 TU and in reasonable filter run lengths (> 10 hours). Cleasby and Saleh's filter runs were conducted using a constant filtration rate and both the coarse media filters and four media or dual media filters were backwashed when the final effluent turbidity exceeded the desired turbidity goal or terminal headloss. As a result of Cleasby and Saleh's study, the study reported on herein included:

a) The effect of using both alum and polymer in the pretreatment system prior to series filtration of high turbidity water.

b) The use of a declining rate operational mode in some of the filter runs to compare the filtration system performance to filter runs using a constant rate operational mode.

c) Backwashing the coarse media filters independently of the dual media filters by establishing separate coarse media effluent turbidity goals and limiting terminal headloss criteria.

MATERIALS AND EQUIPMENT

Raw Water Materials

Introduction

The evaluation of direct filtration of high turbidity waters was carried out in three phases. The first phase involved the use of the jar test to select an optimum chemical coagulant combination based on settled water turbidity. Alum alone and alum in combination with a cationic or one of several different nonionic polymers were evaluated. The raw water suspension was composed of either kaolin or calcium aluminosilicate added to a synthetic water. The particle concentration in the raw water was either 10 or 100 mg/l.

The second phase of the study involved laboratory pilot filter experiments. Two coarse media filters were operated in series with two standard dual media filters. The raw water supply in the filtration experiments consisted of either kaolin or calcium aluminosilicate added to university tap water. The range of kaolin concentrations used varied from 25 to 300 mg/l while calcium aluminosilicate was evaluated only at 100 mg/l. Coagulants used during the second phase of the study included alum and alum in combination with a cationic polymer or one of several nonionic polymers. The performance of the filtration system was based on the total headloss in each set of filters, the coarse media effluent combined turbidity, the dual media effluent combined turbidity, and the coarse media filter run length. Both constant rate and declining rate operation were evaluated.

The third phase of the study involved a field investigation with the pilot filtration system used during the second phase. Des Moines river water was used as the raw source. The coagulants used during the third phase included alum in combination with a nonionic polymer or one of two cationic polymers. Filter performance was evaluated on the same basis as used in the second phase.

Kaolin

The kaolin used in this study was Old Hickory No. 5 Ball Clay manufactured by Old Hickory Company of Paducah, Kentucky. Kaolin belongs to the kaolinite group of clay minerals. The clays in this group are nonexpandable in water and have a low cation exchange capacity of 1 to 10 meq/100 g (97). The particular clay used in this study has a cation exchange capacity of 10 meq/100 gm. A typical analysis of the Old Hickory No. 5 Ball Clay is presented in Table 3. The zeta potential of the clay in tap water and synthetic water at the pH value (7.0) used in this study was approximately -20 mV.

Aluminosilicate

The calcium aluminosilicate particles used in this study were supplied by the Procter & Gamble Company of Cincinnati, Ohio. These particles are identified as P & G identification number IPS-76.001. The analytical characteristics of calcium aluminosilicate are presented in Table 4. The zeta potential of calcium aluminosilicate in tap water at pH 7.0 was approximately -10 mV and in synthetic water at pH 7.0 the zeta potential was approximately zero.

Synthetic water

The synthetic water used in the jar test portion of this study was designed to represent a typical surface water containing hardness and alkalinity. The synthetic water was prepared by adding the following salt concentrations to demineralized water:

210 mg/l $NaHCO_3$

83.25 mg/l $CaCL_2$

30.0 mg/l $MgSO_4$

Table 3. Kaolinite clay particle characterization[a]

Typical Chemical Analysis	
Silicon dioxide	55.86%
Aluminum oxide	30.38%
Iron oxide	0.99%
Titanium dioxide	1.40%
Calcium oxide	0.11%
Magnesium oxide	0.05%
Sodium oxide	0.08%
Potassium oxide	0.17%
Ignition loss	10.57%
Typical Particle Size Analysis	
% minus 20 microns	98.1
% minus 10 microns	93.8
% minus 5 microns	85.0
% minus 2 microns	67.4
% minus 1 microns	57.5
% minus 0.5 microns	47.4
Raw Properties	
Crude color	white
pH (28%)	5.7
Filtration rate (ml)	26
Soluble sulfate (ppm)	114

[a]The above data were provided by the Old Hickory Clay Company, Paducah, Kentucky.

Table 4. Chemical characterization of calcium aluminosilicate, ERI-ASL[a] results

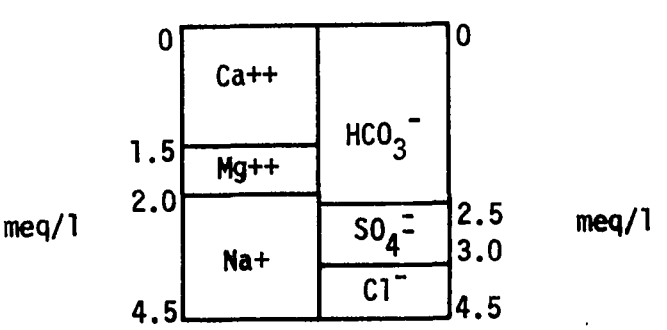
Parameter	Ca-form
% Moisture	21.8 ± 0.7[b]
% Na	1.88 ± 0.03
% Al	13.4 ± 0.3
Ca^{++} exchange cap. (mg $CaCO_3$/g slumino-silicate)	17 ± 4
% PDAS-SiO_2[c]	31.5 ± 2.1
% Total SiO_2	32.5 ± 1.0
% Ca	9.51 ± 0.12

[a]Engineering Research Institute, Analytical Services Laboratory.

[b]Average values and standard deviations based on 9 separate determinations.

[c]Values based on filtration of a 500 ml sample.

The bar graph below shows the ionic characteristics of the synthetic water.



The pH of the synthetic water was approximately 8.5. The turbidity of the synthetic water prior to addition of suspension particles varied with the quality of the demineralized water but was generally less than 0.5 NTU.

Tap water

University tap water was used to prepare the raw water suspensions during the laboratory pilot filtration experiments. The university tap water is supplied by the City of Ames. The water is obtained from wells and is treated by lime softening, filtration, fluoridation, chlorination and stabilization with metaphosphate. The pH of the university tap water is about 8.5. Total hardness varies depending on which wells are used but is generally between 160 to 170 mg/l. Total alkalinity is about 45 mg/l. The turbidity of the tap water was normally in the range of 0.6 - 0.7 NTU.

Chemicals

The chemicals used during this study include those used to prepare the synthetic water, the acid used for pH adjustment and the coagulants used for treatment of the water. The chemicals used, except for the organic polymers, are summarized in Table 5 .

Several polymers were used in the course of this study. These polymers are described in Table 6 .

Table 5. Summary of chemicals used in this study

Chemical	Formula	Formula Weight	Grade	Purpose
Calcium Chloride	$CaCl_2 \cdot 6H_2O$	219.08	Certified	Synthetic Water
Magnesium Sulfate	$MgSO_4$	120.39	Reagent	Synthetic Water
Sodium Bicarbonate	$NaHCO_3$	84.01	Reagent	Synthetic Water
Hydrochloric Acid	HCL	36.46	Reagent	pH Adjustment
Aluminum Sulfate	$Al_2(SO_4)_3 \cdot 18H_2O$	666.42	Technical	Coagulant

Miscellaneous Equipment

Jar test

The six paddle stirrer used for the jar tests was a standard Phipps and Bird Custom 270 manufactured by Phipps and Bird of Richmond, Virginia. The mixing speed ranged from 0 to 270 rpm. The jars used were 2000-ml beakers.

Turbidimeter

The turbidimeter used in this study was a Hach Turbidimeter, Model 2100A supplied by Hach Chemical Company of Loveland, Colorado. The turbidity standards used for calibration were rated in Nephelometric Turbidity Units (NTU).

Electrophoresis apparatus

The Laser Zee Meter, Model 400, manufactured by the Pen Kem Company, Croton-On-Hudson, New York, was used during this study to measure particle zeta potential. The use of this instrument allows for several zeta potential readings to be taken within one minute. Instead of measuring the zeta potential of individual particles this instrument uses a particle cloud method which encompasses the field of particles seen through the dark field microscope. After filling the electrophoresis cell with sample fluid, the zeta potential measurement is made by first applying a voltage across the cell and then adjusting the speed of rotation of a prism. When the particles in the particle cloud appear stationary as the prism is rotated the zeta potential is read directly. The zeta potential value displayed must be corrected for temperature unless the sample fluid temperature is equal to 20° C.

Pilot Filter System

The pilot filter system used in this study was composed of raw water storage tanks, a rapid mix unit, chemical feed pumps, plexiglas filter columns and appropriate appurtenances and a rate controller. A schematic diagram of the pilot plant apparatus is shown in Figure 5.

Raw water tank

Two 2,700-liter steel tanks were used to store the raw water supply. One tank was used as a feed tank as well as for storage while the other tank was used strictly for storage. Mounted on each tank was a variable

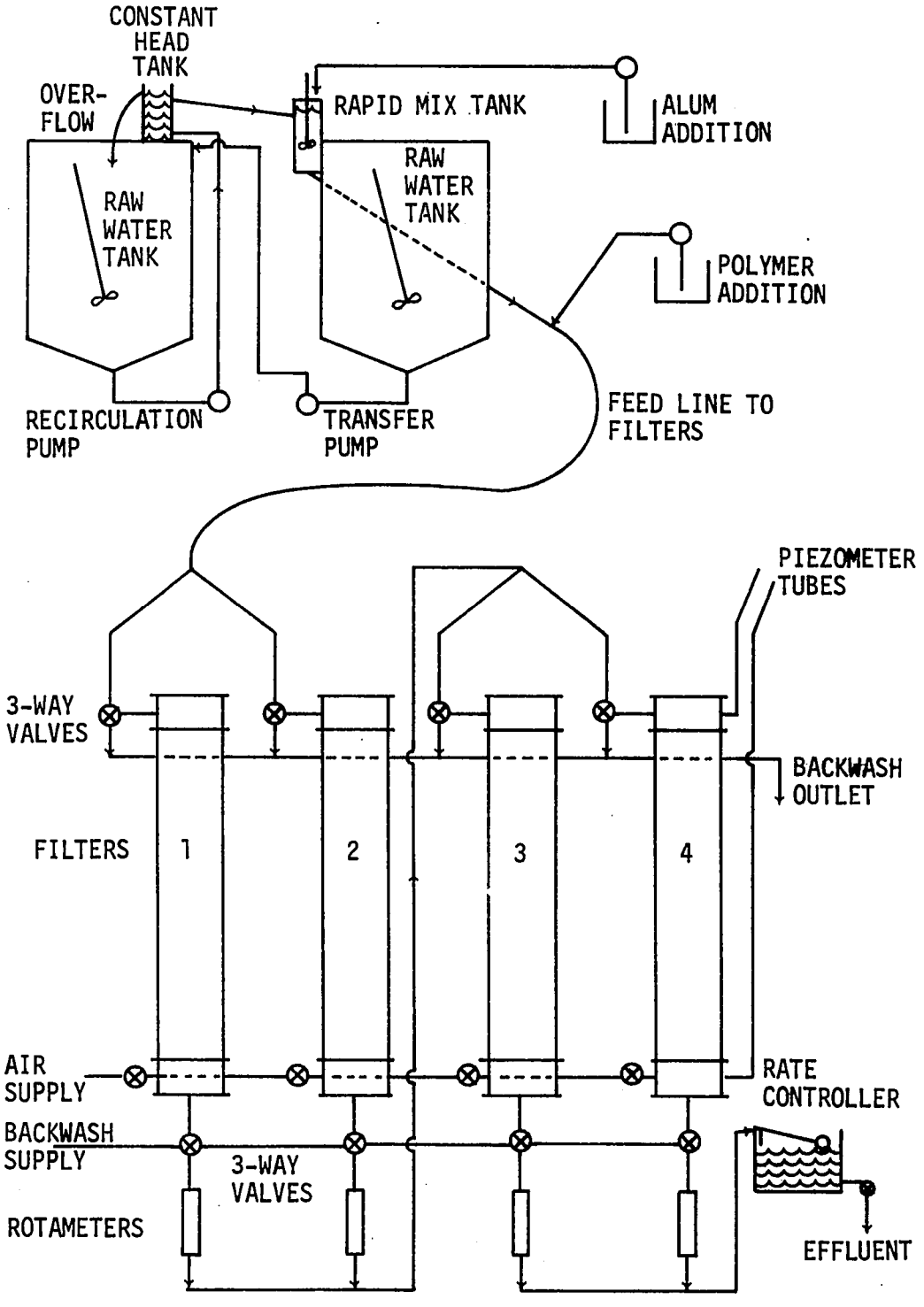


Figure 5. Schematic diagram of the pilot filtration system

Table 6. Polymers used in this study

Polymer Name	Manufacturer	Ionic Character and Type	Molecular Weight	Physical Form
Cat Floc T	Calgon Corporation	High cationic polydiallydimethyl ammonium	30,000-40,000	Clear liquid (20% active)
Dow NP-10	Dow Chemical Company	Nonionic polyacrylamide	-	White, granular solid
Percol 720	Allied Colloids, Inc.	Nonionic polyacrylamide	-	White, granular solid
Percol LT-20	Allied Colloids, Inc.	Nonionic polyacrylamide	-	White, granular solid
Percol LT-22	Allied Colloids, Inc.	Medium cationic polyacrylamide	-	White, granular solid

speed propeller mixer. Both mixers were rated at 1/2 h.p.

The purpose of the propeller mixers was to keep the raw water particles in suspension. A plexiglas constant head tank was located above the feed tank. A recirculation pump continuously pumped the raw water to the constant head tank. Excess water drained back into the raw water feed tank.

Rapid mix unit

The raw water flowed by gravity to the rapid mix tank. This tank was constructed of plexiglas. The rapid mix tank provided 30 seconds of detention time at a filtration rate of 12.2 m/hr. A variable speed propeller type stirrer was used to mix the contents of the rapid mix unit. Throughout the study the stirrer speed was kept at approximately 300 rpm. Alum was fed into the rapid mix unit at the same height as the propeller blades.

Chemical feed pumps

Two small positive displacement pumps were used to feed chemicals at desired points in the filtration system. One pump was used for alum and the other for polymer. Both pumps could be adjusted, but were maintained at a 2 liter/hr flowrate throughout the study.

Filter columns

The pilot filter units used in the laboratory and field investigation studies consisted of four filters. The filters were constructed of clear plexiglas. Each filter was 185.42 cm high and had an inside diameter of 15.24 cm. The bottom 15.24 cm was a calming section and also a support for the underdrain. The underdrain consisted of a plexiglas plate with

0.32 cm diameter holes which was covered by 50-mesh stainless steel screen. The top 12.7 cm of the filter was a removal inlet section. All three sections were flanged and were connected with brass machine bolts and wing nuts.

The filter operation was controlled by eight 3-way valves. Each filter had two. The position of the valves allowed for either normal downflow filter operation, backwashing or for shut down. The bottom sections of the filters were interconnected with 0.32 cm copper tubing for introduction of air for backwash. A small needle valve controlled the air flow to each filter. The bottom sections of the two coarse media filters each contained a 12.7 cm length of 0.64 cm I.D. copper tubing. This piece of tubing was connected to the air supply valve. The end of the tubing was closed and slits were made along the length of the tube. By using this arrangement better distribution of air during backwash was obtained.

The points for measurement of pressure drop across the filter were located at the inlet/outlet section of the top and bottom sections of each filter. Tygon tubing connected these points to a piezometer board where pressure drop was measured by the difference in water level. For the field investigation, the piezometer board was replaced with pressure gages (0-30 psig).

The flow from each filter was recorded with an individual rotameter. The rotameters were calibrated volumetrically at the beginning of the study. Each rotameter was rated at 9.71 liters/minute at 100% flow. Flow through the entire filter system was controlled with a rate

controller. The rate controller consisted of a small tank with a float controlled valve on the inlet, a needle valve on the outlet controlled the rate at which the water left the tank. Once the outlet valve was set at the appropriate rate the float valve maintained a constant head on the outlet valve. With this arrangement a constant rate could be maintained at any outlet pressure from the filter system. Even though the rate controller kept the system operating at a constant rate, changes in operation dictated whether each individual filter operated at constant rate or declining rate. Constant rate was possible if each set of filters were started and backwashed at the same times. Declining rate was possible if one of the filters was backwashed or started in operation at a different time.

Filter media

The pilot filter system was operated with two types of filters. The first two filters contained 132 cm of coarse sand. The second two filters each contained 38 cm of anthracite coal and 38 cm of silica sand. The first two filters are often termed depth clarifiers and the second two filters were typical dual media filters.

The coarse sand media was prepared by sieve analysis using U.S. Standard sieves. A small amount of the coarse sand was placed on top of a series of sieves and then shaken on a motor driven shaker for about five minutes. The sieved media was then placed in buckets. All the media that passed the number 6 sieve (3.36 mm) and was retained on the number 7 sieve (2.83 mm) was again sieved on the shaker for an additional two to three minutes. The sand media used in the coarse media filters

was retained on the number 7 sieve and passed the number 6 sieve resulting in an average media size of 3.08 mm.

The filter media details for the dual media are given in Table 7. The dual media was prepared by first sieving the media with U.S. standard sieves. Again the motor driven shaker was used for a time period of about five minutes. The media retained on a particular sieve was placed in a bucket. The coal media within the size ranges needed was sieved a second time for three to four minutes. The sand was not sieved a second time. The coal and sand media were mixed individually by first weighing the amount of media in the buckets corresponding to the desired effective size and then adding the appropriate weight of the other media sizes to correspond to the desired grain analysis. After the coal and sand were prepared, the sand was first placed in the filters. The filters were backwashed and approximately one centimeter of fine sand was removed. Next, the coal was placed in the filter and the filter was then backwashed two to three times to remove the fines. The coal used for the dual media filters was supplied by Neptune Microfloc Inc., and the sand was supplied by Northern Gravel Company of Muscatine, Iowa.

Experimental Procedures

Phases of study

The direct filtration study was divided into three phases. The results from a completed phase were used to determine operating conditions in a subsequent phase or phases. The first phase involved using a jar test device to narrow the optimum chemical dosages needed for particle

Table 7. Dual media filter media details

Media	Effective[a] Size d_{10} (mm)	Uniformity[b] Coefficient	d_{90} (mm)[c]
Anthracite	1.0	1.35	1.64
Sand	0.50	1.52	1.0

[a]10% of the media by weight is finer than the effective size, d_{10}.

[b]Uniformity coefficient is the ratio of the d_{60}/d_{10}, 60% by weight of the media is finer than d_{60}.

[c]90% by weight is finer than d_{90}.

destabilization. The second phase involved making laboratory pilot filter runs using either kaolin or calcium aluminosilicate as the turbidity causing particle. The third phase involved pilot filter runs using Des Moines river water as the raw water source.

Jar tests

Jar test studies are often used to determine chemical requirements in water treatment systems. A jar test generally uses six beakers of raw water to which a range of chemical dosages are added. The chemicals and beaker contents are mixed at high speed for a brief period to obtain a homogeneous mixture. After high speed mixing the beaker contents are mixed at a slower speed to promote particle growth to a point where the particles will settle by gravity once the mixing is stopped. The high speed mixing is analogous to the rapid mix tank in a water treatment plant and the slow mixing is analogous to the flocculation tank. After

the beaker contents have been slow mixed, the suspended solids therein are allowed to settle under quiesent conditions for a designated period. The success of a jar test is determined by measuring the turbidity of the supernatant in the beakers.

There are several variations in jar testing procedure which can be used for detailed study of a particular coagulant type or dosage. Once a particular type or dosage of a coagulant has been determined to produce acceptable results, further study can be undertaken to optimize mixing times and intensities. This is done simply by using the same chemical treatment but varying the high speed mixing and/or the slow speed mixing times and intensities. The importance of mixing times and intensities to successful particle destabilization is well documented in the literature and is discussed in the literature review section of this report.

A jar test study can be beneficial when either direct filtration or complete conventional treatment is to be used as the water treatment process. By design, the jar test most closely models complete conventional treatment and is therefore most useful in this case. When direct filtration is to be used, the jar test study may or may not be helpful in determining the appropriate coagulant dose. Often times less coagulant is necessary for direct filtration than is indicated by jar test results. The major role of the jar test as related to direct filtration is the determination of a suitable coagulant or combination of coagulants which will destabilize the particles to be removed.

The jar tests performed for this study were used to find a suitable coagulant or combination of coagulants and the approximate optimum dose or

dosages. The jar test results were then used as a starting point in the pilot filtration runs. All jar tests were performed using a rapid mix speed of 200 rpm for two minutes. The slow mix time was usually 30 minutes but in some tests it was varied at 5, 15 and 30 minutes. The slow mix speed was varied at 20, 50 and 90 rpm for each set of initial conditions. The initial pH of the suspension was adjusted to 7.0 $\pm$ 0.2 with 0.20 N hydrochloric acid. All jar tests were made with synthetic water as the suspending medium. Six 2000 ml glass beakers were used to hold 1500 ml each of the test suspension.

The particles used in the jar test studies were kaolin (Old Hickory number 5 Kentucky Ball Clay) or calcium aluminosilicate. The concentration of the particles to be used in the jar tests was either 10 mg/l or 100 mg/l. The particles were weighed (0.15 g for 10 mg/l suspension and 1.5 g for a 1.0 mg/l suspension) to provide a total volume of the stock suspension of 1000 ml, and then soaked in synthetic water for at least eighteen hours to provide hydration of the particles. After soaking, the particles were dispersed in synthetic water with the Tekmar high speed mixer. The mixing speed was 600 rpm for a period of one minute. One hundred milliliters of the stock suspension was added to 1400 ml of synthetic water to make the appropriate particle concentration for the jar test.

The coagulants used in the jar test study were alum, a cationic polymer, alum plus a cationic polymer and alum plus a nonionic polymer. The only cationic polymer used in the study was Cat Floc T. The nonionic polymers studied were Percol 720 and Dow NP-10. All chemical stock solutions were prepared daily using distilled water. Initially, polymer

solutions were prepared at a concentration of 100 mg/l and then diluted to lower concentrations for use in the jar test. The alum stock concentration was 3.75 g/l.

The data collected during the jar test were limited to settled water turbidity measurements. The results of the settled water turbidity measurements were used to select the optimum chemical pretreatment combination and approximate required chemical dosages for use in the direct filtration experiments in the second phase of the study.

Laboratory filter runs

The second phase of this study involved the laboratory pilot filter experiments. This phase was the major emphasis of the study. The purpose of the pilot filter runs was to evaluate the effectiveness of the pilot filter system in treating high turbidity water by direct filtration. In all but two filter runs the suspension material was kaolin. One run was made with calcium aluminosilicate as the suspension material and one run was made with both calcium aluminosilicate and kaolin as suspension materials. The kaolin material was chosen as the primary material since it had been used in previous experiments and its characteristics and properties were well documented. In addition, since kaolin is a clay material and most high turbidity associated with surface waters contains a large portion of clays, it seemed justifiable to use this material. The water used to make the raw water suspensions was university tap water. It was necessary to use the tap water rather than synthetic water due to the large volumes required. As much as 900 l/hr was used at the highest flow rate.

The suspension material was weighed out and allowed to soak in tap water for at least 18 hours before the experiment. The material concentrations used were 25, 50, 100, 200 and 300 mg/1. The material was then dispersed with the high speed Takmar mixer at 6000 rpm for one minute. The dispersed material was then added to 2500 1 of tap water in one of the two raw water tanks. The mixers on the tanks dispersed the material and kept it in suspension. One raw water tank was used as the raw water feed tank and the other tank was used for raw water storage. When only 600 1 of water remained in the feed tank, the water in the storage tank was pumped into the feed tank. The recirculation necessary for the constant head tank, located on the feed tank, also helped keep the particles in suspension. The suspension pH was adjusted to 7.3 $\pm$ 0.3 with 12 N hydrochloric acid. This was done in an attempt to have a pH of 7.1 $\pm$ 0.1 after alum addition. The control of pH was a minor problem throughout the study. There was a tendency for the pH to raise slightly as the feed tank emptied.

The raw water was pumped from the bottom of the feed tank to a constant head tank. Excess water flowed back into the feed tank. From the constant head tank the raw water flowed by gravity to the rapid mix tank where the alum was added. After rapid mixing for 30 seconds (filtration rate = 12.2 m/hr), the water flowed by gravity to the filters. Polymer was added in the feed line to the filters. The flow was split prior to the coarse media filters. After passing through the coarse media filters, flow was recombined and a sample was taken for analysis of its turbidity. The

flow then was again split prior to the dual media filters and then recombined prior to taking the filtered water sample for turbidity analysis before passing through the rate controller. The amount of flow to each filter depended on the mode of operation. Some runs were made at constant rate while others were made using a declining rate operation. Headloss measurements were made with piezometer tubes located at the tops and bottoms of each filter.

In general, turbidity and headloss were recorded hourly during a run. The turbidities of the raw water, the coarse filter effluent and the final effluent were measured. At various times during a filter run, zeta potential measurements were taken of the particles in the raw water, the rapid mix tank, the coarse media effluent and the final effluent. Data obtained from the jar test experiments were used to determine chemical dosages for the initial filter runs using alum and Percol 720. As data were collected from the filter runs, chemical dosages were adjusted to optimize the filtration process. The first four filter runs confirmed that alum combined with a nonionic polymer was a very effective combination for removal by direct filtration of up to 100 mg/l of kaolin. Most subsequent filter runs used a combination of alum and nonionic polymer to study the effect of various concentrations of kaolin and the effect of different operating modes. The nonionic polymer used in most filter runs was Percol LT-20. This polymer is similar to Percol 720 with, reportedly, the only difference being that Percol 720 has an amine group attached to the polymer chain. Several runs were repeated under the same conditions to confirm laboratory procedures and reproducibility

of results. Other runs were made with different polymers or combinations of alum and polymer.

The purpose of the latter filter runs was to collect additional data to develop a relationship between optimum chemical dosage, kaolin concentration and flow rate. To do this the kaolin concentration was varied from 25 to 100 mg/l and flow rates of 12.2, 18.3 and 24.2 m/hr were used to filter the various kaolin concentrations. At a flowrate of 24.2 m/hr, only the 25 mg/l kaolin concentration was used.

Filter backwash

The criteria used to determine when a filter needed to be backwashed were based on either a maximum filter effluent turbidity or a limiting maximum headloss. One or both of the coarse media filters, depending on the operational mode, were backwashed when the combined coarse media filter effluent turbidity exceeded approximately one third of the influent turbidity for two consecutive sampling periods. One or both of the dual media filters, depending on the operational mode, were backwashed if the combined dual media filter effluent turbidity exceeded 1 NTU for two consecutive measurements and showed no indication of decreasing. When all four filters of the filtration system were in operation,a filter was backwashed when a terminal headloss of 2 m was reached in an individual filter. When only the coarse media filters were operated in parallel or when one coarse media and one dual media filter were operated in series, the terminal headloss was increased to 3 m. During various filter runs the operating conditions were such that the above criteria was not applicable. In these cases, the filtration system

was operated for a certain time period which was long enough to establish a trend in the filtration system performance.

The procedures followed for backwashing the coarse media and dual media filters remained consistent throughout the study period. No attempt was made during the study to optimize backwash air and water rates. The air and water rates used were those that assured a relatively clean media after backwash. When a filter needed to be backwashed during a filter run, the flow to that filter was reduced to zero over approximately a 10 minute period. This reduced the effect of rate change on the solids holding capacity in the parallel filter since that filter received the total flow during the backwash period. Backwashing a coarse media filter involved an initial 2 minute water wash at approximately 78 m/hr followed by a 3-4 minute combined air and water wash at the same rate. A final 30-60 second water wash at a rate of 106 m/hr was used to remove the entrapped air from the filter media. The high rate water wash expanded the media approximately 10%. After backwashing, the media depth was sometimes higher than the original level and had to be reduced by tapping on the filter column.

Backwashing a dual media filter involved first lowering the water level to just above the media surface. An air scour was used to break apart the solids layer that often formed on the surface of the media. After the air scour, a water wash at a rate of 78 m/hr was used to remove the accumulated solids. The length of the water wash varied depending on how dirty the media was and the chemical treatment used. Normally the backwash period lasted about 10 minutes. Dual media expansion during backwash was approximately 80 to 90%.

After a filter run was terminated, all filters were backwashed until the backwash water appeared clear. When filter runs were not performed back to back, a chlorine solution was passed through the filter media and allowed to remain in contact with the media overnight. After the chlorine treatment, the filters were again backwashed to remove the chlorine solution. The purpose of adding the chlorine solution was to ensure the removal of polymer and associated solids from the filter media. If a filter run was not initiated at this time, the filters were backwashed again prior to the next filter run.

Photographs of coarse media and particle deposits

Photographs of the media and particle deposits in the coarse media filters were taken during Runs 26 and 27. A 35 mm camera with a 50 mm lense in combination with closeup lense filters was used to photographically record the stages of particle deposition within an operating filter.

Photographs were taken at various times during the filter runs to represent particle deposition characteristics at the beginning, middle, and end of the filter runs. Depending upon the time of filter operation and the influent particle concentration, photographs of the top, (surface to 7 cm) upper (12 cm from surface), middle (65 cm from surface), and lower (95 cm from surface) portions of the filter column were taken. By taking photographs at different depths along the filter columns, the shift in the burden of solids removal with filtration time to deeper depths in the media was recorded. A photograph was also taken of the

surface of the filter. This photograph showed the entrance points to the channels that developed in the filter.

The magnification in the photographs could not be determined precisely since the camera and tripod were moved between photographs and since there was some distortion caused by the curvature of the filter columns. An approximate magnification can be estimated by comparing the size of the media grains in the photographs to the average media grain size (3.08 mm). In general, the magnification was about 1.1 x to 1.2 x.

The photograph of the surface of the filter was taken after partially draining the filter and removing the top plexiglas cover. The filter was drained to prevent distortion that could have been caused by the water above the surface.

Field study

The third phase of this study involved pilot filter runs using Des Moines river water at Boone, Iowa, as the raw water source. This phase of the study had originally been planned to take place when high turbidity conditions existed during spring runoff. Unfortunately, high turbidity conditions did not occur during the field study which was initiated on April 7 and lasted through April 12. Normally, an average river water turbidity of 67 NTU can be expected in April, but during the study period the average river water turbidity was 28 NTU. This abnormal pattern of low turbidity was due to the below normal spring precipitation and below normal snowmelt. Even though the turbidity was lower than normal, the filter system and all appurtenances were moved to the Boone Water Works northwest of Boone, Iowa. The filter system was set up outside next to

a USGS gaging station located within 15 meters of the Des Moines River. Electrical power and backwash water and air were provided by the City of Boone. The Iowa State University Water Resources Mobile Laboratory was used as the laboratory facility.

A raw water intake line was placed in the river approximately 6 meters from shore and 6 meters upstream of the low head dam located at the Boone waterworks. The intake was approximately 0.5 meter below the water surface. A centrifugal pump brought the raw water through a 1.9 cm intake hose to the filter apparatus. Alum was added after the pump by injecting it into the feed line. After the injection point of alum, the flow passed through an in-line motionless mixer to homogeneously mix the alum and raw water. There was approximately 8 meters of hose between the in-line mixer and the polymer injection point. After the polymer injection point, there were approximately another 8 meters of hose to the filter apparatus. No mixing other than that which occurred in the hose and filter intake was provided after the polymer addition. The filter operation and sampling procedures were the same as for the laboratory pilot filter runs.

The filter runs at Boone were made at a flowrate of 12.2 m/hr. At the beginning of the run, the pump pressure was maintained at 10 psig. At this pressure the flowrate could not be maintained consistently so the pressure was increased to 15 psig and no further problems were encountered. Initially, alum and Percol LT-20 were used as the coagulants in this phase. Due to poor performance of this combination of coagulants, additional combinations were tried. Two cationic polymers, Cat Floc T

and Percol LT-22, in combination with alum were evaluated. The problem with the alum and LT-20 was attributed to the nature of the turbidity causing particles. At Boone, the majority of the turbidity was attributable to diatoms.

RESULTS

Flocculation Time

The first portion of this investigation made use of the jar test. The jar test was used to determine the effects of chemical dosage, particle type, mixing speed and mixing time on settled water turbidity. Jar test results were evaluated on the basis of settled water turbidity. Flocculated turbidity was taken in some instances but this parameter provided little consistent and reproducible data. The particles selected for jar test evaluation were kaolin and calcium aluminosilicate. Both particles were tested at concentrations of 10 mg/l and 100 mg/l. These concentrations were selected to be representative of a low and high turbidity water.

The effects of flocculation time on the settled water turbidity are shown in Figure 6. In both tests,the initial kaolin concentration was 100 mg/l, the rapid mix speed was 200 rpm for 2 min. and the flocculation speed was 50 rpm. A settling period of 30 minutes followed the flocculation period. The alum dosages were less in the test at pH 6. The results indicate that a flocculation time of 5 minutes, when the pH is 8, results in a significant reduction in settled water turbidity. Flocculation beyond the 5-minute period can be concluded to have limited effect on the final turbidity values. At a pH of 6, the results are not as clearly distinguishable. A 5-minute flocculation period did result in significant reduction of settled water turbidity but further flocculation resulted in still lower values. When 10, 15 and 20 mg/l of

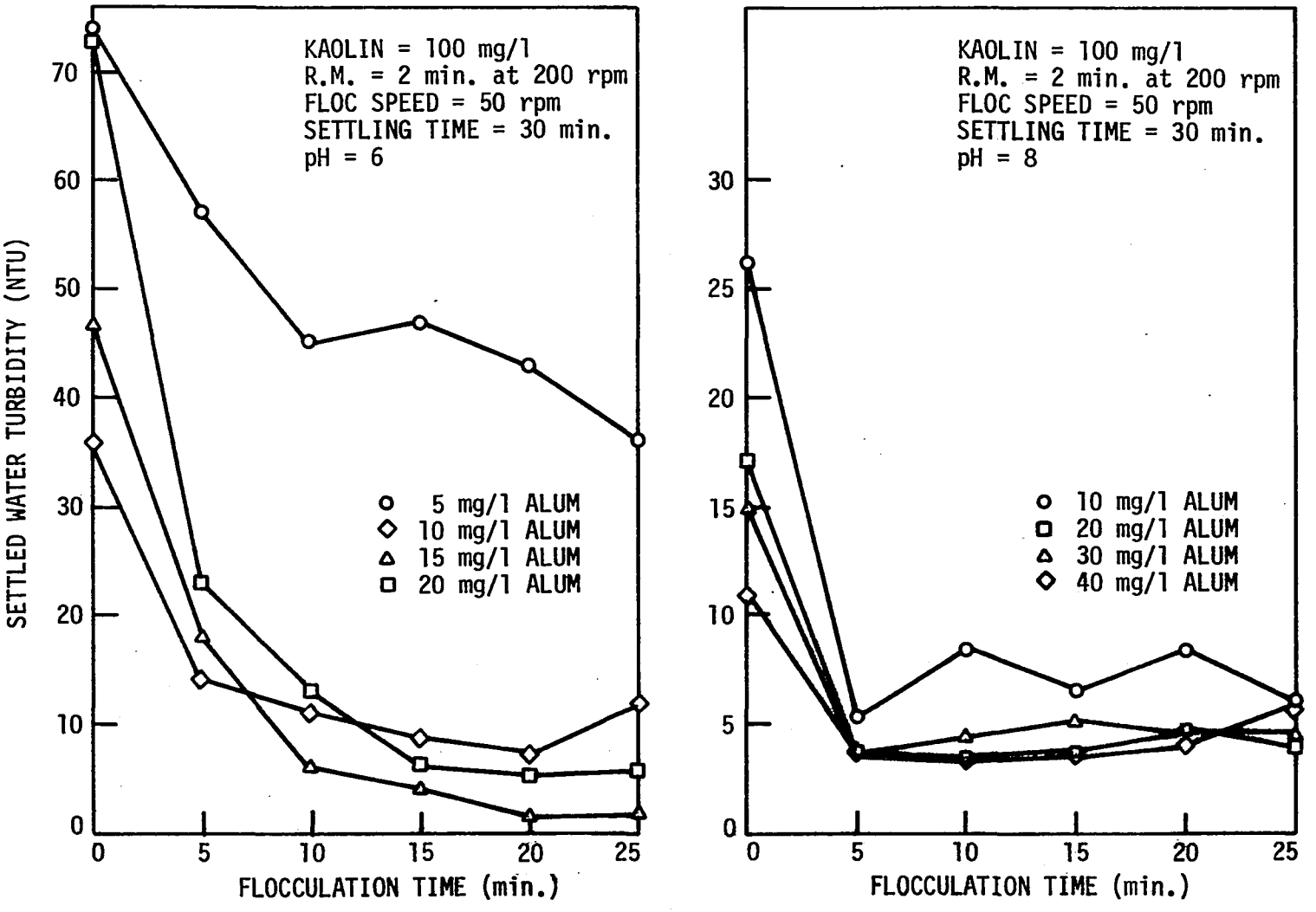


Figure 6. Settled water turbidity vs. flocculation time and alum dosage at raw water pH's of 6 and 8

alum were used, a flocculation time beyond 15 minutes did not produce much additional reduction in settled water turbidity. When 5 mg/l alum was used, an additional turbidity reduction was noticeable up to the total 25 minute flocculation period. The zeta potential values of the particles in the flocculated suspensions are given in Figure 7. Within the limits of experimental error the zeta potential did not change with any degree of consistency as the flocculation period was varied. At a pH of 8, all zeta potential values were negative except when 40 mg/l of alum was used and this dosage resulted in positive values. At a pH of 6 all particle zeta potential values were positive except at an alum dosage of 5 mg/l where the values were negative.

It is difficult and sometimes dangerous to draw too many conclusions from limited data, such as that depicted in Figures 6 and 7. At a pH of 6, destabilization with alum will occur by charge neutralization and at a pH of 8 destabilization will occur by sweep coagulation or sweep coagulation in combination with charge neutralization (Figure 4). For the conditions described in the jar tests, it would appear that sweep coagulation in combination with charge neutralization is less affected by flocculation time. Destabilization by charge neutralization may require additional flocculation time for adequate particle coagulant interaction to occur under the relatively low speed mixing conditions used in the jar tests. When destabilization is accomplished by charge neutralization, Amirtharajah and Mills (3) recommend the use of a high speed and short duration rapid mix since adsorption of alum hydrolysis species to produce charge neutralization is thought to occur in less than

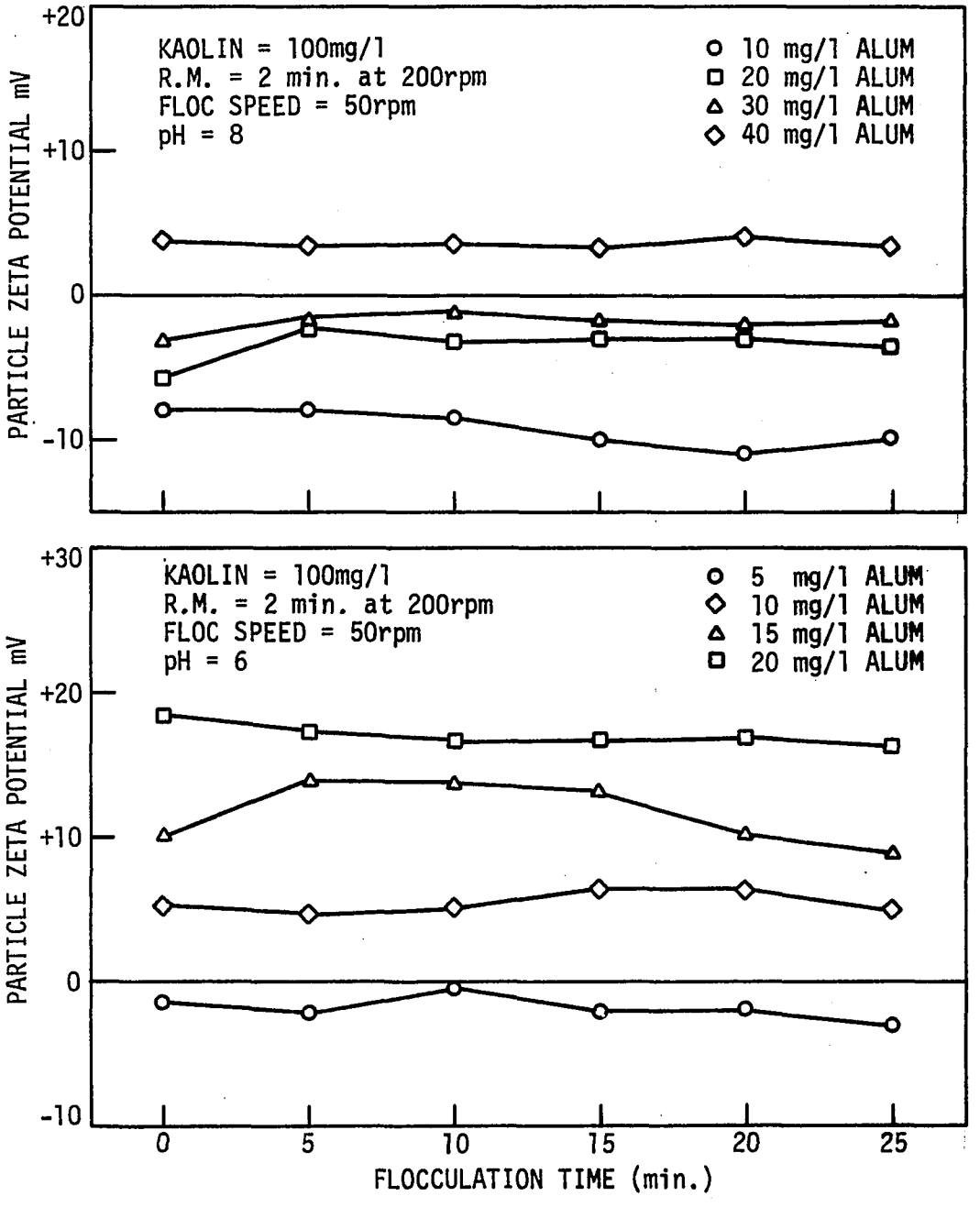


Figure 7. Particle zeta potential vs. flocculation time and alum dosage at raw water pH's of 6 and 8

1 second. A second point that can be illustrated from the data presented in these figures is that of zeta potential control for coagulation processes. Investigators have argued both for and against the applicability of zeta potential as a control parameter in coagulation and filtration. Those arguing for zeta potential control have tried to generalize its application to all systems and those arguing against zeta potential control have found no real correlation to optimum operation within their systems. Most often a zeta potential of zero or slightly negative or positive ($\pm$ 5 mV) is proclaimed as an optimum range. It can be seen from Figures 6 and 7 that problems would be encountered if settled water turbidity would be correlated to particle zeta potential. As an example, in Figure 7, an alum dosage of 5 mg/l at a water pH of 6 results in a slightly negative zeta potential. A similar particle zeta potential value is shown at a pH of 8 for alum dosages of 20 and 30 mg/l. Comparing the curves in Figures 6 at pH values of 6 and 8 clearly shows the difference in settled water turbidity.

Four jar tests were performed to determine the zeta potential of kaolin at six different alum dosages. Clay concentrations were varied at 25, 50, 100 and 200 mg/l. The water used was university tap water. The pH was adjusted with 0.2 NHCL to obtain a pH of 7.0 $\pm$ 0.2 after alum addition. However, the suspension pH when 30 mg/l of alum was used varied from 6.0 to 6.75. Values for particle zeta potential are not reported for a kaolin concentration of 100 mg/l and an alum dose of 7.5 mg/l since the pH adjustment on this sample was overlooked. The beaker contents were mixed at 200 rpm on the six paddle stirrer. Alum was added to the beaker contents and after 30 seconds a sample was withdrawn

Table 8. Particle zeta potential for varying concentrations of kaolin and alum with and without polymer

Alum Dose, mg/l	Kaolin Concentration, mg/l: 25	50	100	200	$\bar{x}$ [a]	s [b]
	Zeta Potential, mV					
0	-20.0	-22.0	-21.4	-19.3	-20.7	1.2
w/polymer	-19.4	-19.6	-20.9	-20.9	-20.2	0.8
5	-13.7	-17.5	-17.7	-16.8	-16.4	1.9
w/polymer	-14.0	-14.1	-15.5	-14.4	-14.5	0.7
7.5	- 9.6	- 9.4	-	-11.7	-10.2	1.3
w/polymer	- 9.5	- 9.2	-	-13.7	-10.8	2.5
10	- 5.1	- 4.1	- 6.4	- 8.7	- 6.1	2.0
w/polymer	- 5.0	- 4.3	- 7.7	- 8.2	- 6.3	1.9
15	0	+ 0.7	0	0	+ 0.2	0.4
w/polymer	+ 0.1	0	0	0	+ 0.025	0.05
30	+11.6	+ 9.9	+ 8.9	+ 8.1	+ 9.6	1.5
w/polymer	+12.2	+ 7.3	+ 8.6	+ 7.7	+ 9.0	2.2

[a] Mean value.

[b] Standard deviation.

for zeta potential measurement. The analysis took about 2.5 minutes to complete. After this period 0.5 mg/l of Percol LT-20 was added to the beaker contents and after 30 seconds of mixing a second sample was withdrawn for particle zeta potential measurement. The results of this series of jar tests are given in Table 8.

As would be expected, the results show that increasing alum dosages, changes the particle zeta potential from negative to a more positive value. With 15 mg/l of alum, the zeta potential is zero. Charge reversal occurs when 30 mg/l of alum is used. There are some differences between particle zeta potential values reported for the different kaolin concentrations. These differences are most noticeable at alum dosages of 7.5 and 10 mg/l where the magnitude of the zeta potential values is relatively low. These differences may show a particle zeta potential dependance on concentration at these alum dosages or they may be the result of experimental error which is often more evident when the zeta potential values are low. The values reported with and without polymer show the same trend at alum dosages of 7.5 and 10 mg/l. The particle zeta potential values before and after polymer addition generally indicate that the nonionic polymer has little affect on the zeta potential values and there is no definite trend in the differences between the values. Both positive and negative changes were noted. The zeta potential values reported at an alum dosage of 5 mg/l may be an exception to the previous conclusions. At this alum dosage, and at all four clay concentrations, the zeta potential was noticeably less negative after the polymer was added. This could be due to the polymer or be the result of the additional mixing period between analyses.

Alum Requirements

A series of three jar tests was used to find an approximate alum dosage for use in the filtration experiments. The jar tests evaluated the effect of alum dosage on the settled water turbidity with 10 and 100 mg/l of both kaolin and calcium aluminosilicate. In all tests, the rapid mix period was 2 minutes at a speed of 200 rpm. An average flocculation period of 30 minutes was used as a convenient operation time so that analyses could be done on individual beaker contents and still have little effect on the results from the other beakers. Previous jar tests had indicated essentially no advantage to extending the flocculation period beyond 5 to 15 minutes but these tests also indicated no detrimental effects resulted with the longer flocculation period. The beaker contents were allowed to settle for 30 minutes after the flocculation period and the results were evaluated on the basis of settled water turbidity. The pH of the beaker contents was adjusted to approximately seven using synthetic water. Flocculation speeds of 20, 50, and 90 rpm were used to represent a range of mixing intensities.

In general, the results showed that at low particle concentrations a medium to high energy flocculation period is desirable at alum dosages less than approximately 10 mg/l. At the higher particle concentrations, the low and medium mixing resulted in the best settled water turbidities. From the jar test data it appeared that the optimum alum dosage for kaolin at 10 mg/l was 5-10 mg/l alum and for kaolin at 100 mg/l the optimum alum dosage was about 20 to 25 mg/l. The optimum alum dosage

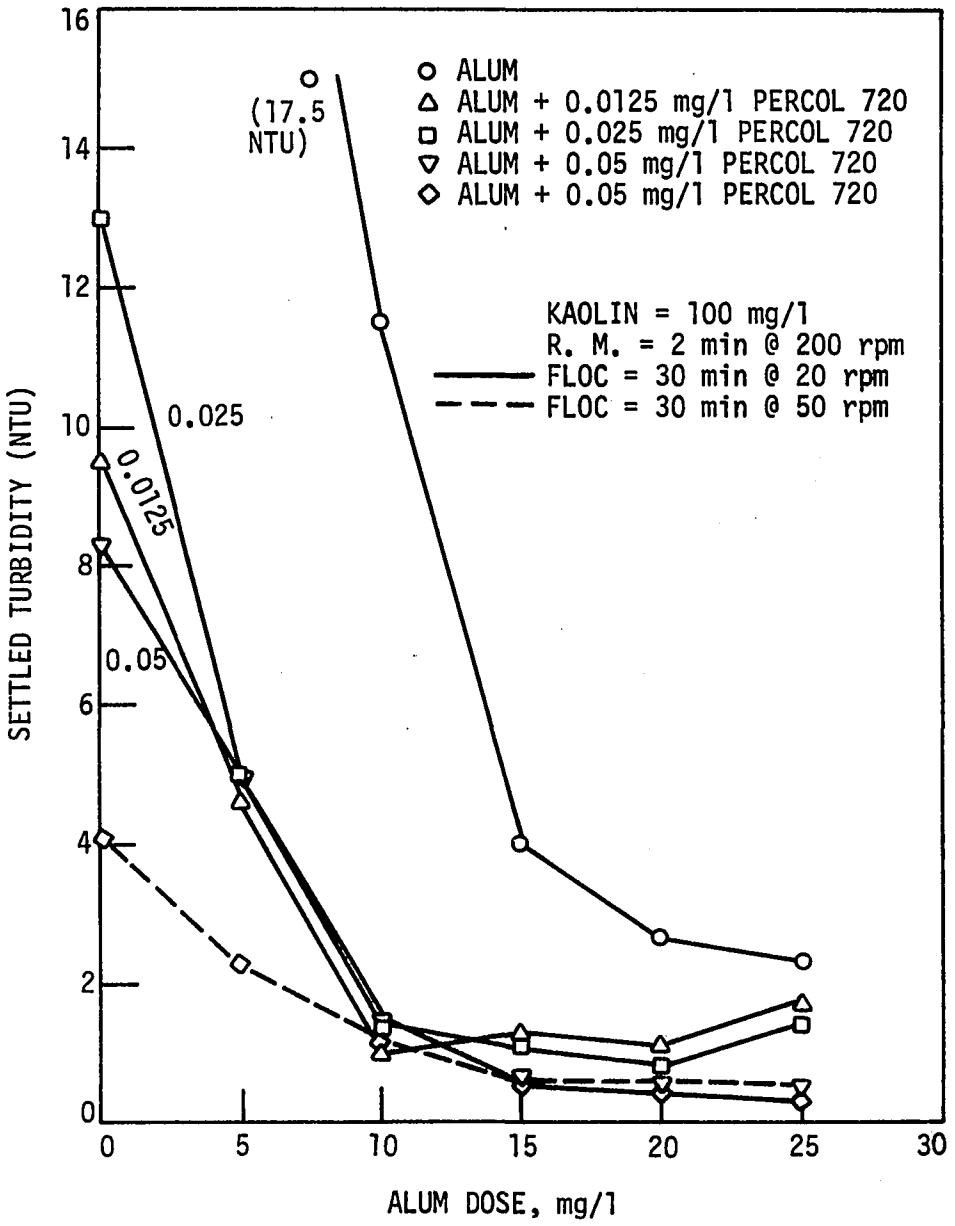


Figure 8. Settled water turbidity of a 100 mg/l kaolin suspension using alum and Percol 720 as coagulants

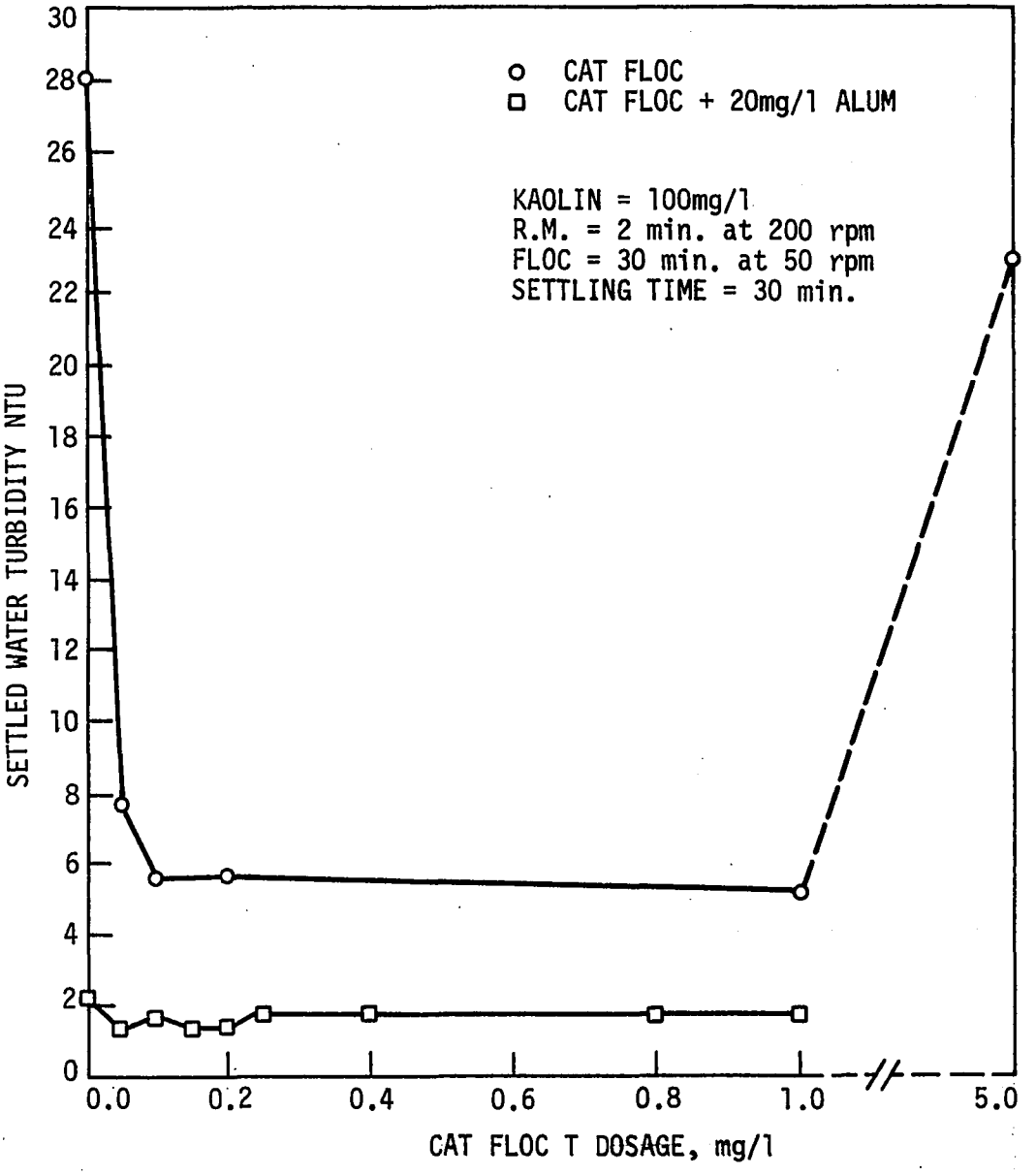


Figure 9. Settled water turbidity of a 100 mg/l kaolin suspension using alum and Cat Floc T as coagulants

for 10 mg/l of calcium aluminosilicate was 5 to 10 mg/l and for 100 mg/l of calcium aluminosilicate the optimum alum dosage was 10 to 15 mg/l, although very little difference in settled water turbidity was found between 5 and 10 mg/l of alum for this particle concentration.

Polymer Requirements

Jar tests were performed with three polymers. Two of the polymers were nonionic and the third polymer was a cationic. The tests were performed with synthetic water, a water pH of 7, a rapid mix period of 2 minutes at 200 rpm, a settling period after flocculation of 30 minutes and a flocculation period of 30 minutes at the speed designated on the figure. Figure 8 shows the results when kaolin was the suspension particle.

The selection of an approximate dosage of Percol 720, a nonionic polymer, was based on the work of Stephenson (86). The results given in Figure 8 for 100 mg/l of kaolin and various combinations of alum and Percol 720 show the effects of the polymer and mixing intensity. At all polymer dosages evaluated, a significant reduction in settled water turbidity occurred when compared to the use of alum alone. The difference was most noticeable at the lower alum dosage. The effect of mixing intensity on settled water turbidity was most obvious for alum dosages below 10 mg/l at the high kaolin concentration. At alum dosages greater than 10 mg/l, mixing speed and polymer concentration, within the range studied, had little effect on settled water turbidity results.

Figure 9 presents the results of the jar tests using 100 mg/l of

kaolin and Cat Floc T, a cationic polymer, in combination with 20 mg/l of alum and as the sole coagulant. It can be seen from this figure that the alum and Cat Floc T combination produces the best results compared to Cat Floc T alone. It should be noted that with the Cat Floc T and alum combination there is little change in settled water turbidity throughout the Cat Floc T dosage range tested. The settled turbidity was between approximately 1.5 to 2.0 NTU. Twenty milligrams per liter of alum could produce this turbidity as shown in Figure 8, without the Cat Floc T present. When Cat Floc T was used as the sole coagulant it appears that a significant improvement in settled water turbidity occurred at a polymer dosage of 0.1 mg/l. Although limited data were collected for polymer dosages between 0.1 and 1.0 mg/l there appeared to be little benefit derived from increasing the polymer dosage above 0.1 mg/l. Five mg/l of Cat Floc T resulted in particle restabilization and therefore poor removal.

Filter Runs

The purpose of the filtration runs was to evaluate the performance of coarse media filters in series with standard dual media filters for direct filtration of high turbidity influents. The major emphasis was the performance of the coarse media filter and how this performance affected the dual media filters headloss and effluent turbidity. Even though the present water quality standards dictate that the final effluent turbidity be 1 NTU or less, a successful filter run during this study was one in which the filtered water turbidity was 0.1 NTU or less. Run lengths in excess of 10 hours were considered necessary at all filtration rates and influent solids concentrations evaluated.

Coarse media and dual media filter performance were evaluated on the basis of effluent turbidity and total headloss. The effluent turbidity values represent combined effluents from each set of filter media types. Headloss values were measured across each filter. Particle zeta potential was also measured during the course of the study to aid in evaluation of chemical treatment on filter performance. A total of twenty eight filter runs were made during the laboratory filter run portion of this study. The filter runs were designed to evaluate the effects of coagulant type and dosage, influent turbidity, filtration rate and operational mode on filter performance. The operating conditions for each filter run are given in Table 9.

The results of selected filter runs are presented in graphical form. It should be noted that time zero represents, on the average, an elapsed time of 20 minutes after the system was started. Some time was needed to check flowrates and chemical feed rates and then to allow the system to equilibrate to the adjustments.

Run 1

The purpose of Run 1 was to check the filter system operation and to determine whether the jar test results from the previous phase were indicative of the required alum dosage. The optimum alum dosage, as determined by the jar test, was about 20 mg/l. In Run 1, alum dosages of 10, 20, and 25 mg/l were used. The turbidity and headloss results are shown in Figures 10 and 11. The coarse media filter effluent turbidity results indicated that an alum dosage of 10 mg/l was too low

Table 9. Summary of operating conditions for laboratory filter runs

Run No.	Flow Rate m/hr	Particle Type	Particle Concentration mg/l	Avg. Influent Turbidity NTU	Alum Dose mg/l	Polymer Type	Polymer Dose mg/l
1	12.2	kaolin	100	99	10,20,25	-	0
2	12.2	kaolin	100	91	20	Percol 720	0.05
3	12.2	kaolin	100	93	20	LT-20	0.05
4	12.2	kaolin	100	93	20	LT-20	0.05
5	12.2	kaolin	100	93	20	LT-20	0.10
6	12.2	kaolin	100	76	10,15	LT-20	0.10
7	12.2	kaolin	100	84	20	NP-10	0.05, 0.20
8	12.2	kaolin	100	82	0	-	0
9	12.2	kaolin	100	79	20	Percol 720	0.05
10	12.2	kaolin	100	79	20	Cat Floc T	0.15, 3.64
11	12.2	kaolin	100	83	15	LT-20	0.05
12	12.2	kaolin	100	81	15	LT-20	0.05
13	12.2	kaolin	100	80	15	LT-20	0.05
14	12.2	kaolin	300	224	20-40	LT-20	0.05, 0.20
15	12.2	kaolin	100	81	10	LT-20	0.05

16	12.2	kaolin	100	85	0	LT-20	0.05
17	12.2	calcium aluminosil-icate	100	32	10	LT-20	0.05
18	12.2	kaolin	200	149	20	LT-20	0.05
19	12.2	kaolin calcium alum-inosilicate	100	99	15	LT-20	0.05
20	12.2	kaolin	50	37	5,7.5,10,15	LT-20	0.05
21	12.2	kaolin	25	19	7.5	LT-20	0.05
22	18.3	kaolin	25	18	7.5	LT-20	0.05
23	18.3	kaolin	50	37	7.5	LT-20	0.05
24	24.4	kaolin	25	20	7.5	LT-20	0.05
25A	12.2	kaolin	100	75	7.5	LT-20	0.05
25B	12.2	kaolin	100	77	15	LT-20	0.05
26	18.3	kaolin	25	20	7,5,15	LT-20	0.05
27	12.2	kaolin	100	89	15	LT-20	0.01, 0.05

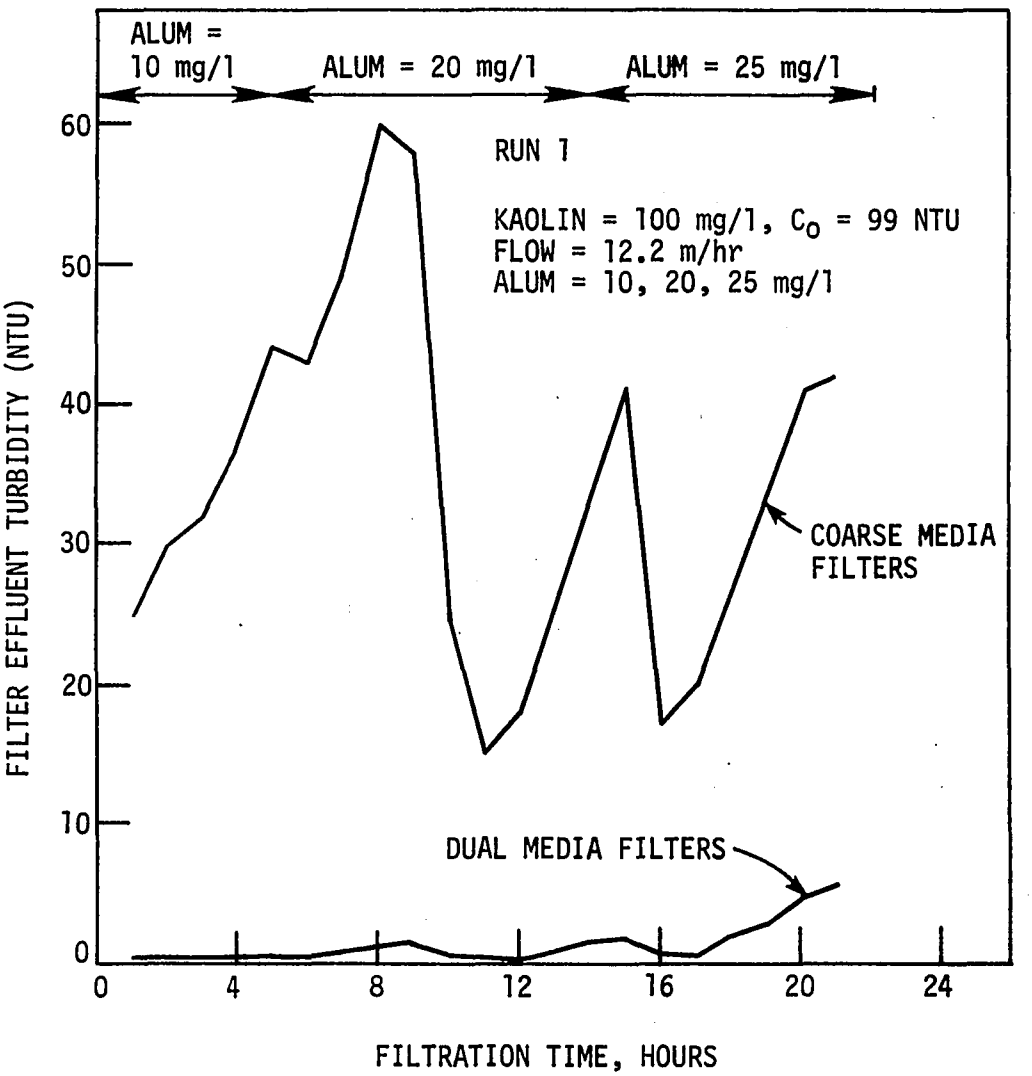


Figure 10. Effluent turbidity vs. filtration time, Run 1, coarse media and dual media filters operated at constant rate

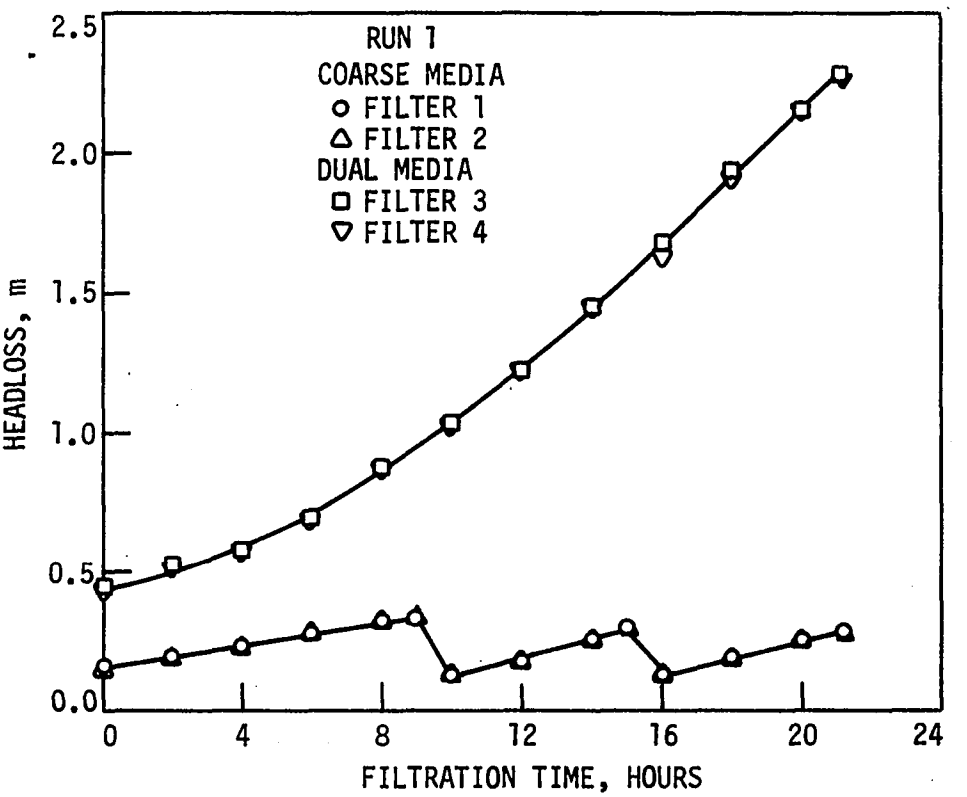


Figure 11. Headloss increase vs. filtration time, Run 1, coarse media and dual media filters operated at constant rate

since the coarse media effluent turbidity started at a relatively high level and then increased rapidly. With alum dosage of 20 mg/l some improvement was obtained. The initial turbidity was lower than it had been when the alum dosage was at 10 mg/l. At an alum dosage of 25 mg/l the initial turbidity was about the same as when 20 mg/l of alum was used. The headloss data indicated that depth filtration was occurring in the coarse media filters. Depth filtration is implied by a relatively straight line headloss curve. The headloss data for the dual media filters plotted somewhat exponentially which would indicate some surface removal, but the exponential headloss characteristics were likely the result of the variable influent concentration. It was concluded from this filter run that alum alone would not be satisfactory as the sole coagulant. The coarse media filter effluent turbidity was high and little headloss developed in the coarse media filters.

Runs 2 and 4

Runs 2 and 4 were made with the same kaolin concentration, flowrate and chemical treatment. During Run 2, the coarse media filters were operated at a constant rate but during Runs 4 these filters were operated in a declining rate mode. Declining rate operation was achieved by beginning a filter run with only one coarse media filter and both dual media filters. The second filter was put into operation later, usually after three hours. Due to this staggered start-up period, the one filter operated at twice the nominal filtration rate during this period. The turbidity and headloss data for these runs are shown in Figures 12

through 15. The filter flowrates during Run 4 are shown in Figure 16.

The effects of the use of the nonionic polymer are well demonstrated by the results of these two runs. The amount of solids retained in the coarse media increased significantly as evidenced by the coarse media headloss and effluent turbidity data. The headloss in the coarse media filters was for the most part linear with time of filtration during Run 4 but in Run 2 a slight curvature is evident in the data which would indicate some surface removal. Operating at a constant rate resulted in a rapid turbidity breakthrough in the coarse media filters. By operating in a declining rate mode the abruptness of the breakthrough was reduced but the coarse media effluent turbidities were not as low as they were when constant rate operation was used. The effect of the low coarse media effluent turbidity for five to six hour durations during Run 2 is illustrated by the extent of headloss in the dual media when compared to dual media headloss during Run 4. During a 23 hour period in Run 2, the net headloss increase in the dual media filters was about 0.5 m while in Run 4 the headloss increase for the same time period was about 2 m. The effluent turbidity from the dual media filters was equivalent in both filter runs, except for the period immediately after backwashing the coarse media filters in Run 2. Most values were below 0.1 NTU except during the first two hours of Run 4 when the coarse media effluent turbidity was high and in Run 2 when the coarse media filters were backwashed. The backwashing of the coarse media filters when the filters are operated at a constant rate usually resulted in stripping

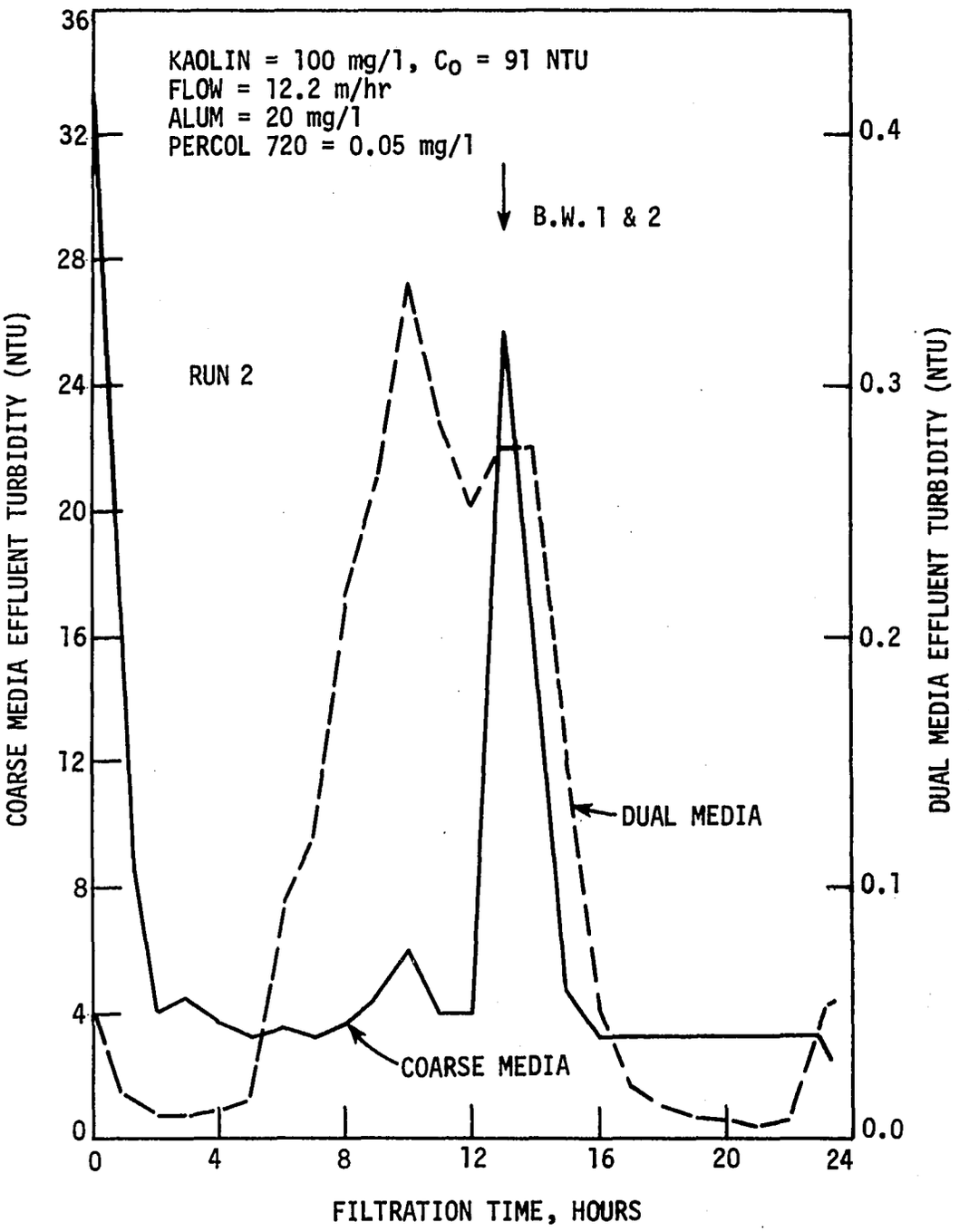


Figure 12. Effluent turbidity vs. filtration time, Run 2, coarse media and dual media filters operated at constant rate

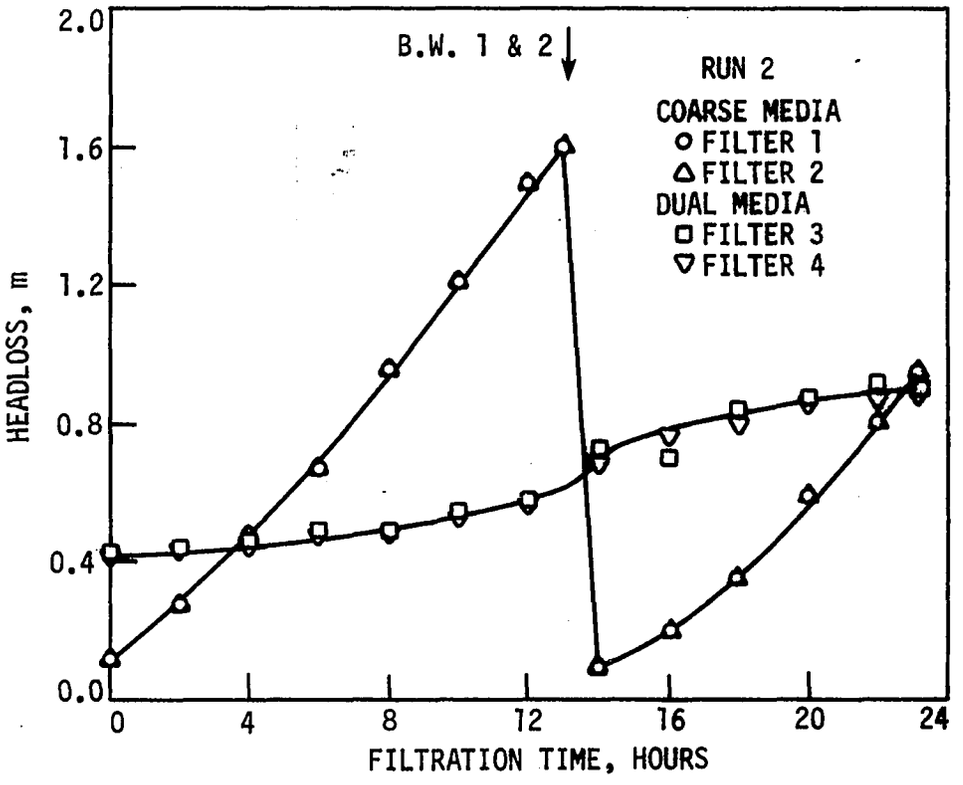


Figure 13. Headloss increase vs. filtration time, Run 2, coarse media and dual media filters operated at constant rate

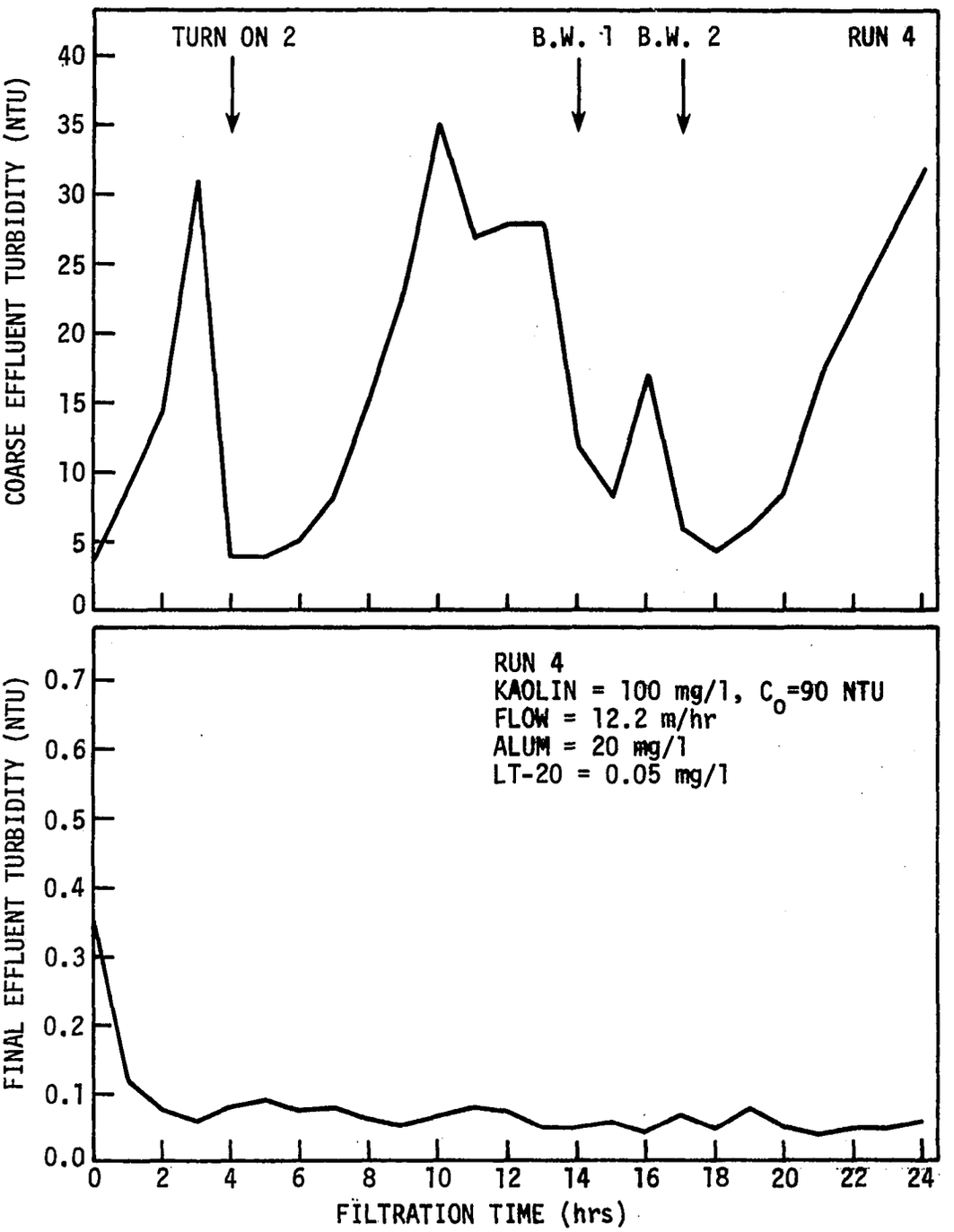


Figure 14. Effluent turbidity vs. filtration time, Run 4, coarse media filters operated in declining rate and dual media filters operated at constant rate

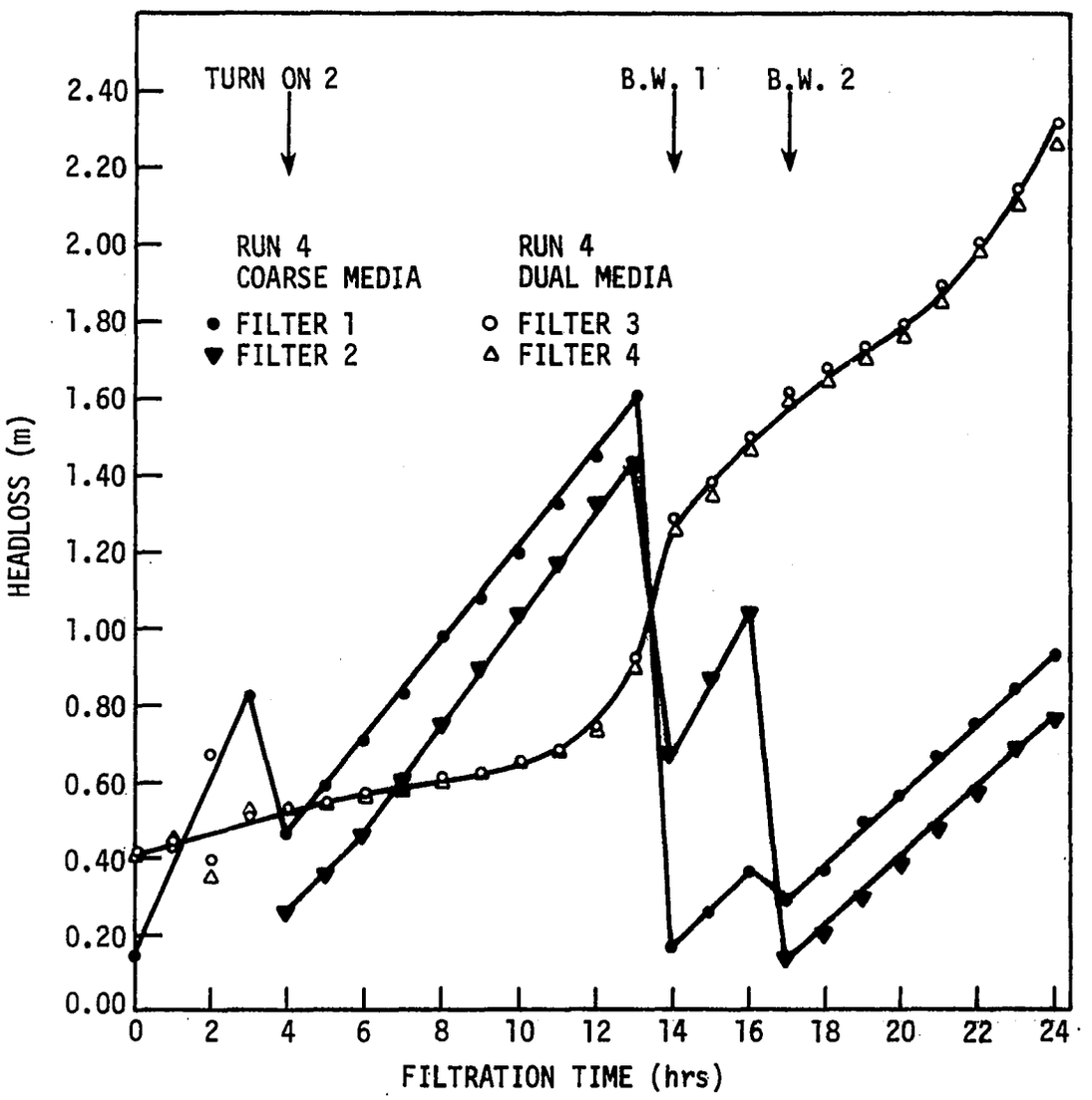


Figure 15. Headloss increase vs. filtration time, Run 4, coarse media filters operated in declining rate and dual media filters operated at constant rate

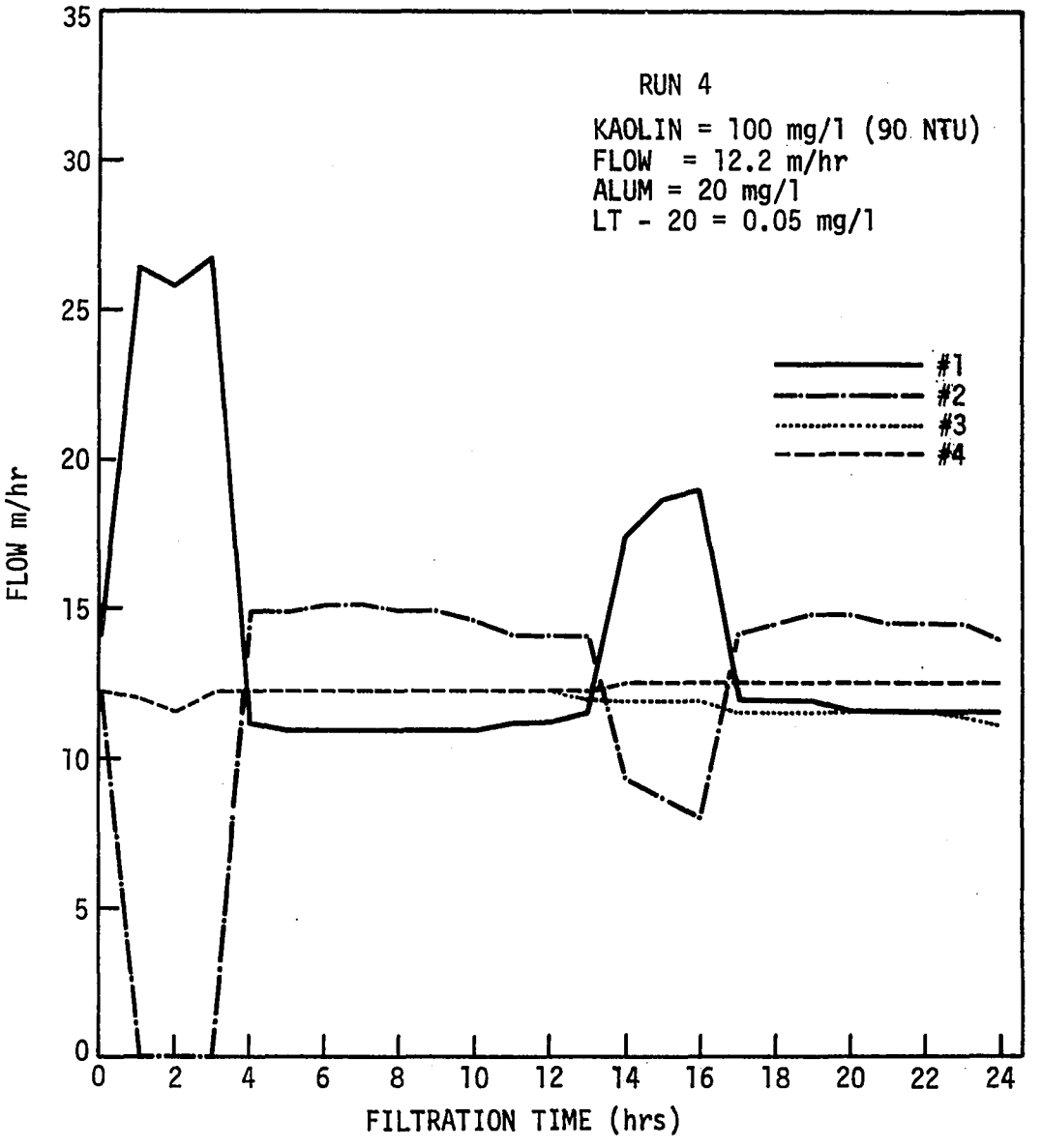


Figure 16. Filter flow rate variations, Run 4, coarse media filters operated in declining rate and dual media filters operated at constant rate

of solids from the operating filter when the other filter was removed from service for backwashing. This is normal and would be expected due to the 100 percent increase in the flow rate to the coarse media filter still in operation. At times, the amount of solids due to the increased flowrate significantly increased the coarse media filter effluent turbidity. The increase in dual media filter effluent turbidity during Run 2 is evidence of the increased dual media filter influent turbidity. The increase in rate of dual media headloss in both Run 2 and Run 4 after coarse media backwash is also indicative of the increased coarse media effluent turbidity. The impact on the dual media headloss during Run 2 appeared to be less than in Run 4. The extent of the impact can be attributed to the fact that the dual media filters during Run 4 were receiving a higher solids loadings during the run while the dual media filters during Run 2 received relatively lower turbidities and therefore had a higher storage capacity at the time the coarse media filters were backwashed. During the coarse media filter backwash in Run 2, the dual media filters apparently were more susceptible to turbidity breakthrough when the high solids load was received.

The flowrate through all filters during Run 2 was constant. During Run 4 the flowrate was constant only in the dual media filters. The changes in flowrate in the coarse media filters is shown in Figure 16. The combined flowrate in the coarse media filters may appear to be slightly higher than in the dual media filters since during this run a constant flow from the coarse media filter effluent line was used for turbidity sampling purposes. Also, the system flow was determined from the rotameters following the dual media filters so there may have been

some slight differences between calibrations for the rotameters. In some runs, the accuracy of the coarse media filter rotameters was questionable as the run progressed and the rotameter tubes and floats become coated with solids.

Run 3

Run 3 was identical to Run 4. The results are not reported herein since operational problems during the filter run resulted in the filters being dewatered.

Runs 5 and 6

The purpose of Run 5 was to determine the effect of a higher polymer dosage on filter performance. The polymer dosage was increased from 0.05 mg/1 to 0.1 mg/1. Because of the system performance during Run 5, Run 6 was made to evaluate the effect of a decreased alum dosage while keeping the polymer dosage at 0.1 mg/1. Both filter runs were made with the coarse media filters operating in a declining rate mode. The effluent turbidity and headloss data for Run 5 are shown in Figures 17 and 18 and for Run 6 in Figures 19 through 20.

The combination of alum and polymer used in Run 5 produced a low coarse media effluent turbidity. During a 24 hour period the coarse media filter effluent turbidity rarely exceeded 2.5 NTU. The headloss in the coarse media filters was generally linear and increased at a high rate. The high rate of headloss development shortened the coarse media filter run lengths. Very little headloss developed in the dual media filters due to the relatively good influent delivered to them by the coarse media filters. The final effluent turbidity was higher during

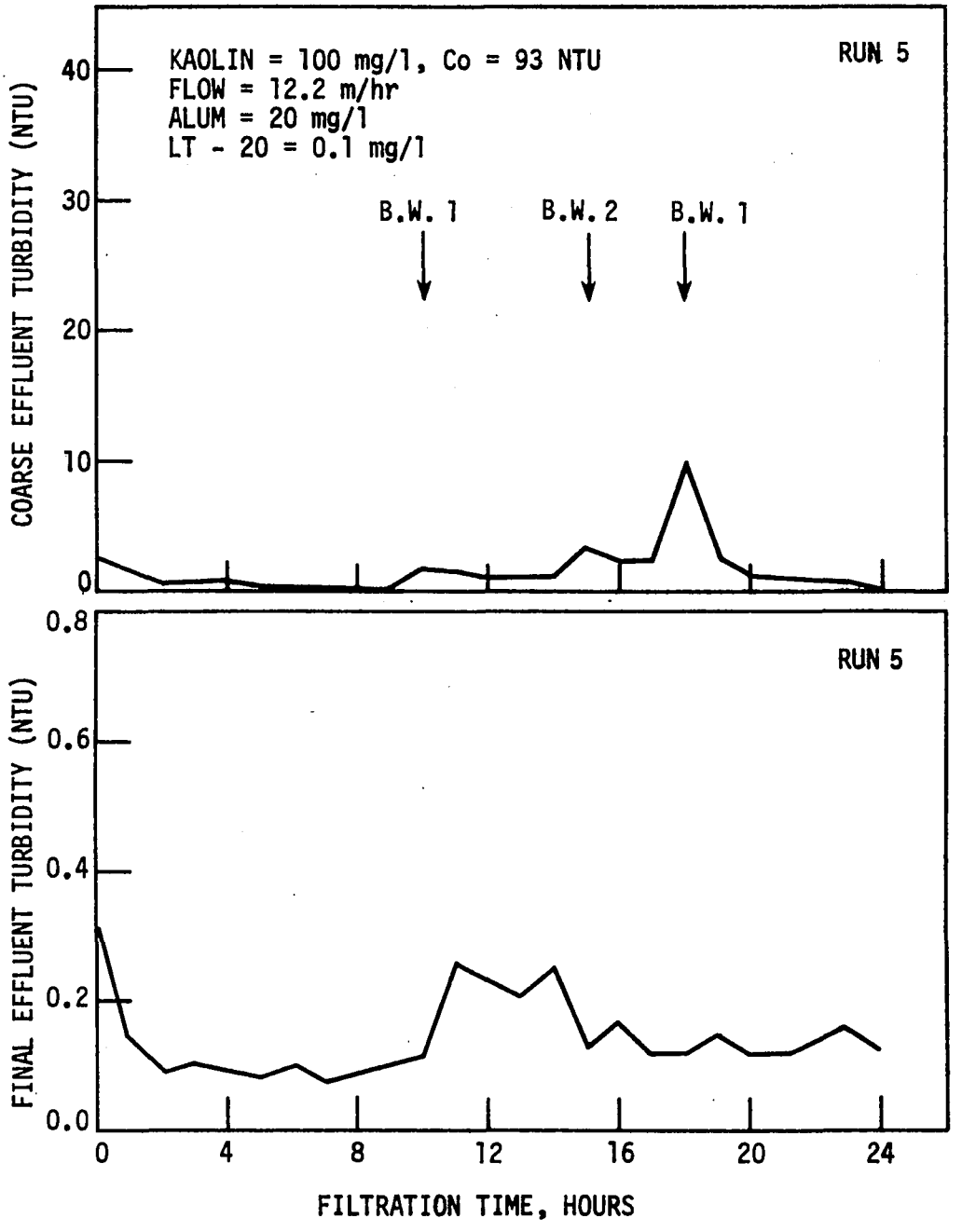


Figure 17. Effluent turbidity vs. filtration time, Run 5, coarse media filters operated in declining rate and dual media filters operated at constant rate

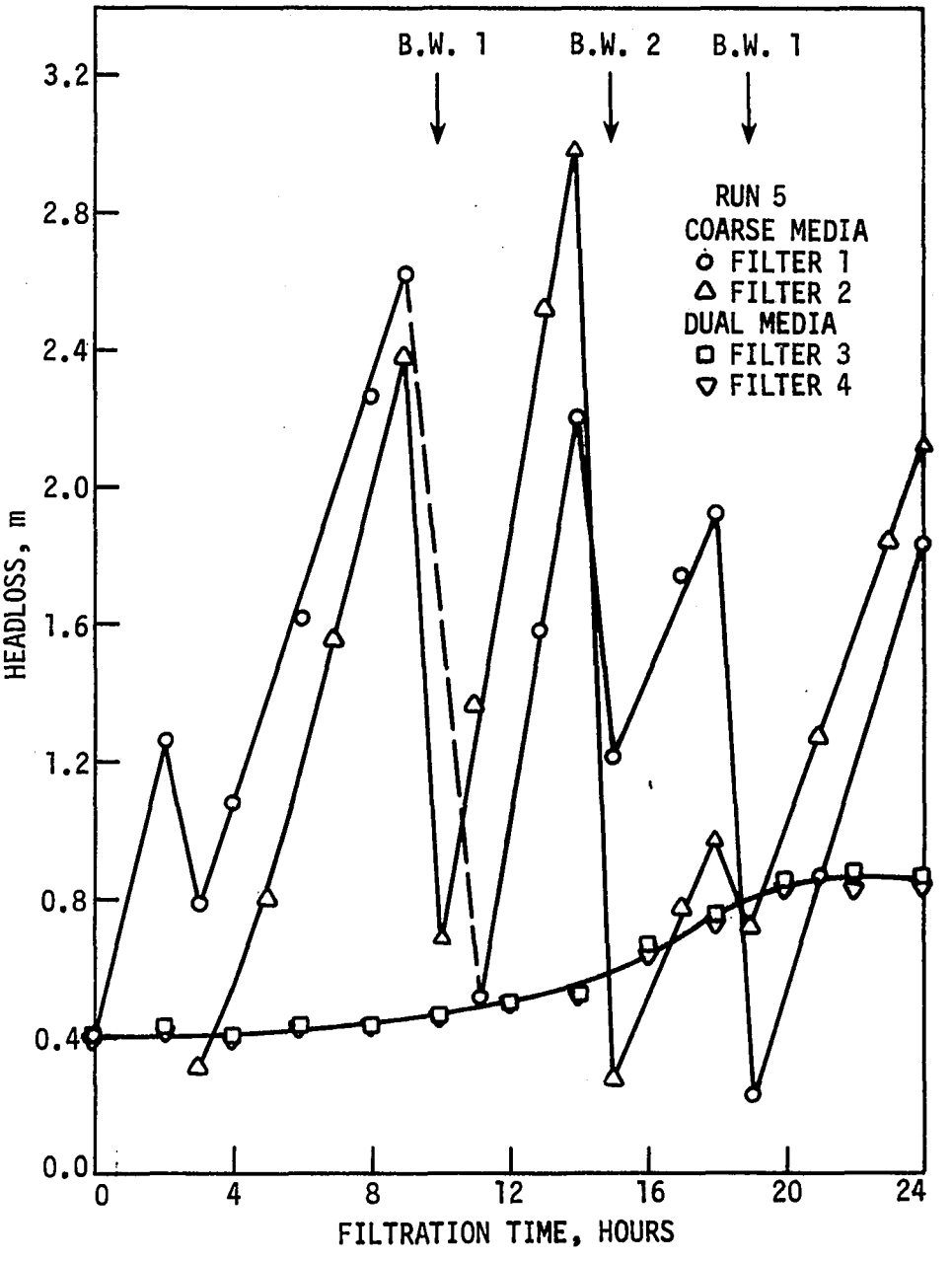


Figure 18. Headloss increase vs. filtration time, Run 5, coarse media filters operated in declining rate and dual media filters operated at constant rate

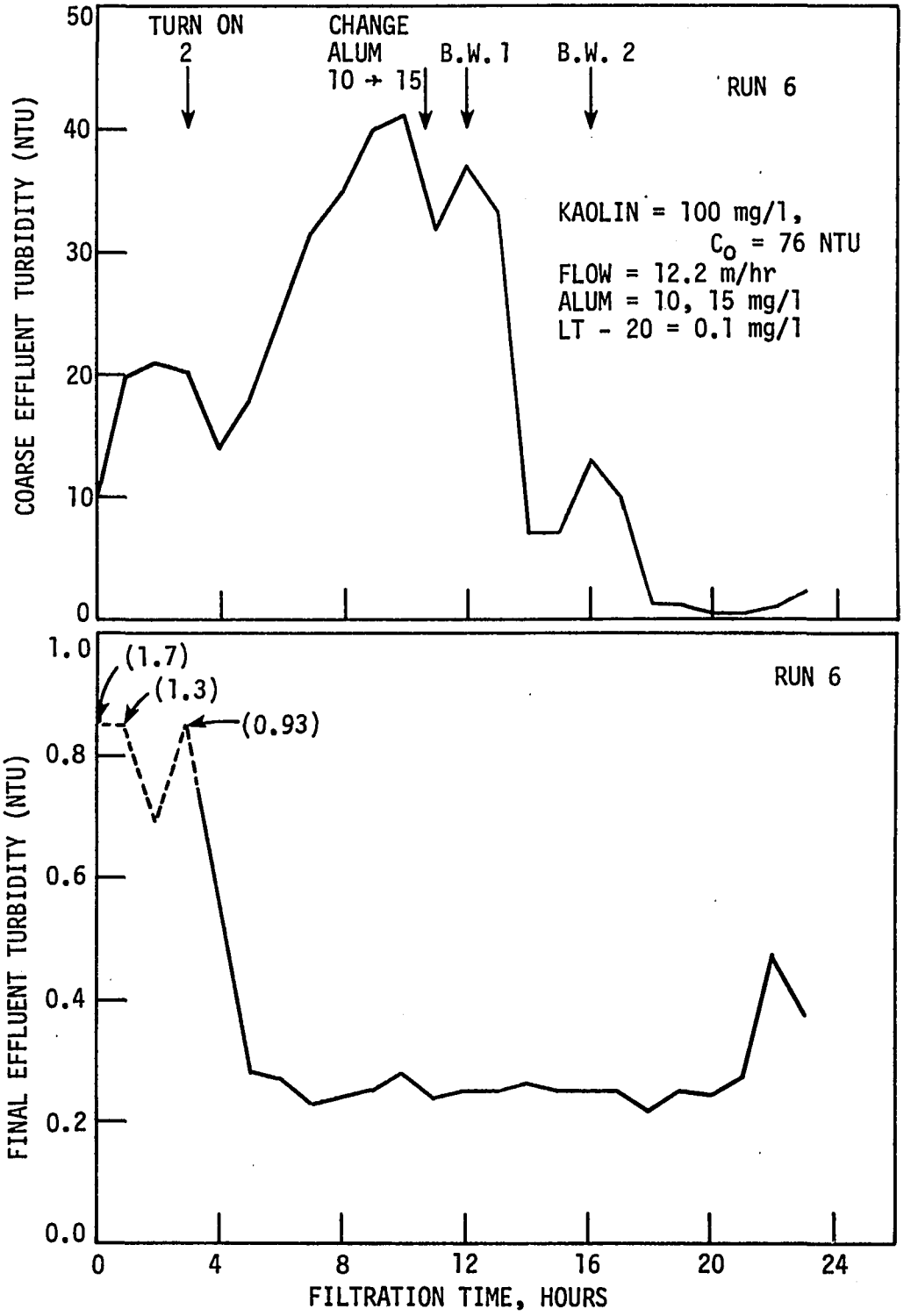


Figure 19. Effluent turbidity vs. filtration time, Run 6, coarse media filters operated in declining rate and dual media filters operated at constant rate

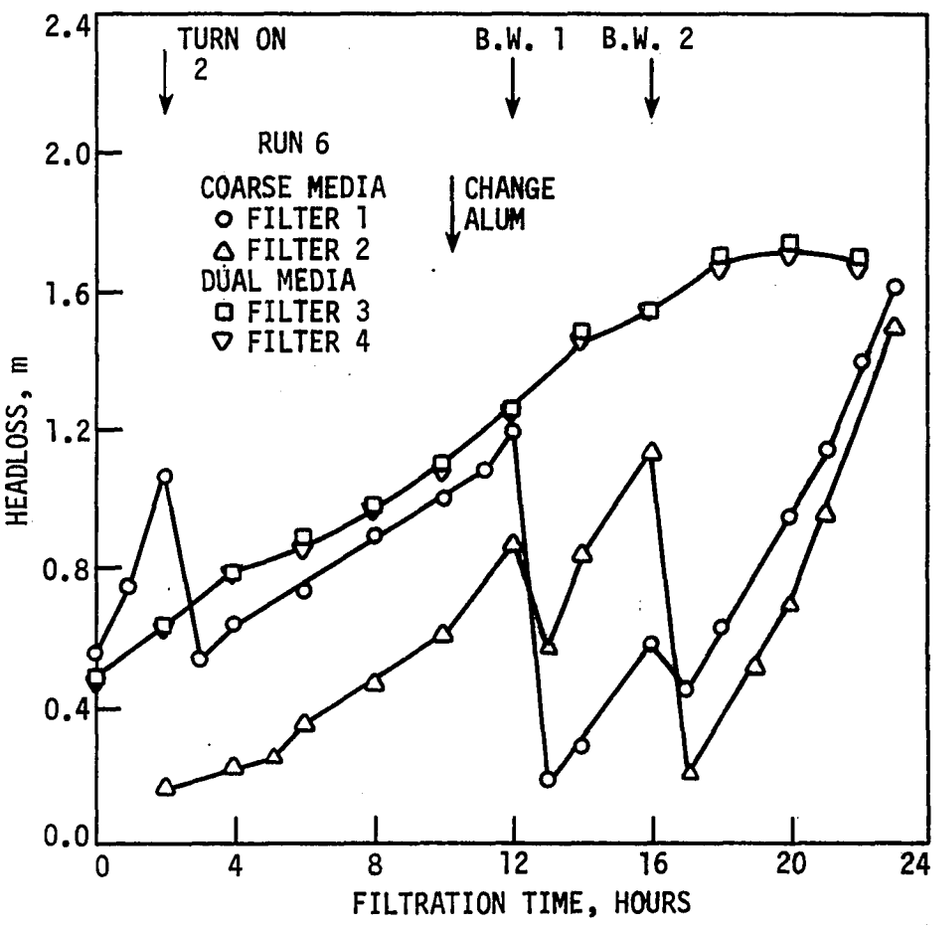


Figure 20. Headloss increase vs. filtration time, Run 6, coarse media filters operated in declining rate and dual media filters operated at constant rate

this run (< 0.2 NTU) when compared to previous runs (< 0.1 NTU). This probably was due to the low influent turbidity and subsequent slow ripening of the dual media filters.

The results from Run 6 indicated that 10 mg/l of alum in combination with 0.1 mg/l of nonionic polymer was not a sufficient alum dosage. The coarse media effluent turbidity did decrease when filter 2 was put into service but only for a short period. The dual media effluent turbidity began to decrease when the second coarse media filter was brought into service. When the alum dosage was increased from 10 to 15 mg/l, the coarse media effluent turbidity decreased the next sampling period but then increased the following period. One coarse media filter was then backwashed and the coarse media effluent turbidity dropped substantially. When the second coarse media filter was backwashed, the turbidity again dropped to levels seen in Run 5. The dual media turbidity never changed much as the run conditions varied. An improvement in final effluent turbidity might have been expected with an improvement of such an extent as that which occurred in the coarse media effluent turbidity. It should be noted that there was no apparent turbidity breakthrough in the dual media filters when the coarse media filters were backwashed. The headloss development pattern in the dual media filters, during the first 17 hours of the filter run, was indicative of depth filtration. After this time period and until the filter run was terminated little additional headloss developed in the dual media filters since the water received from the coarse media filters was of a relatively high quality.

In the coarse media filters the shape of the headloss curve may be attributed to changes in alum dosage and flowrate. The effect of the alum dosage on the coarse media headloss is evident by the higher rates of headloss after the alum was increased.

Run 7

The results of Run 7 are not reported herein for two reasons. First, the polymer NP-10 used appeared to be inferior to the other nonionic polymer. The second reason was an operational problem during the run.

Run 8

The purpose of Run 8 was to observe the filter system performance when no water pretreatment chemicals were used. The results are shown in Figures 21 and 22. As would be expected, both the coarse media and dual media effluent turbidities were high. The dual media effluent turbidity showed the most variation and overall gradually increased with time. Headlosses in both the dual media and coarse media filters was very low, indicating little solids capture.

Run 9

The purpose of Run 9 was to repeat the conditions of Run 4 to determine the effect, if any, of a new supply of the kaolin. The results showed the same patterns as demonstrated in Run 4. The new kaolin supply had no discernible effects on the filter system performance after allowing for slight differences that may have resulted from operational variances.

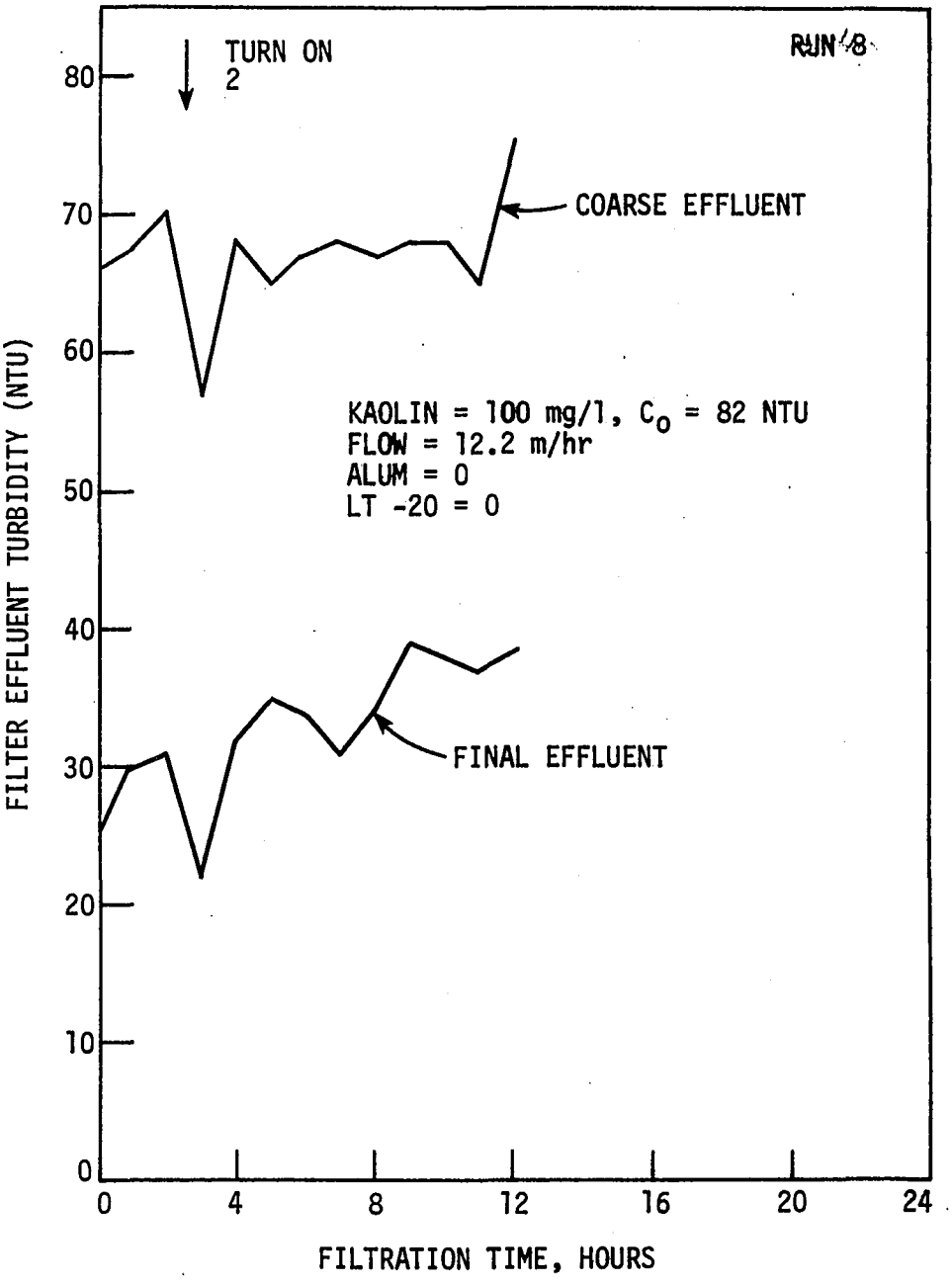


Figure 21. Effluent turbidity vs. filtration time, Run 8, coarse media filters operated in declining rate and dual media filters operated at constant rate

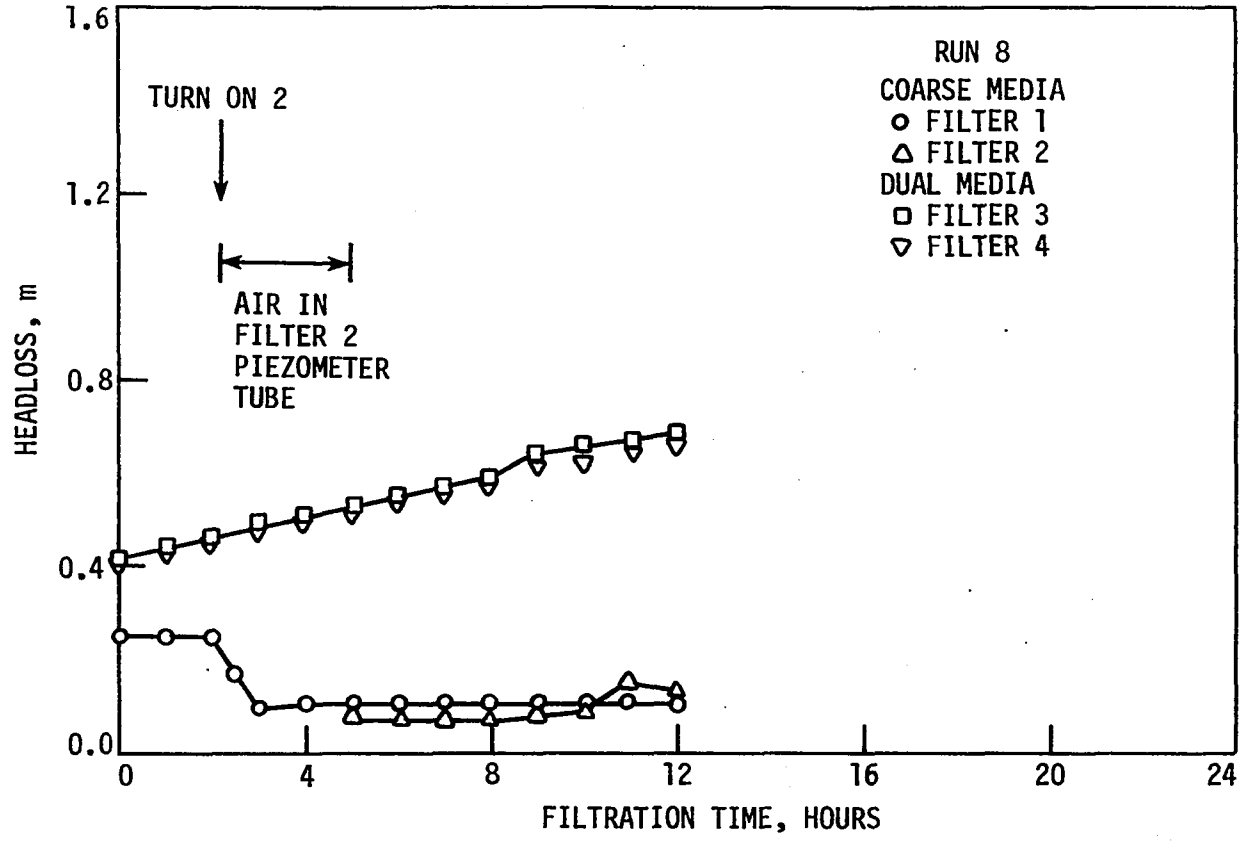


Figure 22. Headloss increase vs. filtration time, Run 8, coarse media filters operated in declining rate and dual media filters operated at constant rate

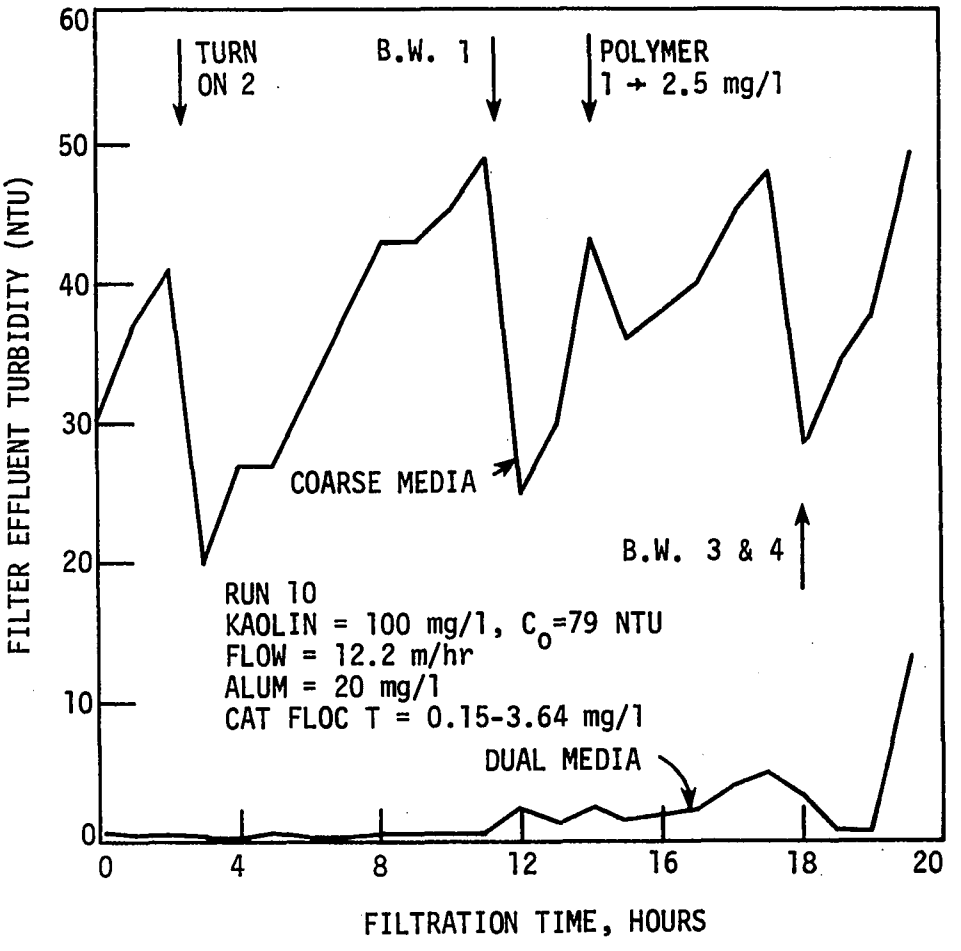


Figure 23. Effluent turbidity vs. filtration time, Run 10, coarse media filters operated in declining rate and dual media filters operated at constant rate

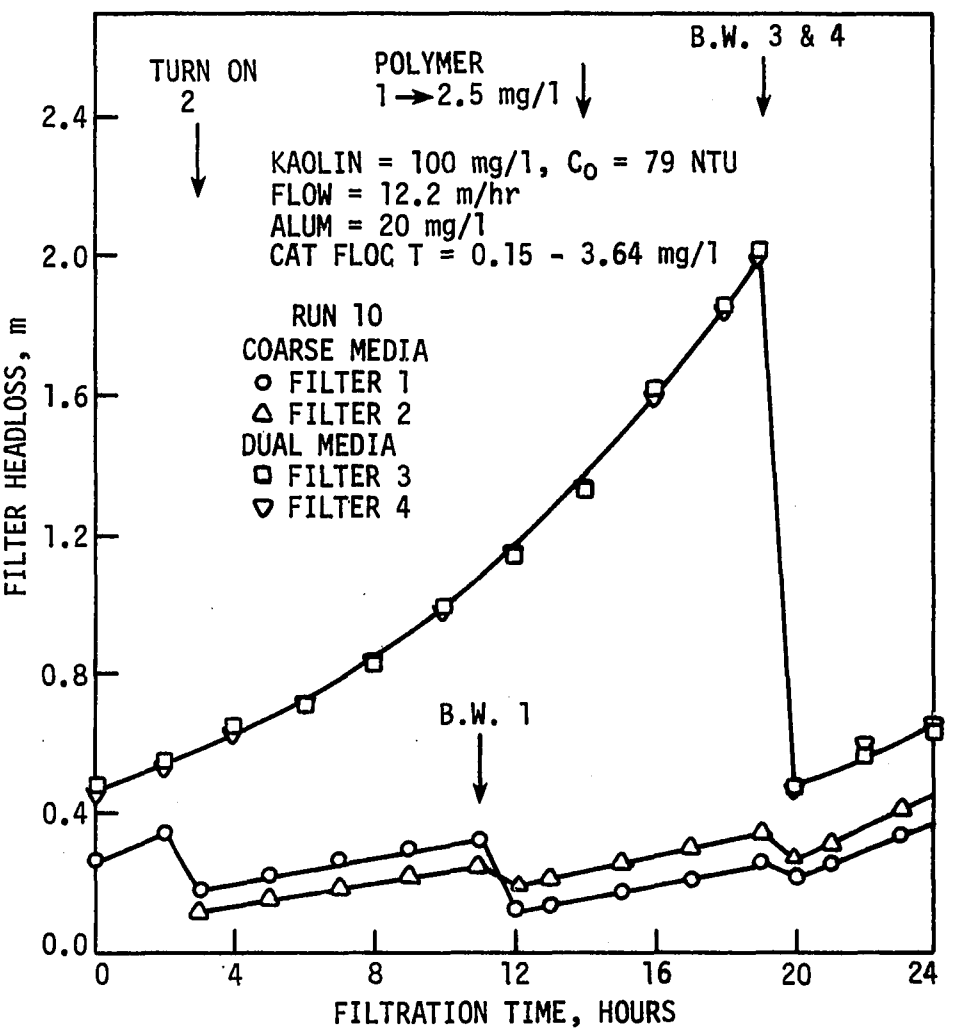


Figure 24. Headloss increase vs. filtration time, Run 10, coarse media filters operated in declining rate and dual media filters operated at constant rate

Run 10

Run 10 was undertaken to evaluate the effects on filter performance of a combination of alum and Cat Floc T, a cationic polymer. The results shown in Figure 23 indicate that the coarse media effluent turbidity remained high throughout the run. As shown in Figure 24, very little headloss developed in the coarse media. The effluent turbidity from the dual media filters remained below 1.0 NTU for the first 11 hours of the run. During the next 8 hours the final turbidity increased from 0.7 NTU to about 4.8 NTU. At this time the dual media filters were backwashed and the final effluent turbidity dropped below 1.0 NTU within 2 hours. Unfortunately a soap suds problem appeared resulting in a sharp increase in final effluent turbidity and therefore the run was terminated. Terminal headloss in the dual media filters was reached after 18 hours. The characteristics of the headloss curve were indicative of filters receiving a variable influent suspended solids concentration.

Run 11

The purpose of Run 11 was to confirm the data obtained in Run 6. A number of operational problems occurred during the latter stages of this run. Run 12 was then made to confirm Run 6 results.

Runs 12 and 13

Runs 12 and 13 were made under identical conditions with the exception of the operational mode. Run 12 was made with the coarse media filters operated in a declining rate and Run 13 was made with the filters operated at a constant rate. The results of Run 12 are shown in Figures

25 through 27 and the results of Run 13 are shown in Figures 28 and 29.

A comparison of the coarse media effluent turbidities during Runs 12 and 13 shows the same trends as found during earlier filter runs. When the filters are operated at a constant rate, the coarse media effluent turbidity remains low for several hours and then rapidly deteriorates to the terminal value. In the declining rate mode, the coarse media effluent turbidity does not reach extremely low values and stay low for several hours. Usually, with the declining rate mode a minimum turbidity lasts only 1 or 2 hours and then begins to rise. The final effluent turbidity data shows very good results for Run 13. The final effluent turbidity in Run 13 exceeded 0.1 NTU on only two occasions, at time 0 and when terminal headloss was reached. The final effluent turbidity in Run 12 showed an uncharacteristically long ripening period at the beginning of the run. Most final turbidities were between 0.1 and 0.2 NTU. The dual media filters were backwashed in Run 12 due to headloss limitations after only 20 hours of operation. In Run 13, the dual media filters were operated for 28 hours before headloss limitations necessitated backwashing.

The coarse media headloss data for both Runs 12 and 13 show an almost linear headloss development. Higher terminal headlosses were possible in Run 13 at constant rate. The dual media headloss curves for both Run 12 and Run 13 showed some curvature due to the variable dual media influent quality. The rate of headloss development in the dual media filters was greatest in Run 12. In both runs, increases in headloss were associated with backwashing the coarse media filters. In

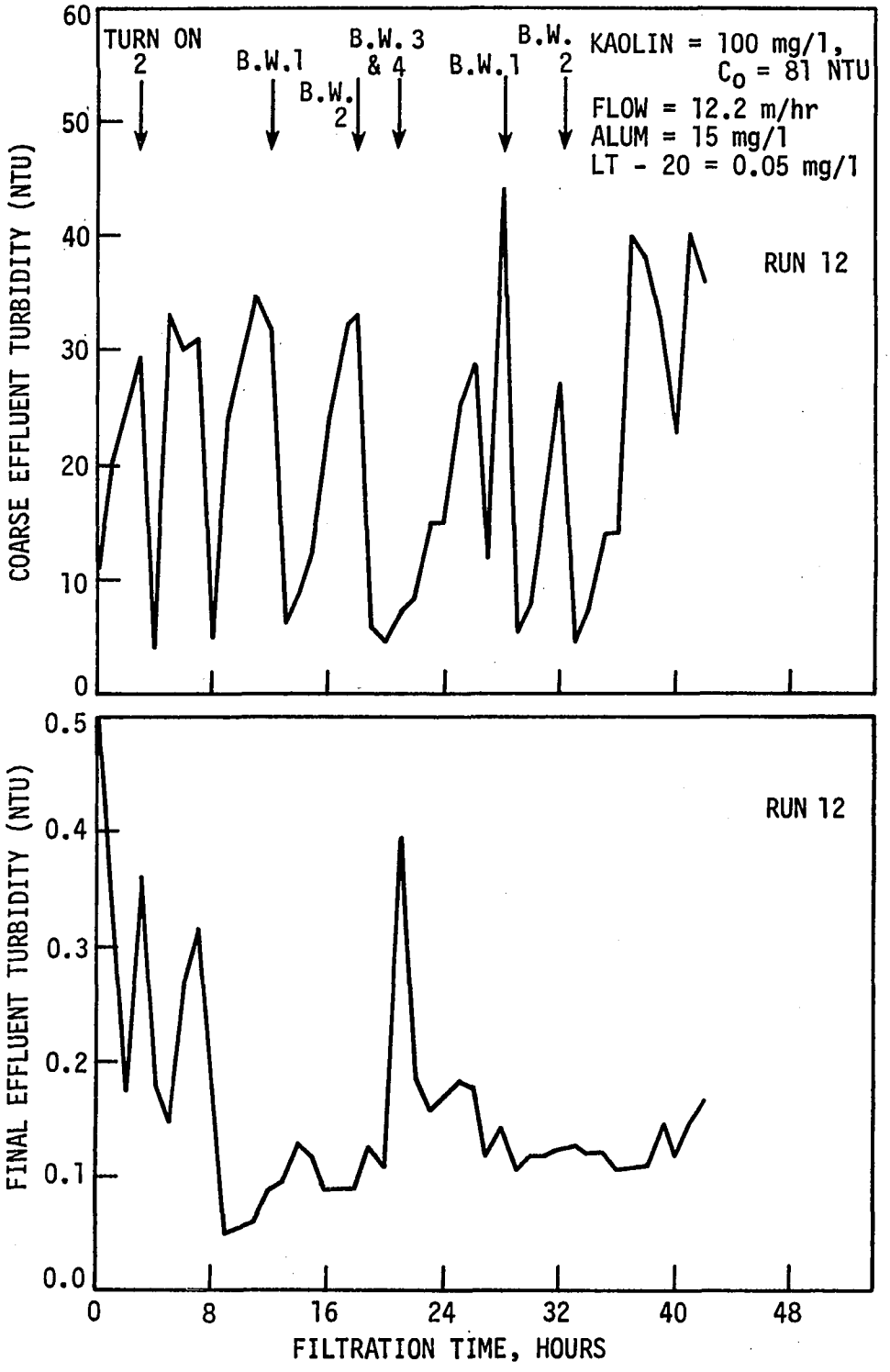


Figure 25. Effluent turbidity vs. filtration time, Run 12, coarse media filters operated in declining rate and dual media filters operated at constant rate

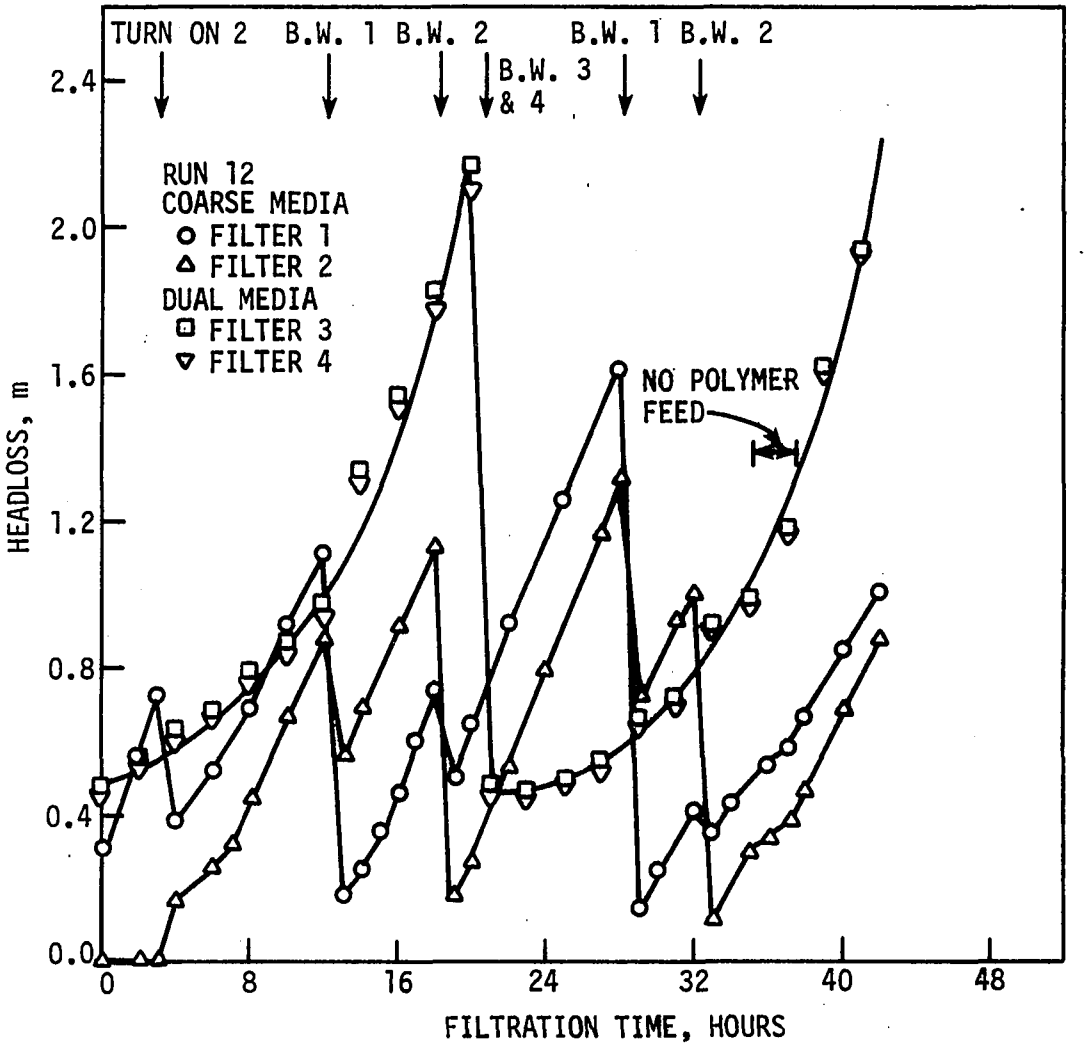


Figure 26. Headloss increase vs. filtration time, Run 12, coarse media filters operated in declining rate and dual media filters operated at constant rate

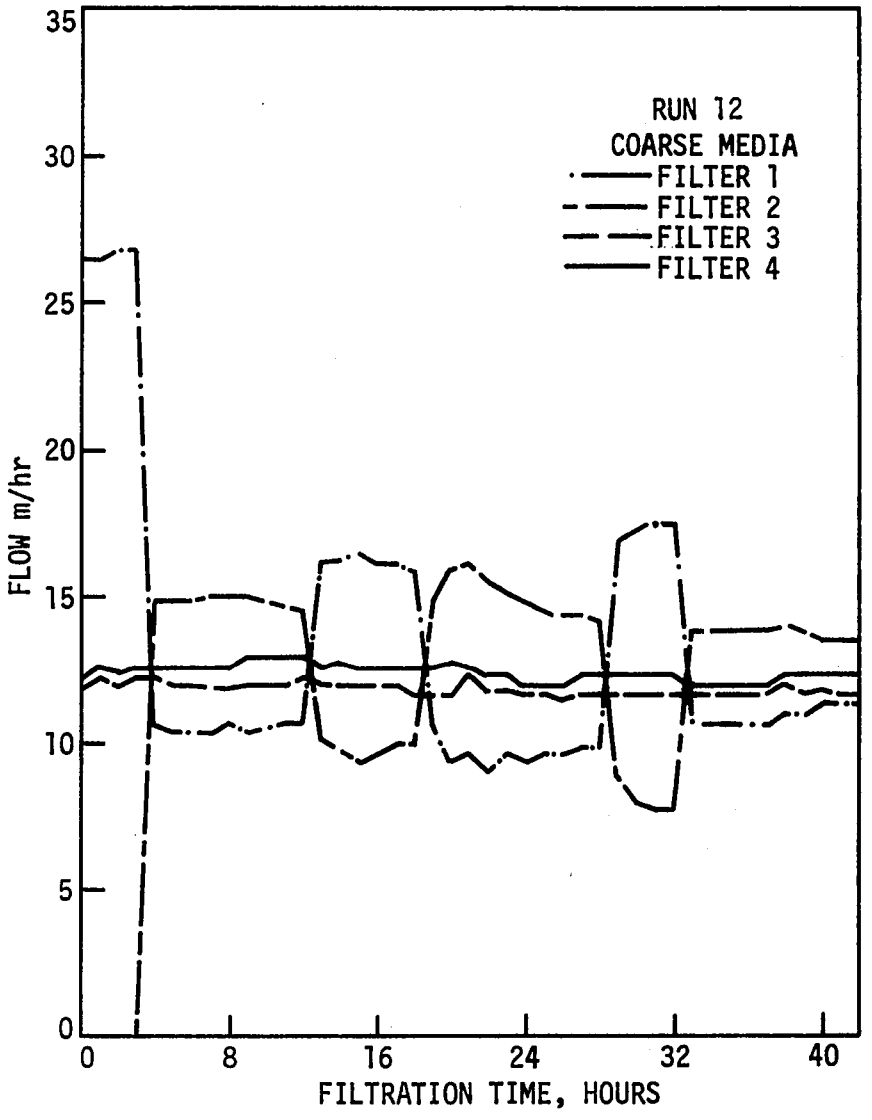


Figure 27. Filter flow variations, Run 12, coarse media filters operated in declining rate and dual media filters operated at constant rate

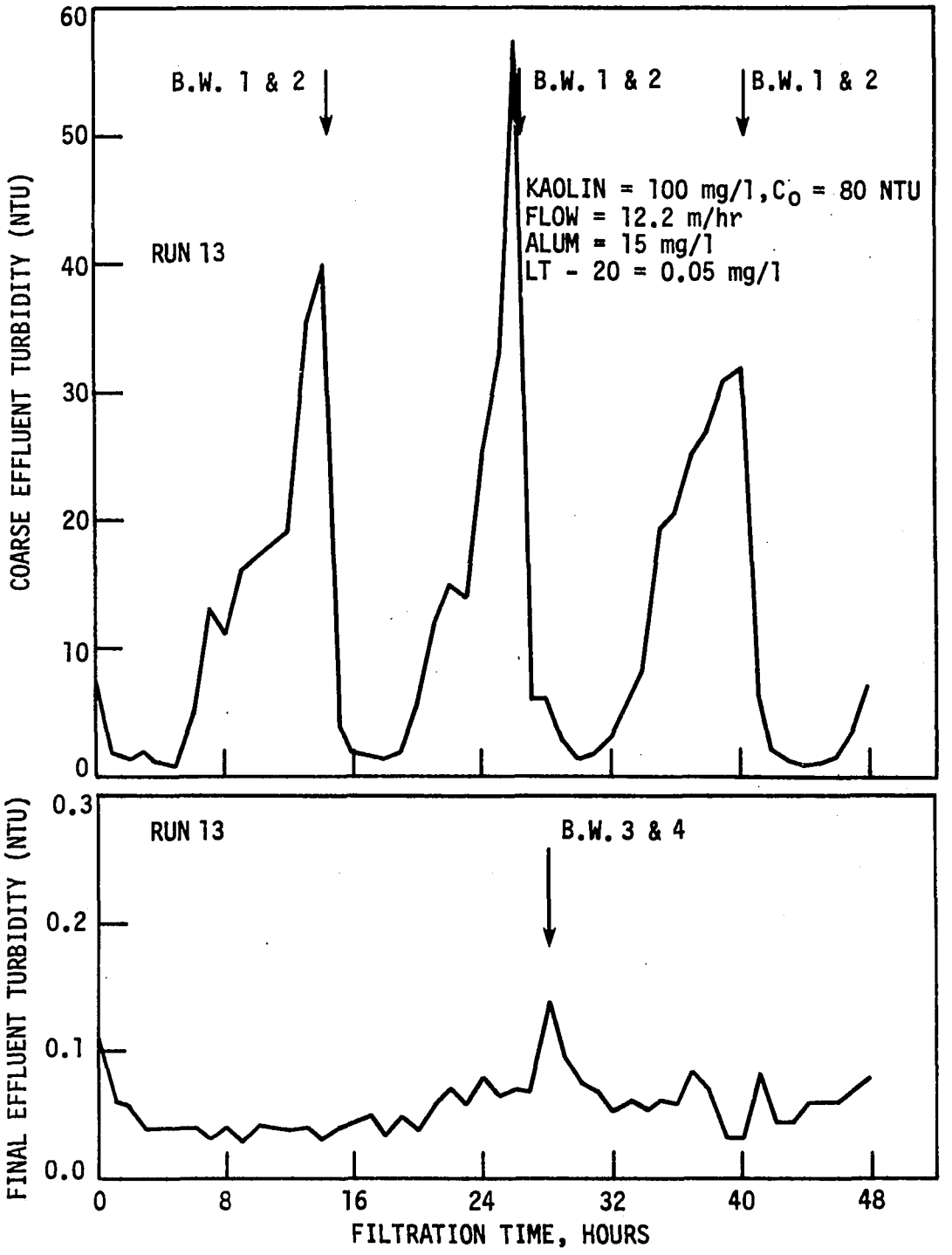


Figure 28. Effluent turbidity vs. filtration time, Run 13, coarse media and dual media filters operated at constant rate

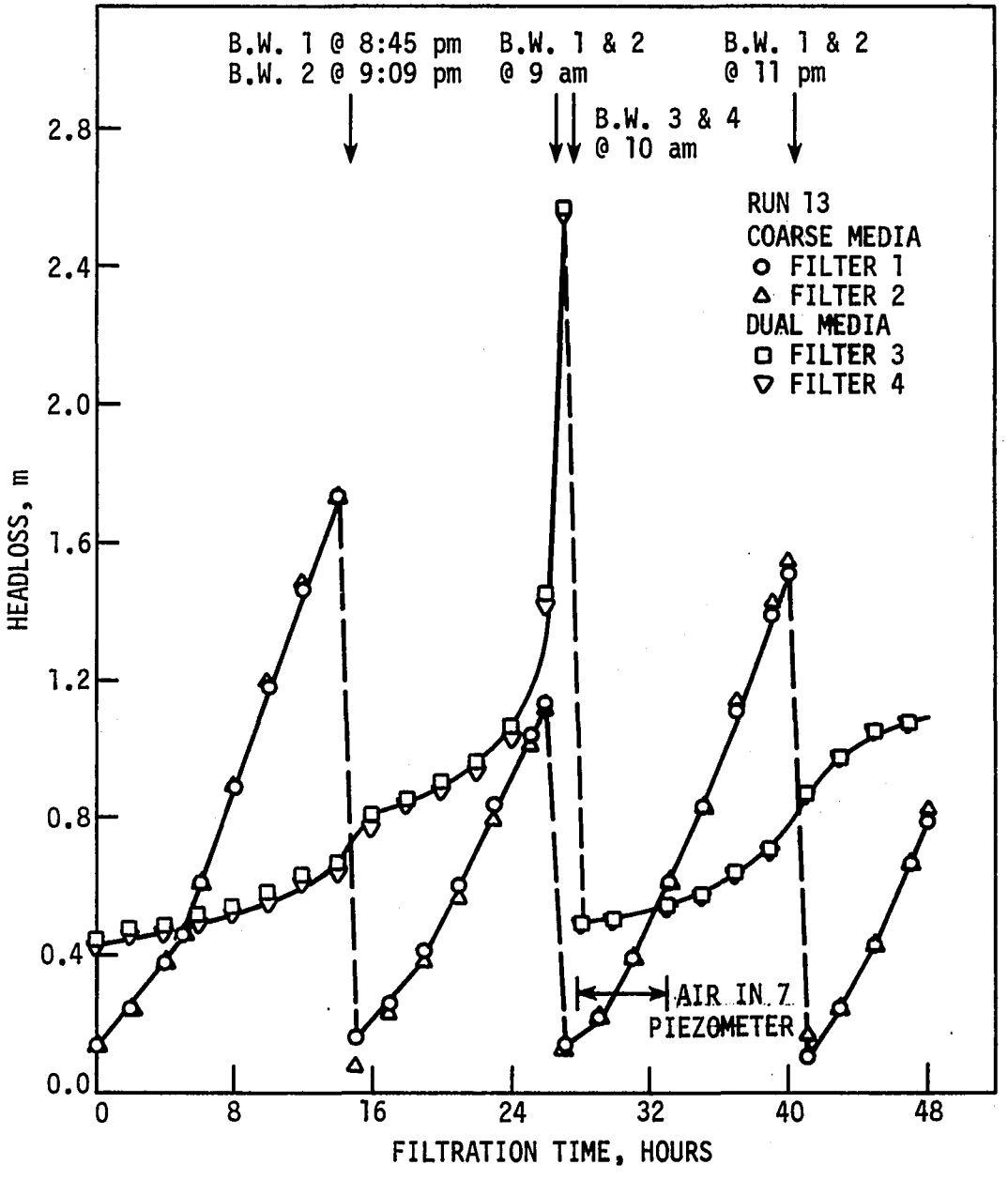


Figure 29. Headloss increase vs. filtration time, Run 13, coarse media and dual media filters operated at constant rate

Run 13 at constant rate the influence of the coarse media backwash was much more evident. In a one hour period in Run 13 the dual media headloss increased from 1.4 m to 2.5 m resulting in a need to backwash the filters.

The variation in flowrates in each coarse media filter during Run 12 as a result of backwashing filters is shown in Figure 27. Run 13 was made at constant rate.

Run 14

The purpose of Run 14 was to evaluate the filter system when 300 mg/l of kaolin was used for the raw water suspension. The results were very poor. Terminal headloss in the dual media filters was reached in 10 hours while the final filtered water turbidity ranged from about 1.0 NTU to over 100 NTU. The headloss in the coarse media filters was very low and effluent turbidity varied radically. The alum and polymer dosages (Table 9) were varied to try and improve the filter system performance but all attempts were unsuccessful.

Run 15

In Run 6, a combination of 10 mg/l of alum and 0.1 mg/l of polymer had proven to be less effective than a combination of 15 mg/l of alum and 0.1 mg/l of polymer. Run 6 was performed in the declining rate mode. As a result of the good results experienced in Run 13, Run 15 was conducted to determine whether 10 mg/l of alum and 0.05 mg/l of polymer would be successful when the filters were operated at constant rate.

The coarse media effluent turbidity data showed a trend in between constant rate and declining rate results in previous runs. The typical

constant rate result of low turbidity was observed but the period in which low turbidity was observed was shortened, as in typical declining rate results. The turbidity was excellent throughout most of the run, although a higher than normal initial turbidity was measured. The coarse media headloss was relatively low in all filter cycles. Some slight exponential behavior was noticeable. The dual media headloss showed exponential behavior and occurred at a high rate. The dual media filters were backwashed after 22 hours due to headloss limitations. The significance of the coarse media breakthrough during backwash was not as evident in this run as in Run 13.

Run 16

Run 16 was made to determine the effects of the polymer, Percol LT-20 on the filtration system performance. Percol LT-20, a nonionic polymer, was used at a concentration of 0.05 mg/l. As would be expected, the headloss in the coarse media filters was low, increasing by only 0.2 m in 11 hours. The headloss in the dual media filters increased by 0.6 m in the same time period. The turbidity results were somewhat of a surprise. The effluent turbidity from the coarse media filters increased from about 20 NTU to 38 NTU in a 10 hour period. The final effluent turbidity ranged from about 0.3 NTU to 5.4 NTU during the same time period. For the first 8 hours of the run, the final turbidity was below 1.0 NTU.

Runs 17 and 19

Runs 17 and 19 involved the use of calcium aluminosilicate. Run 17 used 100 mg/l of calcium aluminosilicate as the sole raw water particle

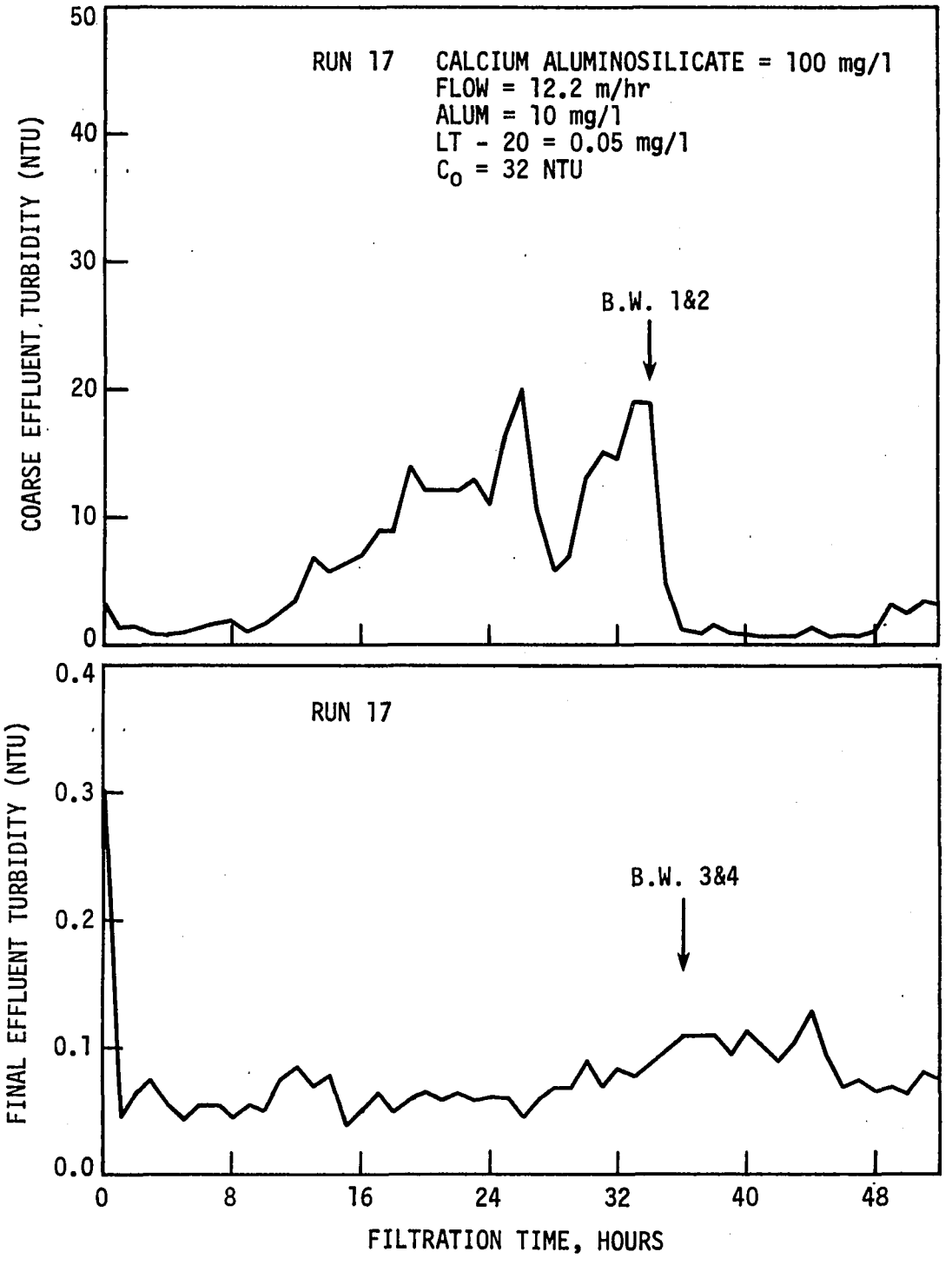


Figure 30. Effluent turbidity vs. filtration time, Run 17, coarse media and dual media filters operated at constant rate

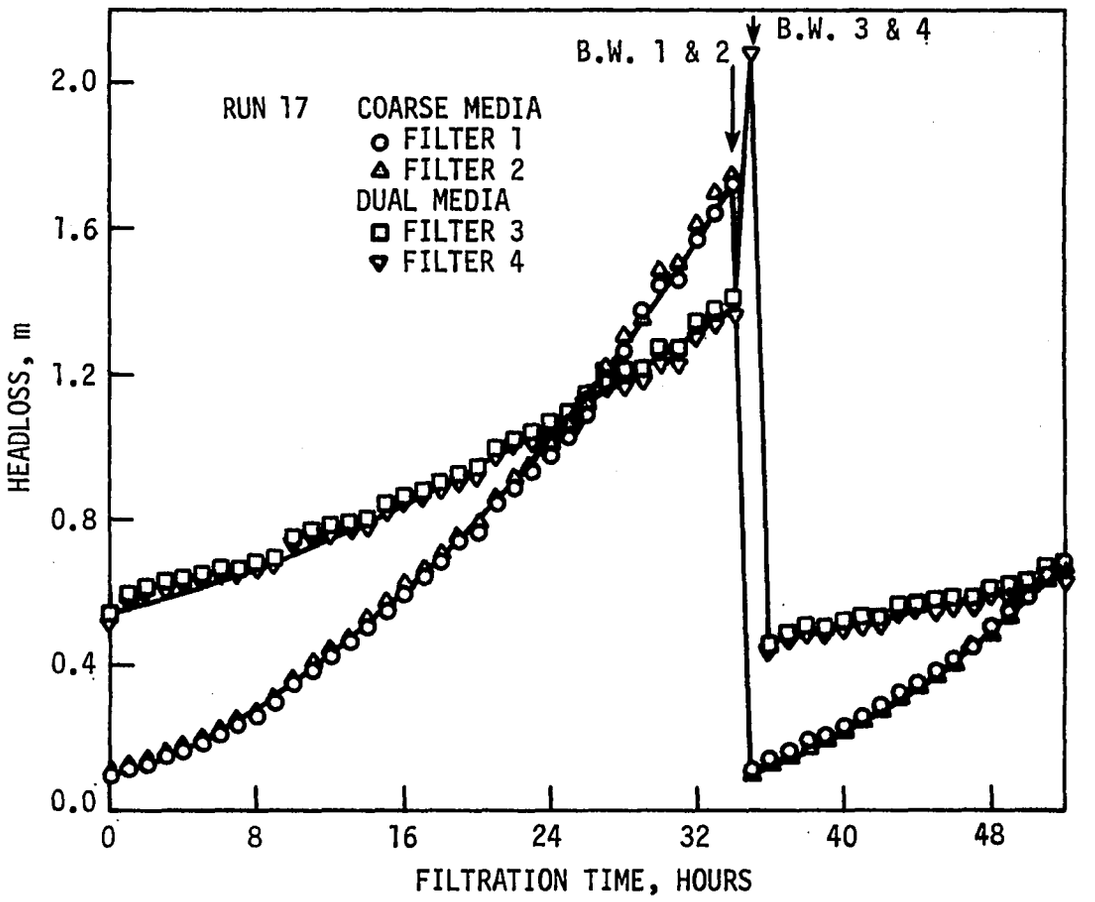


Figure 31. Headloss increase vs. filtration time, Run 17, coarse media and dual media filters operated at constant rate

while Run 19 used calcium aluminosilicate and kaolin at 100 mg/l each. The turbidity created by 100 mg/l of calcium aluminosilicate was about half that when 100 mg/l of kaolin was used. Both filter runs were made at a constant rate. The alum dosage of 10 mg/l used in Run 17 was the optimum determined in the jar tests. The results for Run 17 are shown in Figures 30 through 31. The coarse media effluent turbidity results show that low turbidities were obtained and turbidity breakthrough was more gradual than when kaolin was used in previous runs. The final effluent turbidity remained near 0.1 NTU or less throughout the run duration. An extended ripening period was noticed in the dual media filters.

The coarse media headloss exhibited exponential behavior as did the dual media headloss to some extent. Most of the particles were removed in the coarse media filters as evidenced by the total headloss after 33 hours. About 1.7 m of headloss developed in the coarse media during this time while about 0.80 m developed in the dual media. When the coarse media filters were backwashed, the dual media headloss increased from 1.4 m to 3.0 m which necessitated dual media backwash.

The results of Run 19, not presented, were similar to the results of Run 13 since both filter runs were operated at constant rate and in both filter runs 100 mg/l of kaolin was used in the raw water supply. The only difference between the runs was the addition of 100 mg/l of calcium aluminosilicate in the raw water supply in Run 19. From the results of Run 19, it appeared that the presence of the additional particles had

little effect on the dual media effluent turbidity. The coarse media effluent turbidity was affected in respect to the duration of the period of low coarse media effluent turbidity between the backwashing of the coarse media filters and the onset of effluent turbidity breakthrough. Although the coarse media filter runs were shorter in Run 19, the rapidity of breakthrough seemed to be less severe. The overall rate of headloss buildup was slightly greater in Run 19 than in Run 13 which is indicative of the higher influent solids concentration. The rate of headloss buildup in the dual media filters in Run 19 was greater than in Run 13. The same trend with respect to high dual media headloss resulting from coarse media backwash was present in Run 19. The dual media filter runs in Run 19 lasted about 23 hours whereas in Run 13, the dual media run length was 27 hours.

Run 18

The purpose of Run 18 was to evaluate the filter system performance when filtering water with a kaolin concentration of 200 mg/l. Twenty milligrams per liter of alum was used in combination with 0.05 mg/l of Percol LT-20. The results of this filter run showed the alum and polymer combination to be ineffective in producing acceptable results. Coarse media filter runs lasted only about 4 hours before the turbidity exceeded 60 NTU. Dual media filter runs lasted about 6 hours before the effluent turbidity exceeded 1.0 NTU. Coarse media headloss was low while the dual media headloss built up very rapidly, as much as 0.1 m per hour. Near the end of the filter run, the flow rate was decreased to 6.1 m/hr

and all the filters were backwashed. This reduced the coarse media effluent turbidity to less than 5 NTU but within 6 hours the coarse media effluent turbidity had reached 60 NTU. The dual media effluent turbidity remained at about 0.1 NTU through this period.

Runs 20, 21 and 25A

The purpose of these filter runs was to evaluate the filter system performance with the same chemical pretreatment but with different influent clay concentrations. Run 20 was the first filter run using less than 100 mg/l of kaolin so this run also includes data for various alum dosages. The results for Run 20 are given in Figures 32 through 35.

The results from Run 20 indicated that 7.5 mg/l of alum in combination with 0.5 mg/l of LT-20 produced the most desireable results. Fifteen and ten milligrams per liter of alum in combination with the polymer produced coarse media effluent turbidities often below 1.0 NTU. Reducing the alum dosage to 5 mg/l resulted in elevated turbidity values. The use of 7.5 mg/l of alum provided results which were similar to previous runs. The turbidity reached a minimum level for a period of several hours and then proceeded to breakthrough. This alum dosage was selected as the optimum for this reason. It was not the purpose of the coarse media, in this application, to remove the solids to a level which would meet present water quality standards of 1.0 NTU. The purpose of the coarse media filters was to relieve only a portion of the solids removal burden from the dual media. The dual media filter effluent turbidity remained below 0.1 NTU for most of the filter run. Exceptions

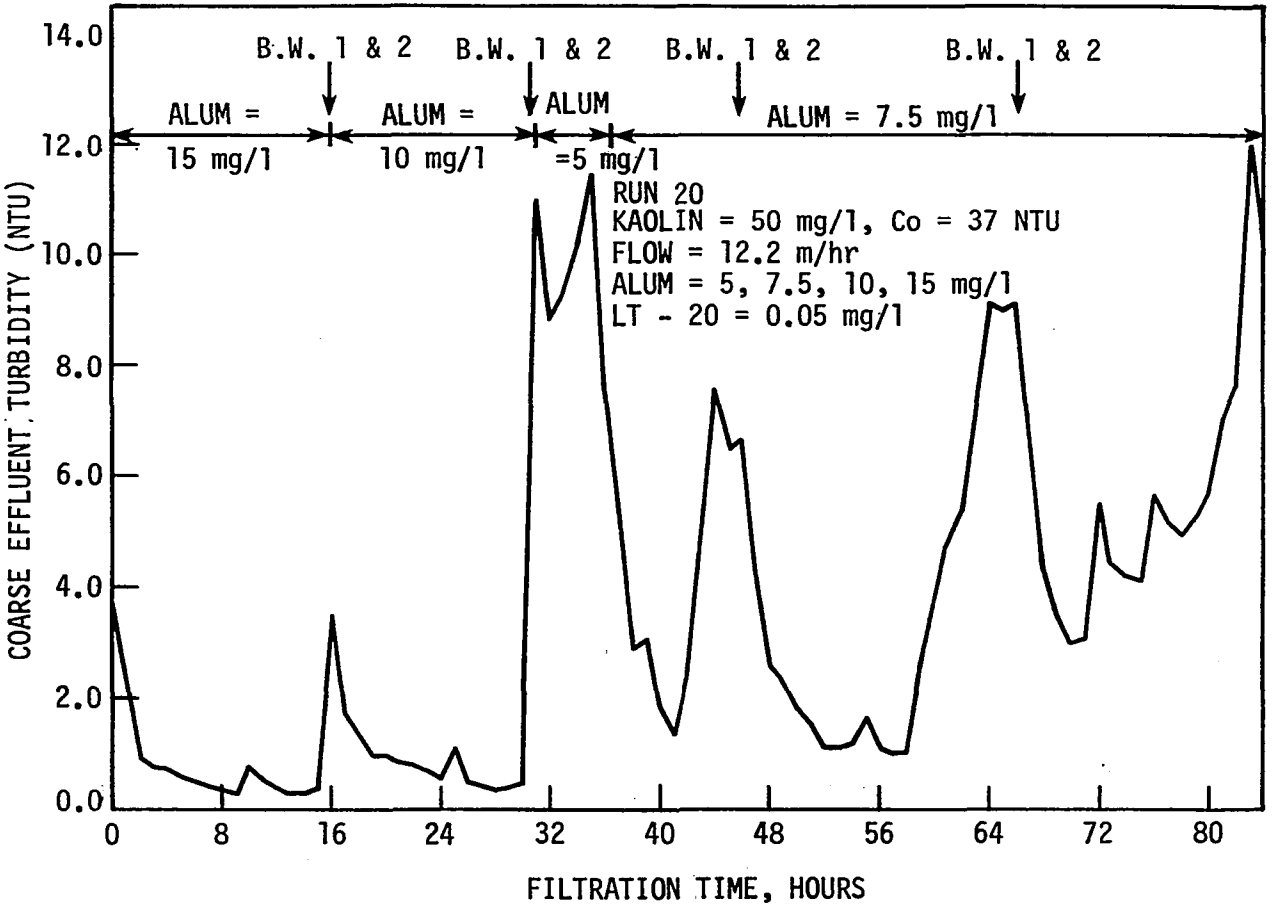


Figure 32. Coarse media effluent turbidity vs. filtration time, Run 20, coarse media filters operated at constant rate

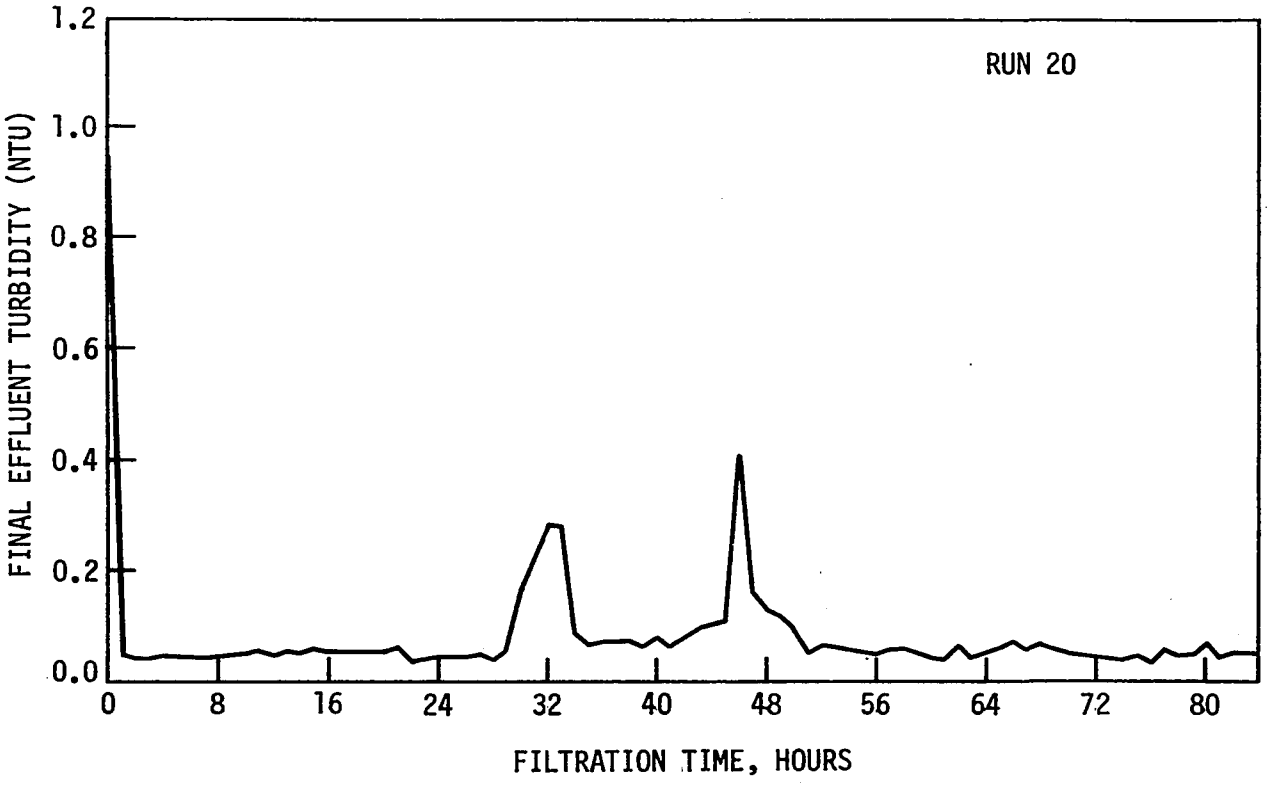


Figure 33. Dual media effluent turbidity vs. filtration time, Run 20, dual media filters operated at constant rate

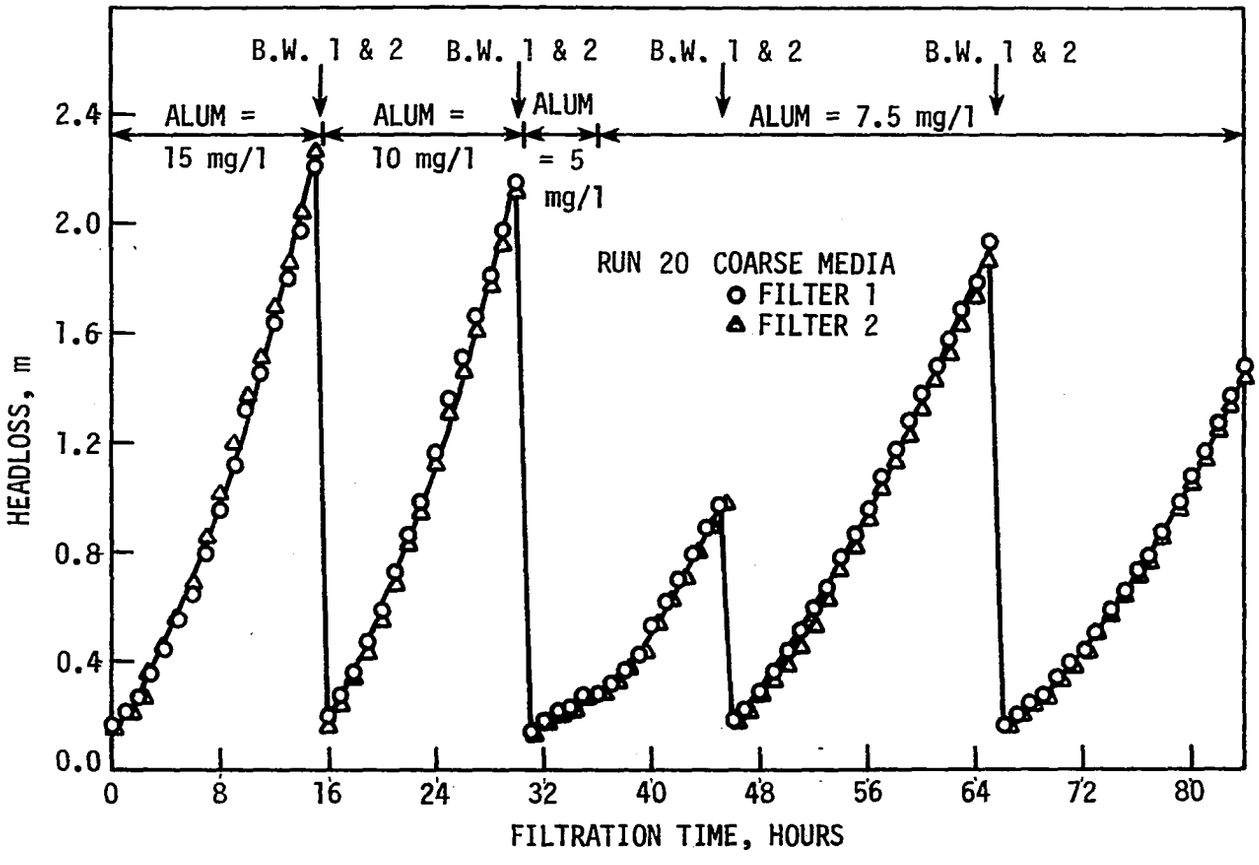


Figure 34. Coarse media headloss increase vs. filtration time, Run 20, coarse media filters operated at constant rate

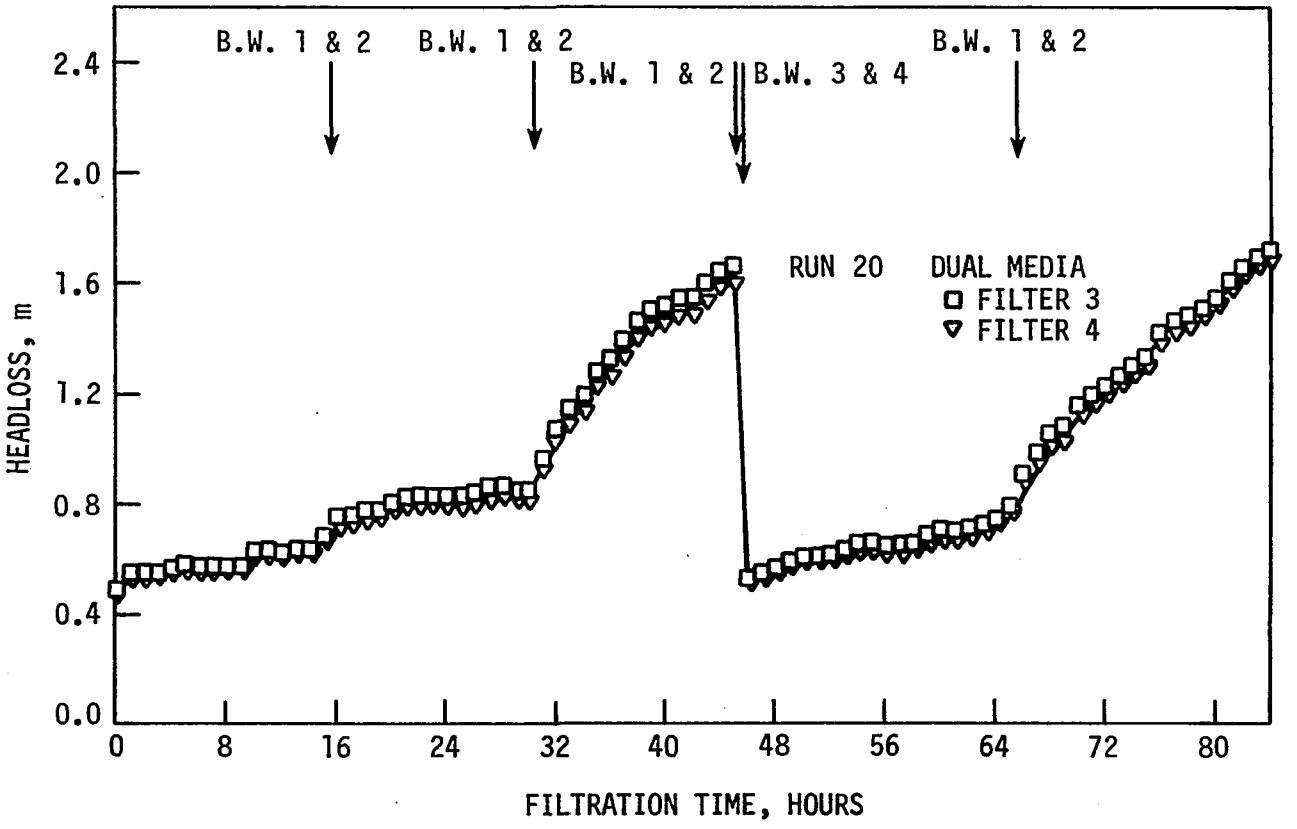


Figure 35. Dual media headloss increase vs. filtration time, Run 20, dual media filters operated at constant rate

to this were during the initial filter ripening period, during the period when coarse media filter turbidity breakthrough occurred due to an insufficient alum dosage and immediately after the coarse media filter backwash. The coarse media filter headloss curves show that a significant amount of headloss developed in the coarse media filters. At an alum dosage of 7.5 mg/1, turbidity breakthrough occurred at close to terminal headloss. The headloss development in the dual media was low (0.3 m) through the first 30 hours of operation. After a coarse media filter backwash and high coarse media filter effluent turbidities, resulting from an insufficient alum dosage (5 mg/1), the headloss increased an additional 0.8 m in a 16 hour period. After backwashing the dual media headloss buildup again remained low (0.3 m in 20 hours), but increased after the coarse media filters were again backwashed and continued to increase at a constant rate until the filter run was terminated.

The results of Run 21 are shown in Figures 36 through 37. The coarse media effluent turbidity in Run 21 exhibited a pattern typical of a standard depth filter. A ripening period was evident in the first stage of filtration. This was followed by a period of minimum turbidity change. The filters were backwashed due to limitations in available headloss for the system. Headloss in excess of about 2.0 m in the coarse media made it difficult to maintain the proper flow rate. Final effluent turbidity remained below 0.1 NTU for all but 3 out of 68 sample periods. Headloss development in the dual media progressed at a

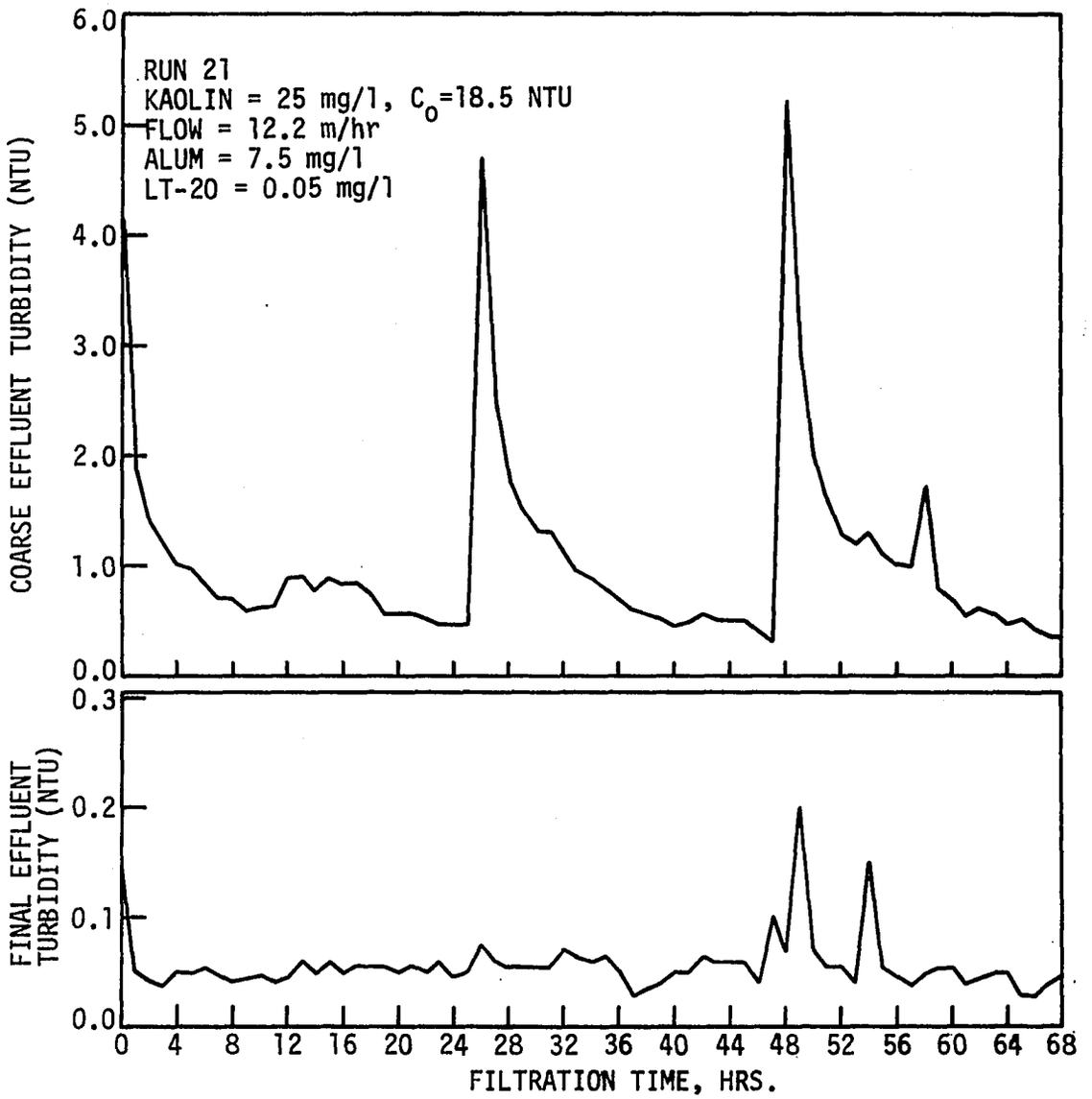


Figure 36. Effluent turbidity vs. filtration time, Run 21, coarse media and dual media filters operated at constant rate

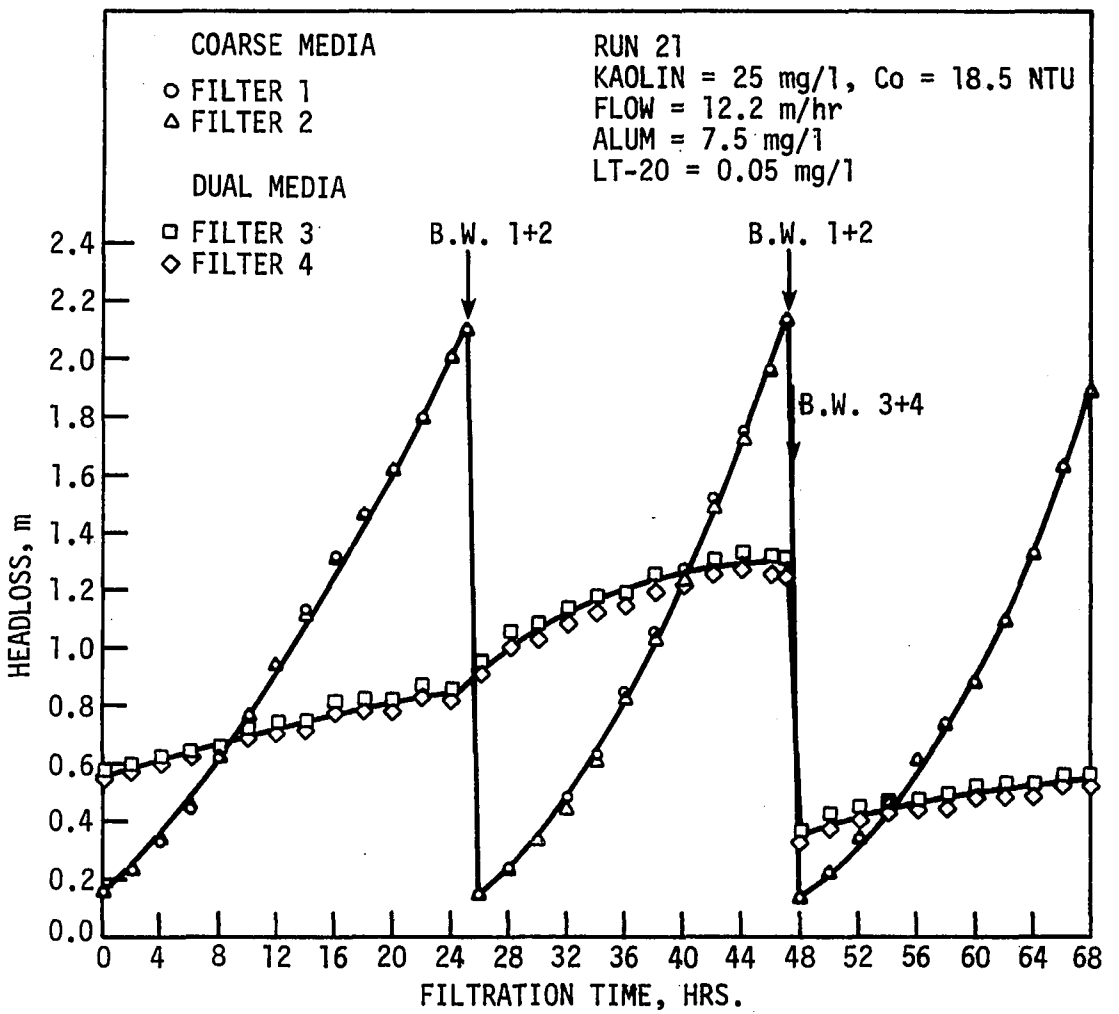


Figure 37. Headloss vs. filtration time, Run 21, coarse media and dual media filters operated at constant rate

low rate. After 47 hours of operation, the headloss had only increased by about 0.7 m. At this time the filters were backwashed to avoid any problems that might have resulted from extending the run length for a longer period of time.

Run 25A was made to provide data for the evaluation of a model used to predict filter performance. The data for Run 25A, shown in Figures 38 through 39, indicate that both coarse media and dual media filter run lengths would be shortened by the chemical pretreatment used in this run compared to earlier runs using larger alum dosages. The coarse media effluent turbidity was lowest at the beginning of the run and increased almost constantly to over 40 NTU after 10 hours. The headloss in the coarse media was linear. The dual media headloss was linear through the first 7 hours but in the next 3 hours the rate of headloss development doubled. The rate of headloss development in the dual media filters clearly showed the detrimental effects of influent turbidities in excess of 30 NTU. Normally, the coarse media filters would have been backwashed after eight hours but additional data was desired for model predictions.

Runs 22 and 23

The purpose of Runs 22 and 23 was to evaluate filter system performance at higher flow rates. The dual media filters were not used during these two runs so that the headloss in the coarse media filters could be extended to a higher value. In Run 22, 25 mg/l of kaolin was used for the raw water suspension. The headloss reached over 3.2 m in two 21 hour filtration cycles. The effluent turbidity data indicated a long filter

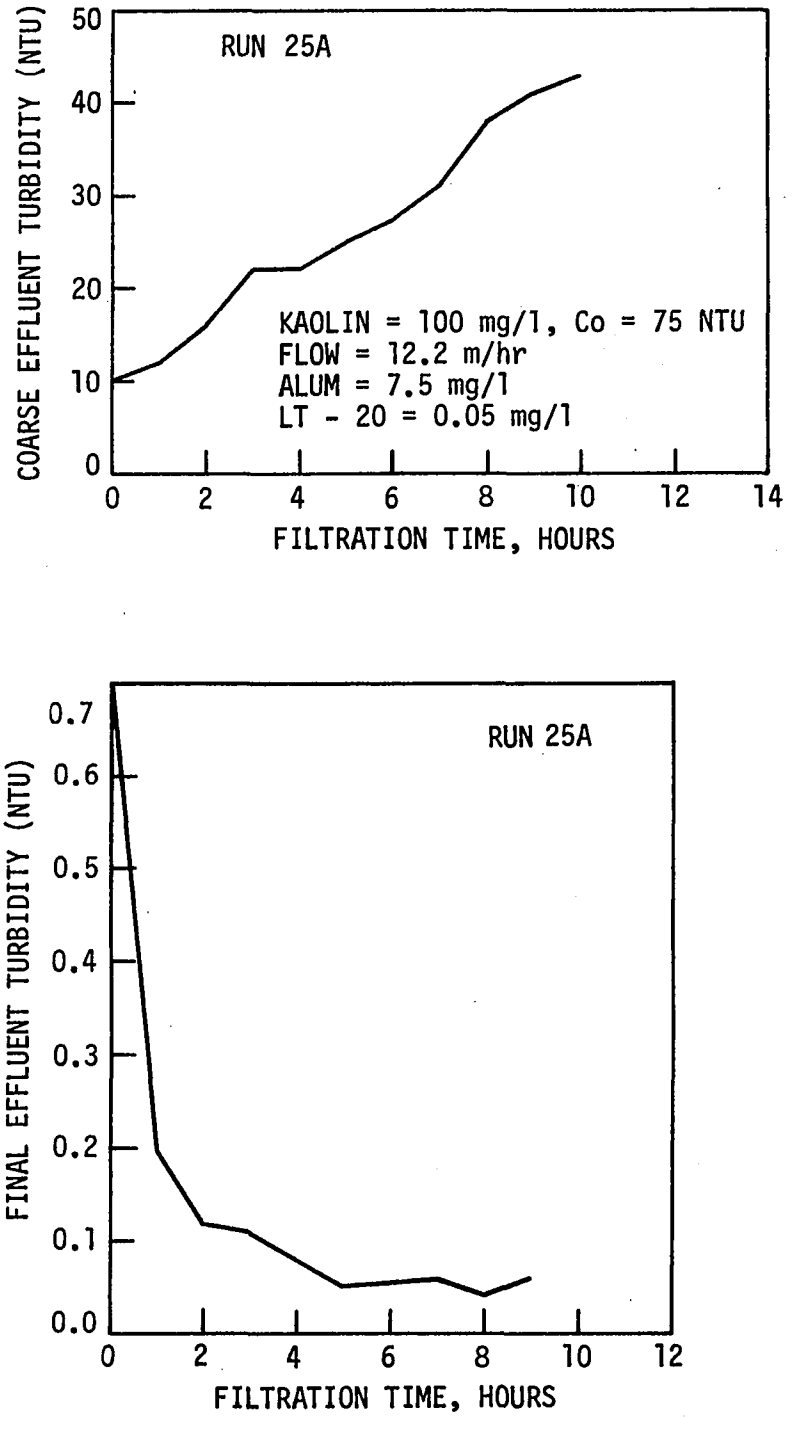


Figure 38. Effluent turbidity vs. filtration time, Run 25A, coarse media and dual media filters operated at constant rate

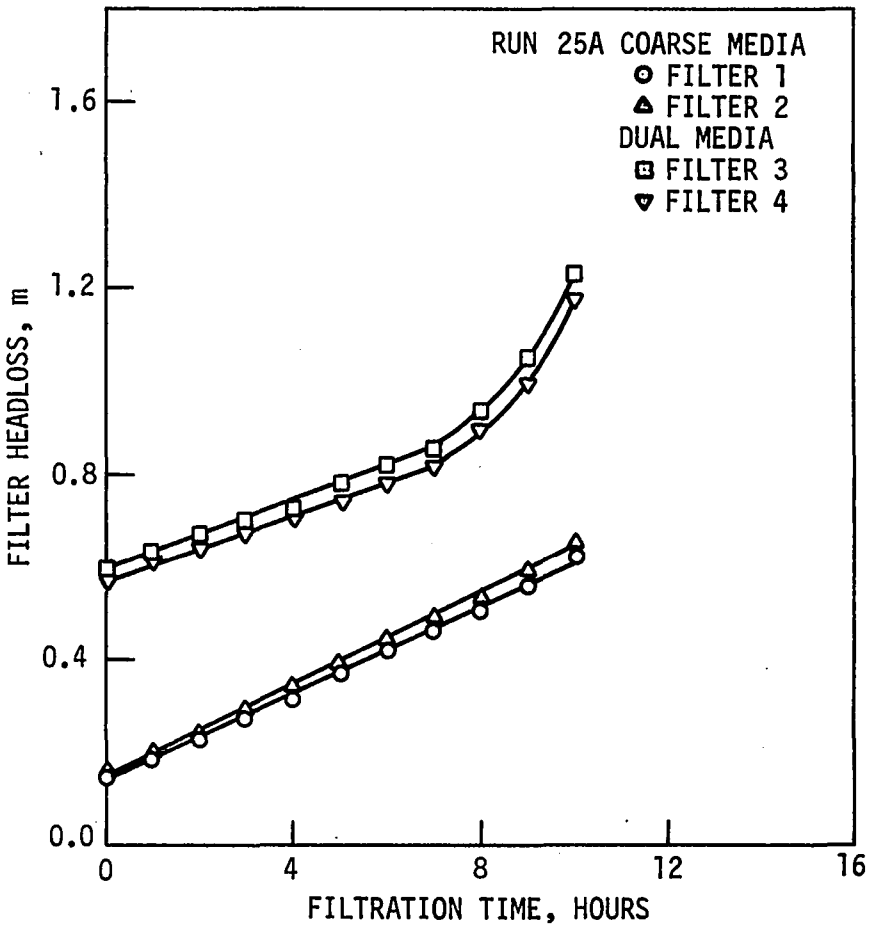


Figure 39. Headloss increase vs. filtration time, Run 25A, coarse media and dual media filters operated at constant rate

ripening period with the minimum turbidity produced at the end of the end of the filter cycle. Coarse media filter effluent turbidity values were less than 1.0 NTU for approximately one half of the filter run.

In Run 23, 50 mg/l of kaolin was used for the raw water suspension. The data for Run 23 appears in Figures 40 and 41. Although the criteria for backwashing the coarse media in terms of turbidity would have been met at a run length of 15 hours, the run was extended to 18 hours to a terminal headloss of about 3 m. The turbidity criteria for backwashing was two consecutive effluent turbidities exceeding 10 NTU which was about 30 percent of the influent turbidity of 36 NTU. The second filter cycle during this run also lasted 18 hours. During the second cycle the headloss again reached 3 m but the effluent turbidity did not exceed 5.5 NTU. This may be attributed to a ripening effect since the filters were cleaned between filter runs with chlorine bleach.

Run 24

Run 24 was made to test filter system performance at a flow rate of 24.4 m/hr. A kaolin concentration of 25 mg/l was used. At this flow rate, the headloss in the filter influent lines was too great to allow extending the run length beyond a terminal headloss of 2.2 m, since the flow rate could not be maintained. To remedy this, only one coarse media filter was used for the second filtration cycle and that filter operated at a flow rate of 24.4 m/hr. Results of this run are shown in Figures 42 and 43. Headloss development and effluent turbidity characteristics were similar for both cycles. The terminal headloss in the second cycle was 2.8 m. At the end of both cycles, the effluent turbidity was still

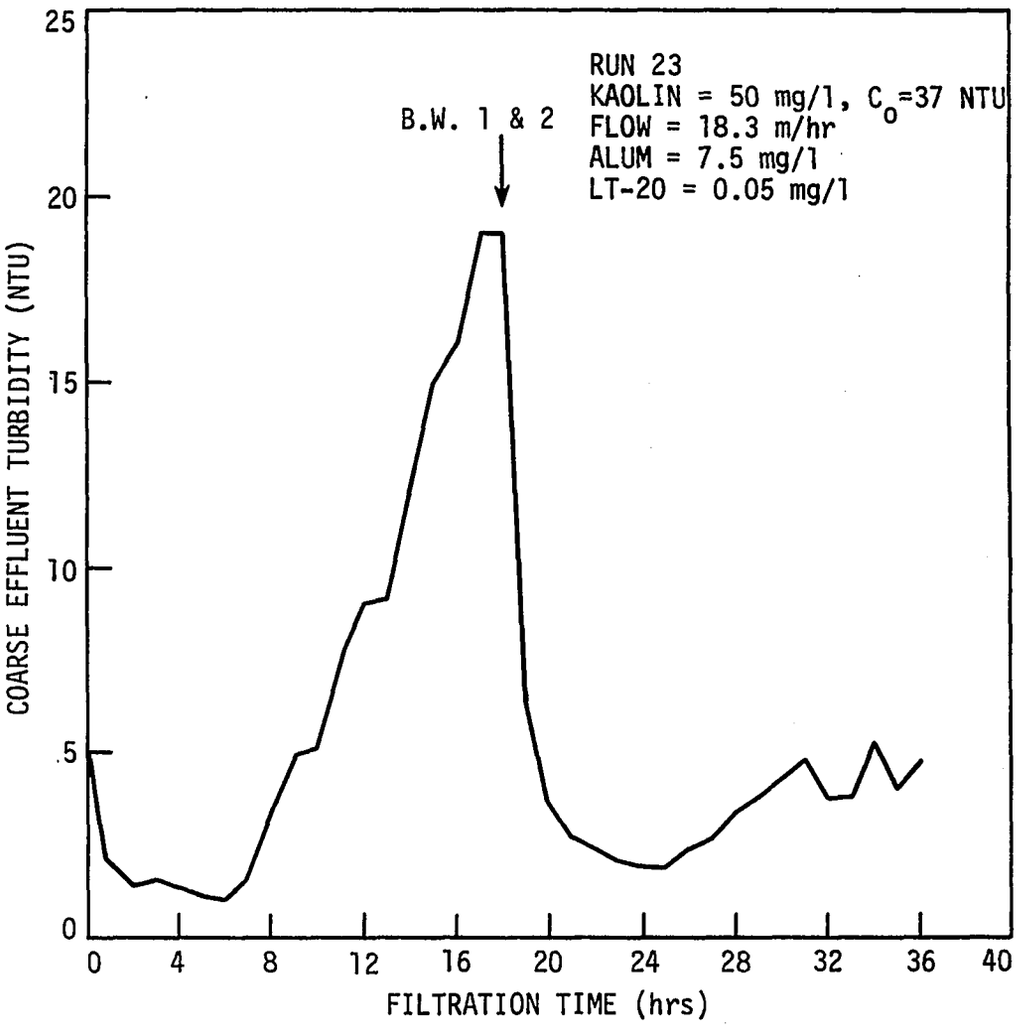


Figure 40. Coarse media effluent turbidity vs. filtration time, Run 23, coarse media filters operated at constant rate

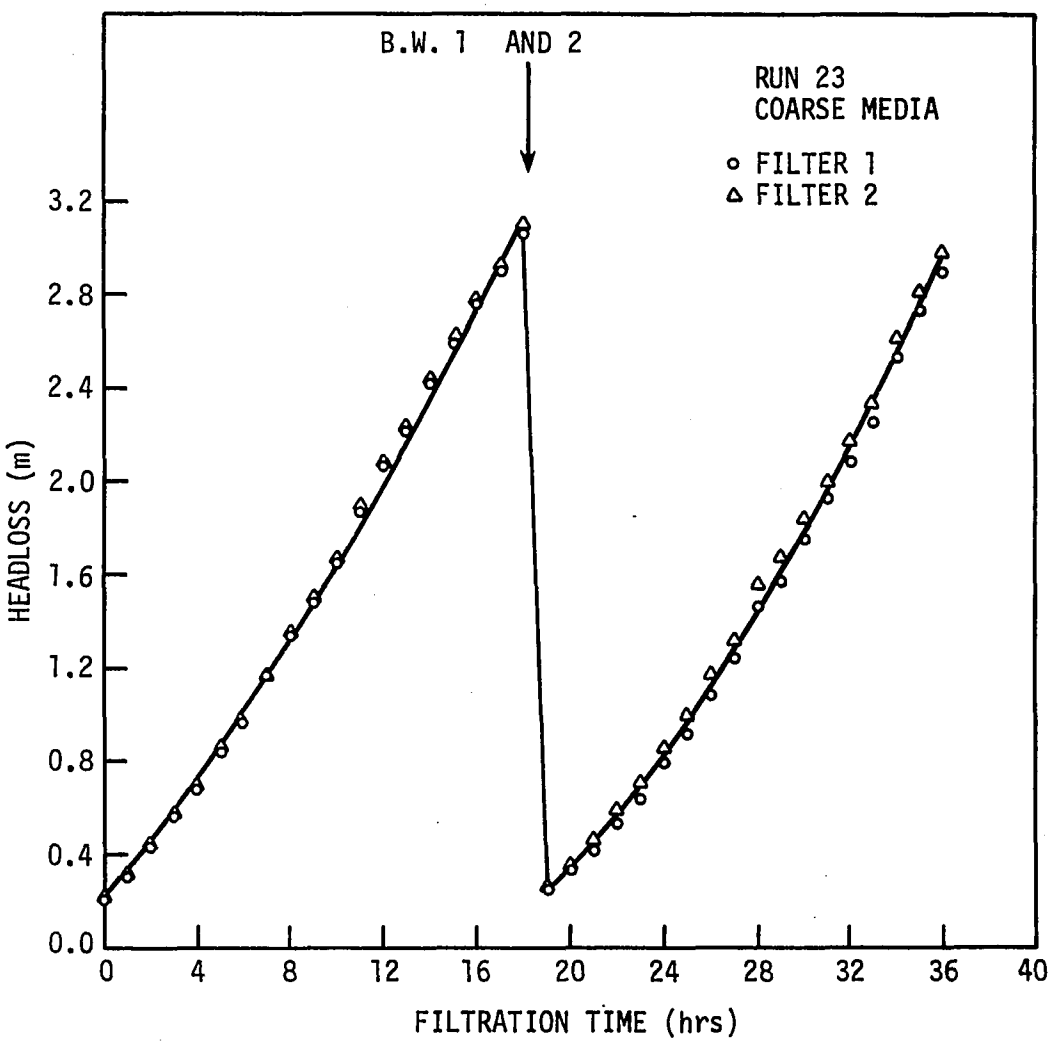


Figure 41. Coarse media headloss increase vs. filtration time, Run 23, coarse media filters operated at constant rate

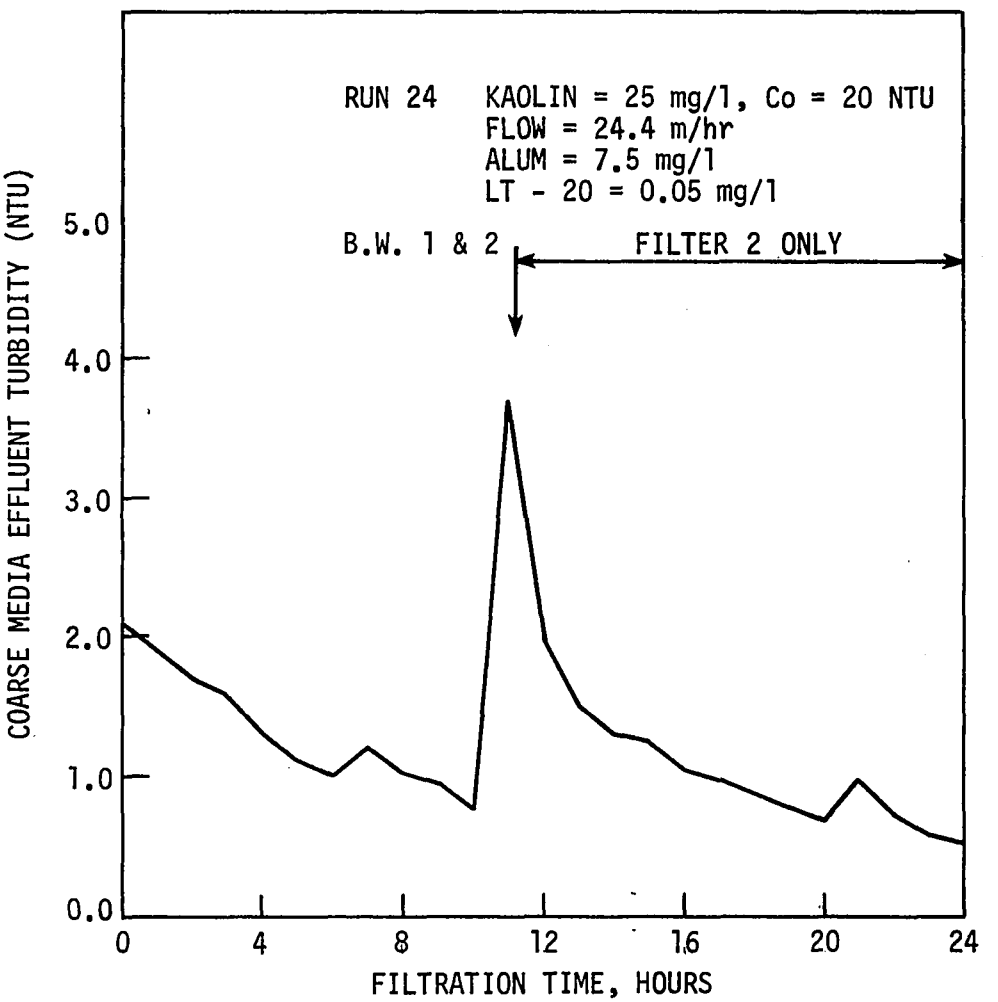


Figure 42. Coarse media effluent turbidity vs. filtration time, Run 24, coarse media filters operated at constant rate

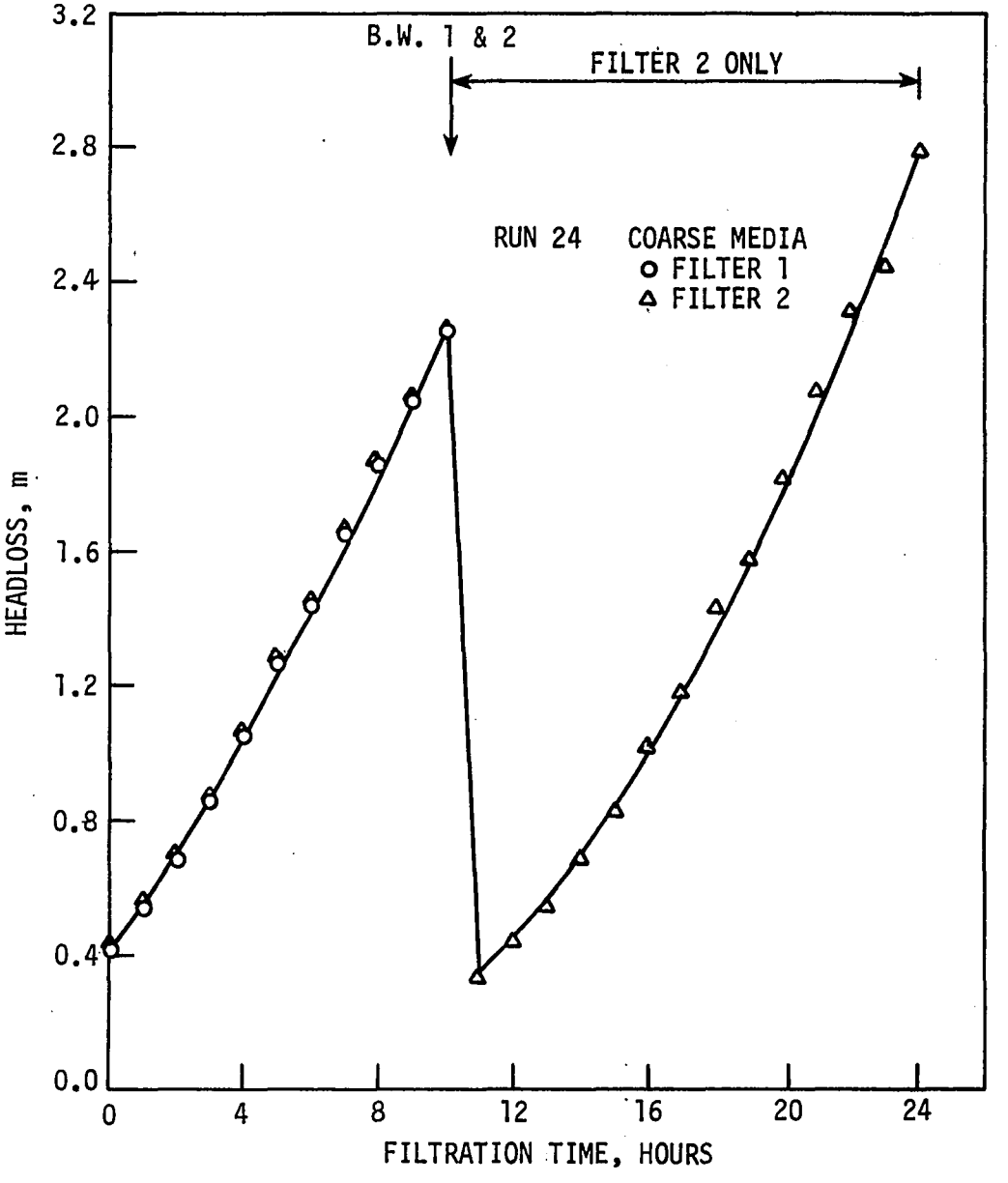


Figure 43. Coarse media headloss increase vs. filtration time, Run 24, coarse media filters operated at constant rate

decreasing.

Run 25B

In Run 25B, both the coarse media and dual media filters were operated in a declining rate mode. Instead of staggering the start up of a set of filters both sets were started at the same time and then one filter of each set was backwashed at some predetermined time and the other filter in the set remained dirty. Data for this run are shown in Figures 44 through 47. The chemical pretreatment and filtration rate used in Run 25B were the same as used in Run 12 and Run 13. In Run 12, the coarse media filters were operated in a declining rate mode and the dual media filters were operated at a constant flow rate. In Run 13, both the coarse media filters and dual media filters were operated at a constant flow rate. The data from this run showed no unusual trends in coarse media turbidity or headloss development compared to Run 12. The dual media headloss data showed the reduced effects during coarse media backwash as also shown by Run 12 when compared to Run 13. The variation in flow rates in each filter during the run as a result of backwashing filters is shown in Figure 47.

Run 26

Throughout the pilot filter system study, a pattern developed with respect to how solids were deposited in the coarse media filters. The pattern varied with influent solids concentration and to some extent with flow rate. The purpose of Run 26 was to make a run with low turbidity raw water and record the pattern of solids deposit with photographs

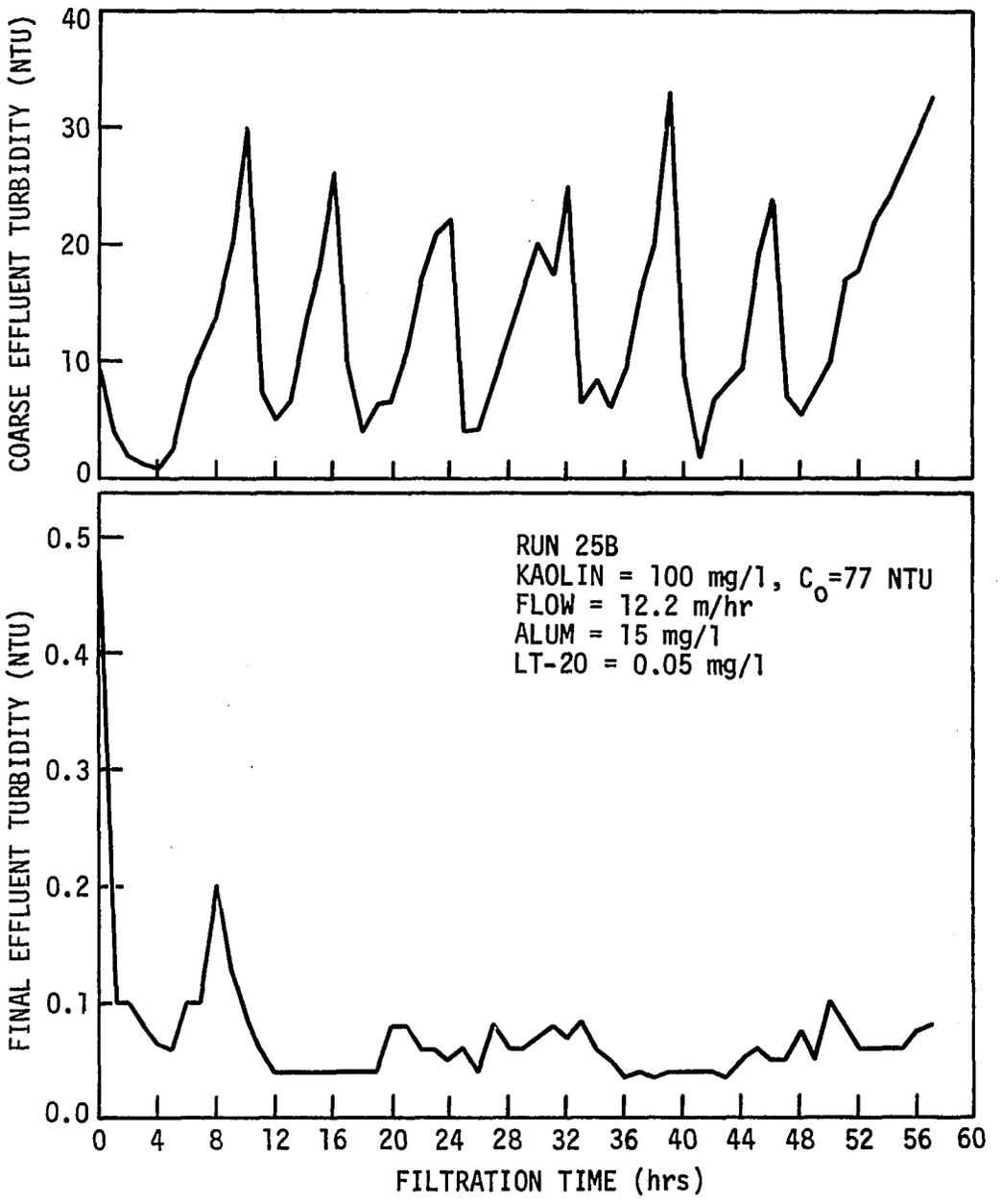


Figure 44. Effluent turbidity vs. filtration time, Run 25B, coarse media and dual media filters operated in declining rate

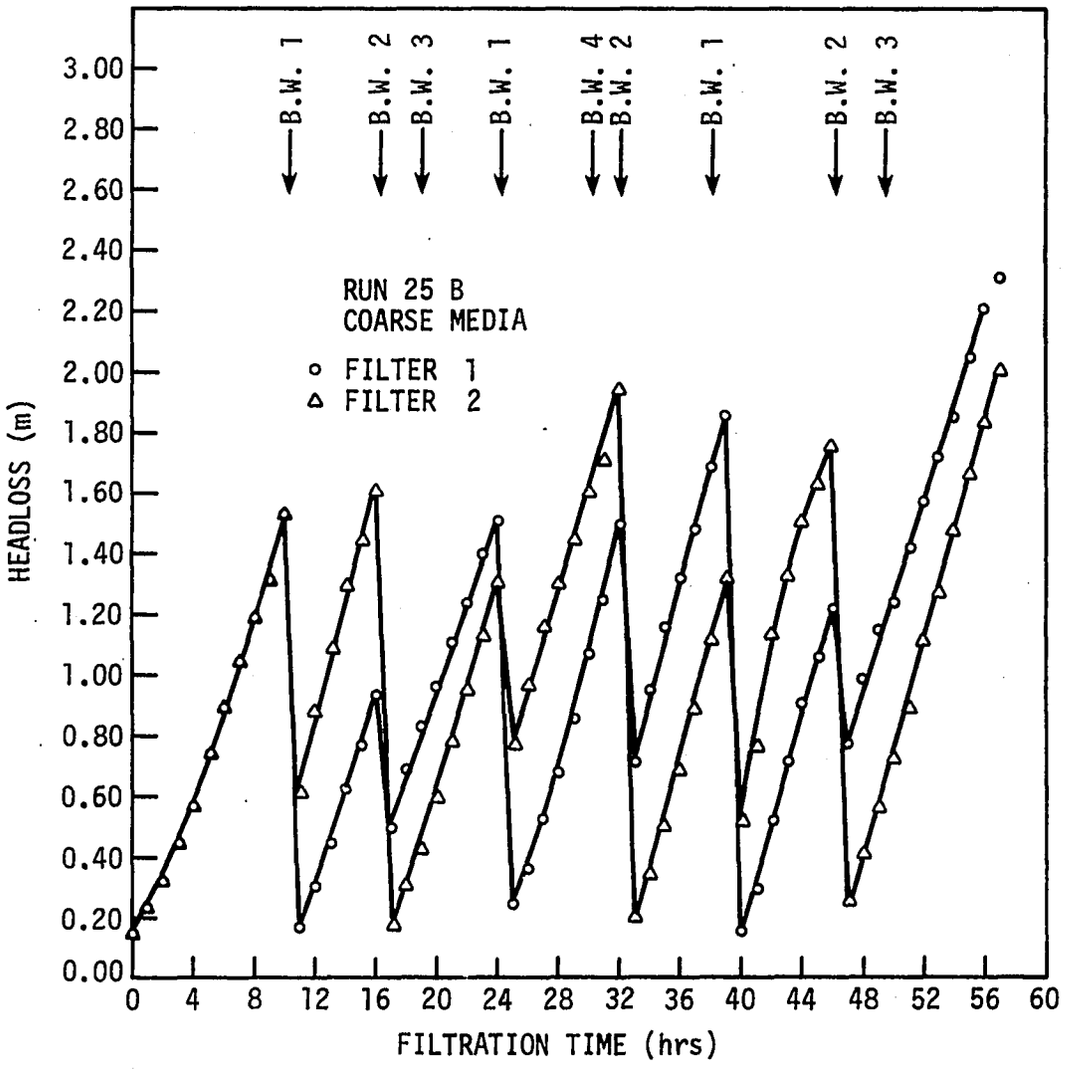


Figure 45. Coarse media headloss increase vs. filtration time, Run 25B, coarse media filters operated in declining rate

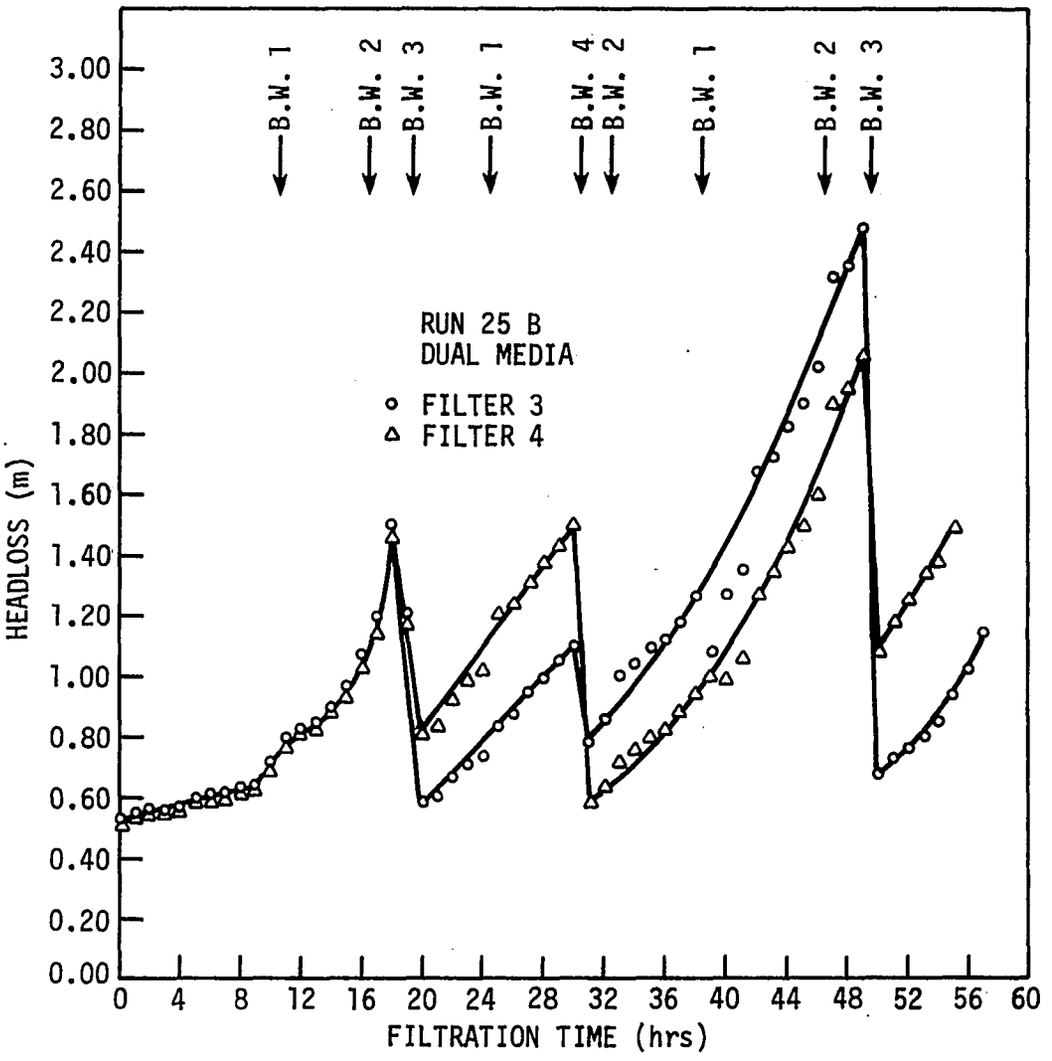


Figure 46. Dual media headloss increase vs. filtration time, Run 25B, dual media filters operated in declining rate

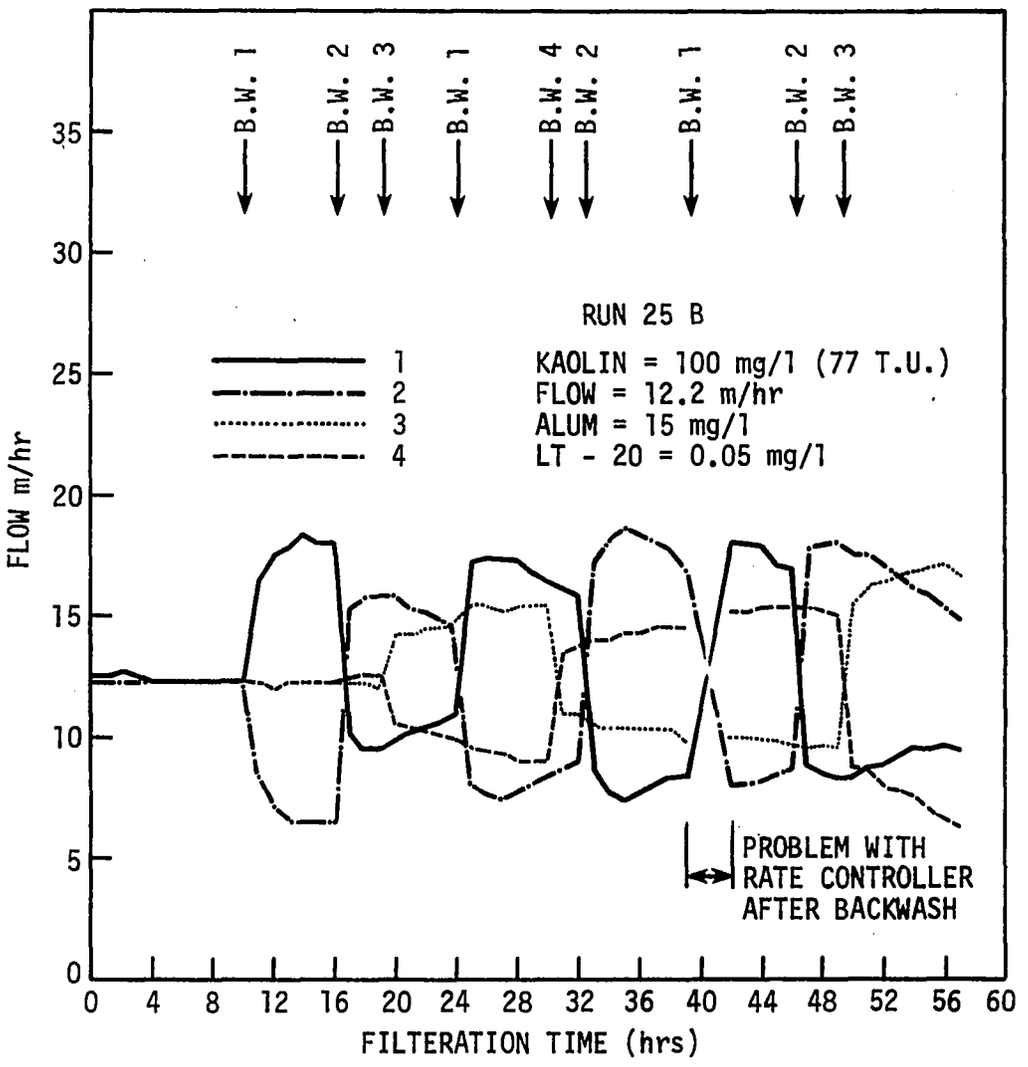


Figure 47. Filter flow rate variations, Run 25B, coarse media and dual media filters operated in declining rate

of the coarse media filters. The photographs are discussed in the discussion of results section of this report. During the first 3.0 hours, the alum dosage was 15 mg/l after which it was lowered to 7.5 mg/l. The LT-20 dosage was 0.05 mg/l throughout the filter run.

Run 27

Run 27 was made for two reasons: To record photographically the solids deposition behavior at 100 mg/l of kaolin, and to make a filter run without pH adjustment. The raw water pH without adjustment was about 8.0; alum addition lowered the pH to 7.6. The photographs will be discussed along with those of Run 26 in the discussion of results section of this report. Results of Run 27 can be compared to Run 25B since the only operational difference between the runs was the pH of the suspension. In Run 25B, the suspension pH after alum addition was about 7.0 and in Run 27 the suspension pH after alum addition was about 7.6. Run 27 results compared favorably with those obtained in Run 25B. Coarse media and dual media headloss rates and values exhibited the same patterns in both runs. Coarse media effluent turbidity was allowed to reach slightly higher values in Run 27, but the pattern in terms of rate of breakthrough was similar to Run 25B results. In both filter runs, the final turbidity was less than 0.1 NTU for most of the run.

Field study

The original intent of the field study portion of this research was to evaluate the filter system performance and optimum chemical pre-treatment requirements found in the laboratory when the raw water was a highly turbid surface water. Due to this years abnormal weather

patterns, a high turbidity surface water was not produced in the Des Moines River. Instead, the surface water at the time of the field investigation (April 7-12) was relatively low in turbidity, about 30 NTU, but high in diatom levels. An analysis of the water revealed that about 85% of the diatoms present belonged to the genus Stephanodiscus. These diatoms are round, disc-shaped organisms with an average size of about 10 μm. The only other genus of diatom present in significant numbers was Nitzschia. These diatoms are rod shaped and were about 30-40 μm in length. A high number of Euglena was also present. The diatom levels were high enough to impart a redish brown color to the water. In a water quality sampling program at Saylorville and Red Rock Reservoirs, diatom levels in the river water are measured on a monthly basis. One of the sampling points is just upstream of the low head dam on the Des Moines River at Boone. The river water used during the field investigation was pumped from behind the low head dam to the filtration system. The diatom level on March 23, 1981, at this site was about 42,000 cells/ml. On April 20, 1981, the diatom level was about 36,000 cells/ml. A study by Beckert (8) indicated that the levels reported in March and April were in the upper 10% of the maximum levels encountered over a 15-year period.

During the field study, the flow rate through the filters was maintained at 12.2 m/hr. The mode of filter operation was declining rate in both sets of filters. Various combinations of alum with either a nonionic or cationic polymers were used to enhance diatom removal.

the pH of the river water was adjusted with concentrated HCL to reduce the pH to about 7 to 7.2 after alum addition.

The turbidity results of the field study are shown in Figures 48 and 49. The headloss results are not presented due to the questionable and irratic behavior exhibited by the pressure gages used to measure the pressure drop across the filters.

The filter run was actually started before the zero time indicated in Figures 48 and 49. This was done to provide an operating period to acclimate the system to the river water treatment and to correct any operational difficulties. The optimum chemical pretreatment as determined in the laboratory studies was used but proved to be unsuccessful.

The raw water turbidity for the whole field test period averaged about 28 NTU with a standard deviation of 2.6. During the first 45 hours of the field filter runs an attempt was made to produce acceptable results (a filtered water turbidity below 1.0 NTU) with the chemical treatments, both types and dosages, that were found to be effective in the laboratory pilot runs. An alum dosage of 20 mg/l was used in combination with a nonionic polymer (Percol LT-20) dosage of 0.05 mg/l for the first 14 hours of the filter runs. With this combination of pretreatment chemicals, final effluent turbidity below 1 NTU could not be achieved. The alum dosage was reduced to 15 mg/l and the polymer dosage was increased to 0.1 mg/l for the next 31 hour period. This chemical pretreatment combination also proved to be ineffective in producing a final effluent turbidity below 1 NTU. At this time it was

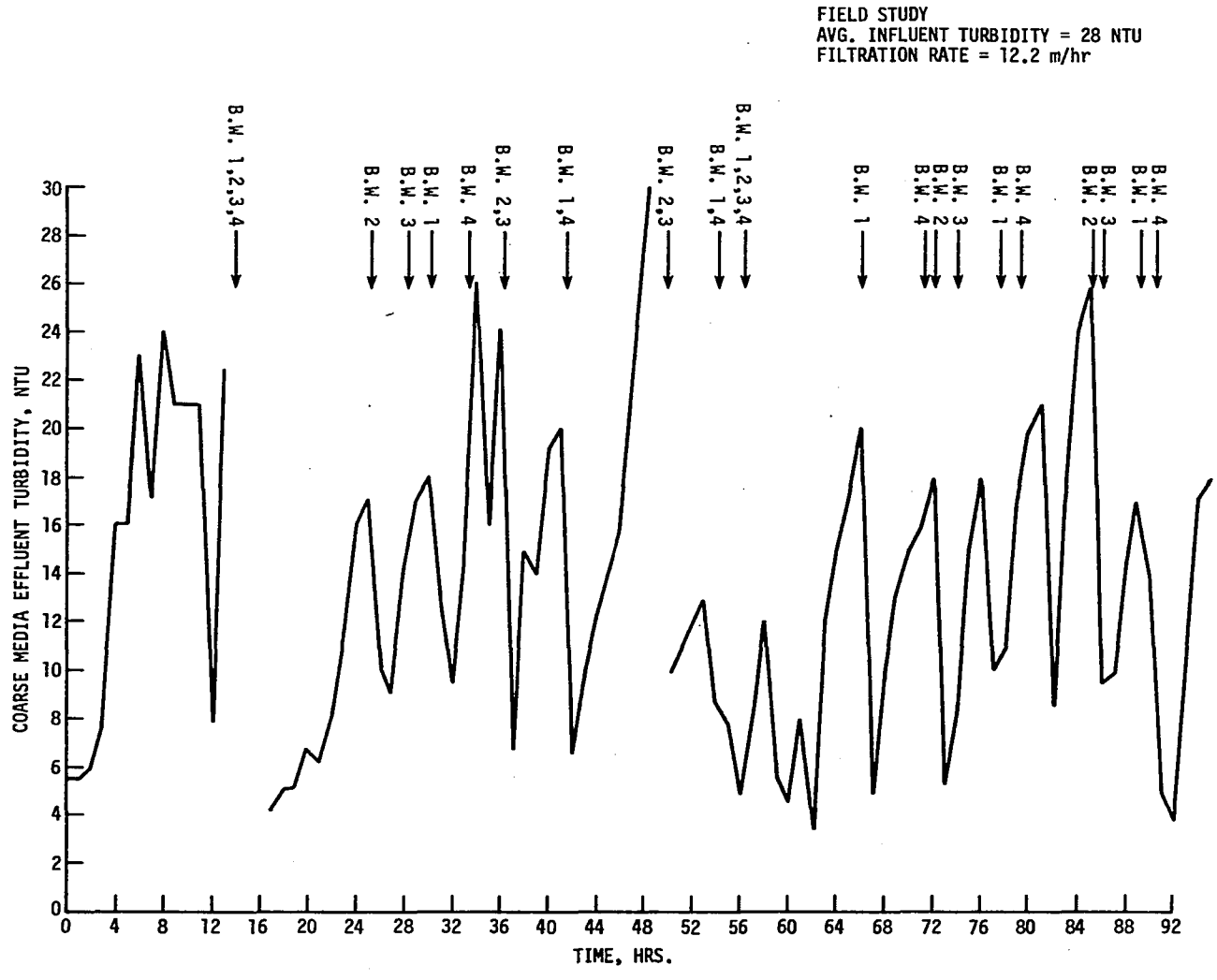


Figure 48. Coarse media effluent turbidity vs. filtration time, field study, coarse media filters operated in declining rate

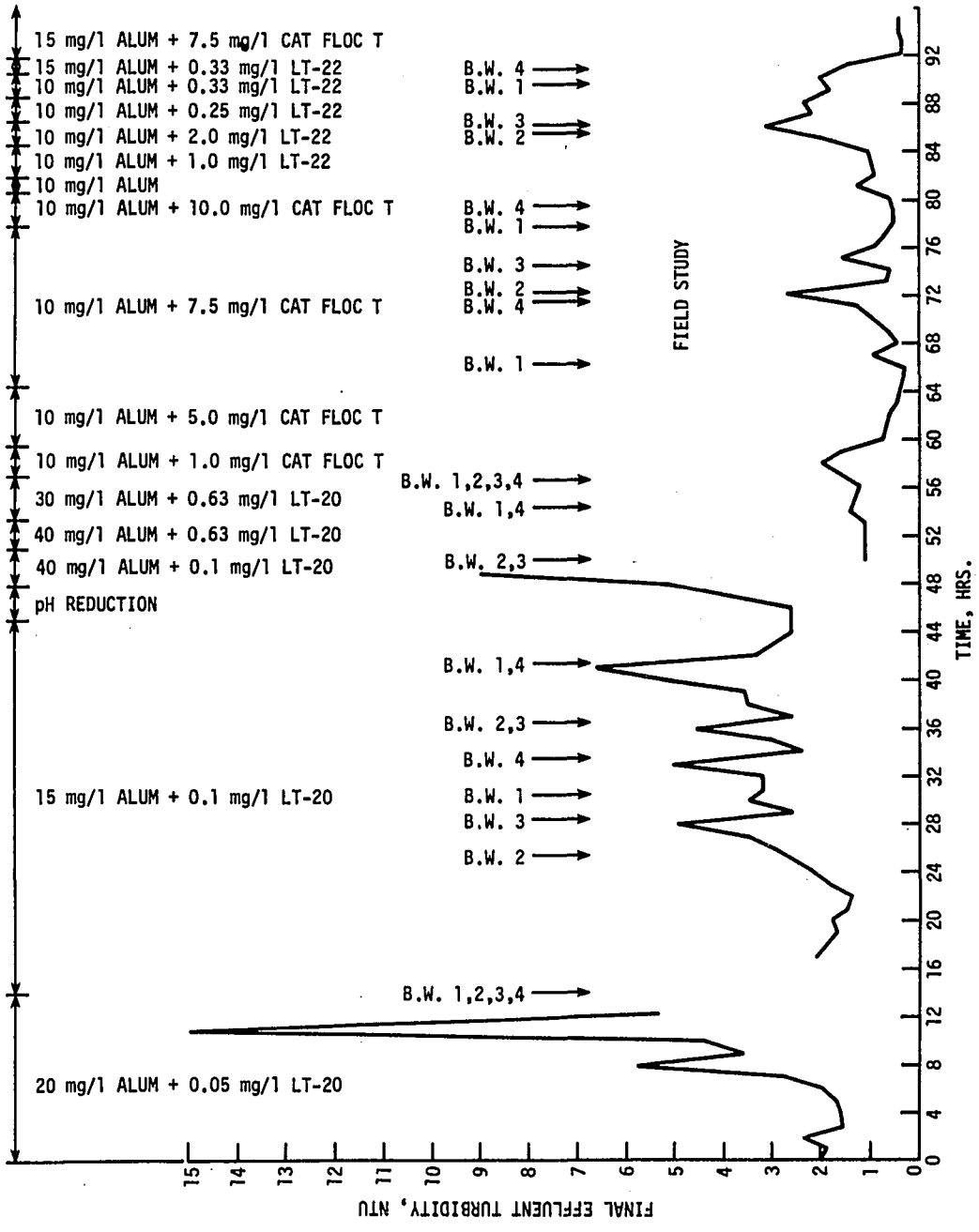


Figure 49. Dual media effluent turbidity vs. filtration time, field study, dual media filters operated in declining rate

thought that part of the problem with reaching the desired final effluent turbidity level was associated with a high color level in the river water. To enhance the removal of color, the raw water pH was reduced to about 6.5 and the alum dosage was maintained at 15 mg/l and the polymer dosage was maintained at 0.1 mg/l. Again, a final effluent turbidity below 1 NTU could not be obtained. The alum dosage was increased to 40 mg/l and the polymer dosage was first maintained at 0.1 mg/l and later increased to 0.63 mg/l. These combinations of chemical pretreatment reduced the final effluent turbidity to between 1.0 and 1.5 NTU after one coarse media filter and one dual media filter were backwashed. The use of 40 mg/l of alum resulted in a significant amount of floc formation prior to the coarse media filters. The alum dosage was reduced to 30 mg/l in an attempt to reduce the amount of floc formation. The final effluent turbidity remained between 1.0 and 1.5 NTU. When it was determined that the turbidity of the raw water was attributable to a high diatom count, the chemical pretreatment was changed to alum plus a cationic polymer and the pH was maintained between 7.0 and 7.5.

The use of 10 mg/l of alum in combination with either 1, 5, 7.5 or 10 mg/l of a cationic polymer, Cat Floc T, as the pretreatment chemicals resulted in achieving the 1.0 NTU final effluent turbidity goal. The only time when the final effluent turbidity was above 1.0 NTU was immediately after filter backwash and during turbidity breakthrough in the dual media filters.

The use of alum in combination with Cat Floc T appeared to have the most effect on the dual media filters. The coarse media filters developed less than 1.5 m of headloss when alum and Cat Floc T were used and the coarse media filter effluent turbidity results were not much different than when alum and Percol LT-20 were used. When alum and Cat Floc T were used the rate of headloss increase in the dual media filters averaged 0.3 m/hr.

Another cationic polymer, Percol LT-22, was used in combination with 10 mg/l of alum as the pretreatment chemicals. At polymer dosages between 0.25 and 2.0 mg/l, the final effluent turbidity was not reduced to the levels obtained with Cat Floc T and was generally above 1.0 NTU. Increasing the alum dosage to 15 mg/l produced a slight improvement in final effluent turbidity. Changing the polymer to 7.5 mg/l of Cat Floc T and maintaining the alum dosage at 15 mg/l resulted in the production of a final effluent water turbidity below 0.5 NTU for the final 4 hours of the field study filter run.

DISCUSSION OF RESULTS

Jar Tests

In this study the jar test was used to determine which coagulants would be effective for particle destabilization and an approximate optimum chemical dosage for their use. From the jar test results, it was found that flocculation time required for effective performance was dependent upon flocculation speed and particle concentration. The length of the flocculation period necessary to produce a settleable floc at pH 8 was less than anticipated. At a pH of 6, the required flocculation period needed was greater than at pH 8, especially for low alum dosages. The results of the jar tests reinforced the decision to rely only on flocculation that would occur within the influent line and the coarse media filters in subsequent pilot testing of coarse media and dual media filters operated in series. At the optimum chemical dosages and the higher particle concentrations, low settled water turbidities occurred after only 5 to 10 minutes of flocculation. At lower particle concentrations, flocculation time and mixing intensities became more influential in producing a settleable floc particle. This was probably due to the decreased chance of particle-coagulant interactions at the lower particle concentrations.

The jar tests indicated that the amount of alum needed to bring particle zeta potential to zero was not affected by either flocculation speed or time. In general, the particle zeta potential remained relatively unchanged at flocculation periods ranging from 0 to 25 minutes. This

result is contrary to the findings of Monscvitz et al. (64) in their investigation using Lake Mead water. These investigators found that the amount of alum required to produce a zero particle zeta potential was reduced from 22 mg/l when no flocculation was used to 8 mg/l when 30 minutes of flocculation was used. Jar tests at pH values of 6 and 8 showed that particle zeta potential was not directly related to optimum settled water turbidity (Figure 6). At a pH of 6, similar settled water turbidity results were shown with alum dosages between 10 to 20 mg/l. The particle zeta potential values ranged between + 5.0 mV and + 17.0 mV (Figure 7). At an alum dosage of 5 mg/l the particle zeta potential was about -2.0 mV and the settled water turbidity was much higher than with the other alum dosages evaluated. At a pH of 8, the alum dosages varied between 10 and 40 mg/l and the particle zeta potential ranged from about -10.0 mV to + 4.0 mV. Similar settled water turbidity trends were found for all alum dosages at a water pH of 8.0. More alum was required at a water pH of 8.0 than at a pH of 6 to bring the particle zeta potential to zero.

The chemical combinations which produced the optimum settled water turbidity in the jar test also produced the optimum filtered water turbidity using the pilot plant filters. The optimum dosages of coagulants were lower for filtration than for production of a settleable floc in the jar tests. The difference in required dosage between jar tests and filtration have also been noted by Conzett (25) and O'Melia (66). O'Melia claims the differences in required dosages may be due to adsorption of coagulant onto the glassware in the jar test procedure.

Nevertheless, the importance of the jar test as related to the selection of the optimum chemical coagulants for direct filtration should not be underestimated.

The purpose of coagulation in the jar test is to produce a large settleable floc particle. The mixing intensity during a jar test is usually constant during the flocculation period. In direct filtration a small dense floc is more desirable since the larger flocs are generally less stable in the presence of shear forces and can be broken apart if formed prior to filtration or within the filter during direct filtration. In filtration, the intensity of agitation or mixing within the filter will change as the filter run progresses, since solids accumulate within the filter media pores resulting in an increase in interstitial velocity. The intensity of agitation within a filter may also vary as a function of filter depth since the upper portion of a filter in gravity or downflow filtration will become clogged with solids before the lower portion of the filter.

Filter runs

One of the major objectives of this research was to evaluate the use of coarse media filters in series with dual media filters as an alternative method of direct filtration which can be used with higher water turbidity levels. Most of the pilot filter runs used kaolin as the suspension particle at concentrations between 25 and 300 mg/1. The performance of the filtration system was evaluated on the basis of effluent turbidity and filter headloss under varying conditions of chemical

pretreatment, flow rate and raw water turbidity.

The quantitative effects of kaolin concentration and flow rate on the coarse media filter run length are summarized in Figure 50. Table 10 gives the source of the data used to construct Figure 50. During the study, a lower limit of 10 hours was used as the minimum desired coarse media filter run length and served as one of the criteria for a successful run. In general, a dual media filter run length can be expected to last a minimum of 20 to 30 hours if the coarse media filters are operating properly. The results in Figure 50 show that this criterion could be met at a kaolin concentration of up to about 125 mg/l and a flow rate of 12.2 m/hr. At kaolin concentrations in excess of 125 mg/l, the filter run lengths were shorter and usually had to be terminated due to turbidity breakthrough. Higher flow rates than 12.2 m/hr were not evaluated at this kaolin concentration since the coarse media run lengths were just above the minimum allowable and higher flow rates would have shortened the filter cycle. Flow rates less than 12.2 m/hr were also not evaluated because 12.2 m/hr was set as the minimum practical flow rate. When the kaolin concentration was 100 mg/l or greater, the coarse media filter runs were always terminated due to turbidity breakthrough. (Note: run lengths could have been increased by using a deeper coarse media.)

At kaolin concentrations of 25 and 50 mg/l, the coarse media filter runs were terminated due to headloss. The run lengths in Figure 50 were based on the time to reach more than 2 m of headloss. The 2 m headloss limit was used as a criterion for run termination since it is a common

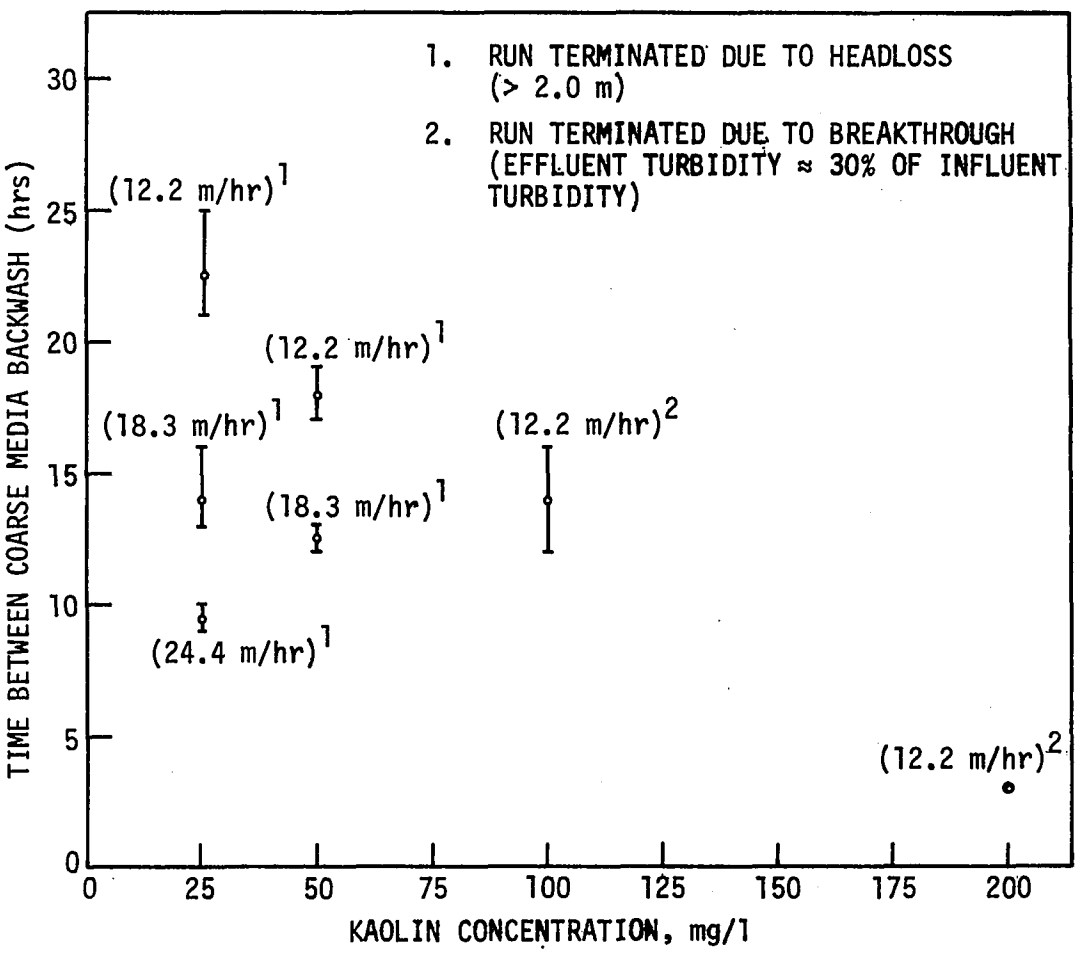


Figure 50. Time between coarse media filter backwash vs. kaolin concentration

Table 10. Source of data used to construct Figure 50

Suspension Concentration, mg/l	Run No.	Chemical Feed, mg/l Alum	Polymer	Filtration Rate, m/hr
25	21	7.5	0.05	12.2 (constant)
	22 and 26	7.5	0.05	18.3 (constant)
	24	7.5	0.05	24.4 (constant)
50	20	7.5	0.05	12.2 (constant)
	23	7.5	0.05	18.3 (constant)
100	13	15	0.05	12.2 (constant)
	6, 12, 25B and 27	15	0.05	12.2 (declining)
200	18	20	0.05	6.1 (constant)
	18	20	0.05	12.2 (constant)

limiting headloss; and headloss in excess of 2 m in the pilot filter system resulted in difficulties in maintaining the desired flow rate. If more pressure drop were available, the coarse media filter run lengths could have been increased. As an example, in Run 23 using a kaolin concentration of 50 mg/l and a flow rate of 18.3 m/hr, the coarse media filter run was extended to about 18 hours with a terminal headloss of about 3.0 m by removing the dual media filters from the flow scheme. The results in Figure 50 indicate that there was a direct correlation between coarse media run length and influent kaolin concentration at a particular filtration rate. At a given filtration rate, the coarse media filter

run length increased as the kaolin concentration used decreased. At kaolin concentrations of 25 and 50 mg/l the run lengths were roughly inversely proportional to filtration rate. The higher the filtration rate the shorter the coarse media filter run length. A 50 or 100 percent increase in filtration rate results in approximately the same percentage decrease in filter run length.

In all filter runs included in Figure 50, alum was always used as a primary destabilizing chemical. In addition to alum, polymers were used as filter aids. The most often used polymer was Percol LT-20, (approved for water treatment), a nonionic polymer similar to Percol 720 (not approved for water treatment). The alum and nonionic polymer dosages found to be the most successful in jar test experiments were higher than those proved necessary for direct filtration. The use of alum alone was demonstrated in Run 1 not to be a successful pretreatment alternative. Alum dosages of 10, 20, and 25 mg/l were used with limited success in producing the desired quality of effluent, i.e. a turbidity consistently below 0.5 NTU. With the lower alum dosage, a reasonable minimum turbidity could not be reached. At the higher alum dosages the coarse media filter effluent turbidity was reduced but the filter run lengths were short, averaging only about 6 hours. The headloss development in the coarse media filters was relatively low when alum was the sole coagulant (Table 11, Run 1). On the average, 0.026 m of headloss per hour developed at a filtration rate of 12.2 m/hr. A large portion of the solids removal burden was placed on the dual media filters. The headloss development in the dual media filters averaged about 0.087 m/hr and the dual media

Table 11. Summary of filter effluent turbidity and headloss increase during the laboratory pilot filter runs

Run No. (D)[a] (DB)[b]	Coarse Media Filter Run Length hrs.	Coarse Media Filter Net Head-loss Increase m	Coarse Media Filter Headloss Rate of Increase[c] m/hr	Coarse Media Filter Effluent Turbidity, NTU Terminal/Average	% Turbidity Removal in the Coarse Media Filters Terminal/average	Dual Media Filter Run Length, hrs.	Dual Media Filter Net Head-loss Increase, m	Dual Media Filter Headloss Rate of Increase m/hr	Dual Media Effluent Turbidity NTU Terminal/ Average ± Standard Deviation
1	9	0.18	0.02	58/42	41/58	21	1.83	0.087	5.5/1.35 ± 1.4
	5	0.15	0.03	41/28	59/72				
	5	0.14	0.028	42/30	58/70				
2	13	1.49	0.115	22/11	76/88	23	0.48	0.021	0.038/0.08 ± 0.1
4(D)	13	1.46	0.152	28/15	70/84	24	1.9	0.079	0.06/0.08 ± 0.06
	12	0.78	0.144	17/16.5	82/82				
5(D)	9	2.22	0.341	0.16/0.81	99.8/99.1	24	0.45	0.019	0.13/0.15 ± 0.06
	11	2.68	0.455	3.4/1.0	96.3/98.9				
	9	1.40	0.423	10/2.8	89/97				
	9	1.87	0.263	0.27/2.2	99/7/97.7				
6(D)	12[d]	0.66	0.100	37/27	51/65	24	1.2	0.05	0.37/0.43 ± 0.38
	14[e]	0.98	0.104	13/25	83/66				
	11[f]	1.42	0.168	2.4/7.0	96.8/90.8				
8	12	-	-	75/67	9/18	12	0.28	0.023	39/33 ± 5.1
	9	-	-	75/67	9/18				

9(D)	11	0.71	0.102	33/19.3	58/76	20	1.42	0.071	0.16/0.13 ± 0.09
	13	0.79	0.082	29/20.8	63/74				
	15	1.39	0.103	32/20.3	59/74				
10(D)	11	0.05	0.005	49/36	38/54	19	1.52	0.08	4.8/1.35 ± 1.34
12(D)	12	0.80	0.103	33/23	60/71	20	1.71	0.086	0.11/0.17 ± 0.12
	14	0.096	0.099	33/23	59/72	21	1.69	0.08	0.17/0.15 ± 0.06
	15	1.42	0.117	44/18	46/78				
	13	0.83	0.118	27/16	66/81				
	13	0.86	0.073	36/22	56/73				
13	14	1.56	0.110	41/12.6	49/84	27	2.10	0.078	0.07/0.05 ± 0.02
	11	0.96	0.090	57/14.3	29/82				
	13	1.40	0.110	32/14.2	60/82				
15	9	0.52	0.058	32/17	60/79	22	1.85	0.084	0.04/0.06 ± 0.09
	11	0.78	0.071	30/19	63/77				
	13	1.07	0.082	42/19	48/77				
17	34	1.62	0.048	19/8	41/75	34	0.85	0.025	0.09/0.07 ± 0.04

[a] Filter runs using a declining rate mode in the coarse media filters.

[b] Filter runs using a declining rate mode in both sets of filters.

[c] Average rate when declining rate operation is used.

[d] 10 mg/l alum.

[e] 10 + 15 mg/l alum.

[f] 15 mg/l alum.

Table 11. continued

Run No. (D)[a] (DB)[b]	Coarse Media Filter Run Length hrs.	Coarse Media Filter Net Head-loss Increase m	Coarse Media Filter Headloss[c] Rate of Increase m/hr	Coarse Media Filter Effluent Turbidity, NTU Terminal/Average	% Turbidity Removal in the Coarse Media Filters Terminal/ average	Dual Media Filter Run Length, hrs.	Dual Media Filter Net Head-loss Increase, m	Dual Media Filter Headloss Rate of Increase m/hr	Dual Media Ef-fluent Turbidity NTU Terminal/ Average ± Stan-dard Deviation
19	10	1.24	0.124	36/15	64/85	21	1.41	0.067	0.08/0.06 ± 0.01
	10	1.36	0.136	30/15	70/85	23	1.36	0.059	0.04/0.07 ± 0.03
	9	1.00	0.111	34/13	66/87				
	9	1.03	0.114	35/17	65/83				
20	15[f]	2.09	0.139	0.36/0.82	99/98	45	1.14	0.025	0.11/0.09 ± 0.14
	14[d]	1.94	0.139	0.47/0.96	99/97	38	1.18	0.031	0.05/0.07 ± 0.06
	14[g]	0.84	0.060	6.5/6.3	82/83				
	19[h]	1.75	0.092	9.0/3.5	76/91				
21	26	1.92	0.074	4.7/1.1	75/94	47	0.74	0.016	0.10/0.06 ± 0.02
	20	2.00	0.100	0.31/0.86	98/95				
	20	1.92	0.096	0.34/1.22	98/94				
22	22	3.11	0.141	0.41/1.1	98/94				
	19	3.08	0.162	0.36/1.3	98/93				
23	18	2.90	0.161	19/7.1	49/83				
	17	2.75	0.162	4.8/3.4	87/91				
24	10	1.85	0.185	0.75/1.3	96/93				
	13	2.45	0.188	0.52/1.2	97/94				

25A	10	0.46	0.046	43/26	43/65	10	0.62	0.062	0.06/0.14 ± 0.19
25B (DB)	10	1.38	0.138	30/9.3	61/88	18	0.98	0.054	0.04/0.09 ± 0.10
	16	1.44	0.161	26/10.5	66/86	30	0.98	0.060	0.07/0.08 ± 0.08
	13	1.32	0.144	22/12.4	71/84	31	1.88	0.075	0.05/0.05 ± 0.02
	15	1.74	0.165	25/12.8	68/83				
	14	1.60	0.185	33/13.8	57/82				
	13	1.55	0.201	24/12.7	69/84				
	16	2.00	0.168	33/15.6	57/80				
26	30	3.22	0.107	6/4	70/80				
27 (DB)	9	1.34	0.149	25/9.4	72/89	22	0.80	0.036	0.035/0.07 ± 0.03
	15	1.90	0.185	30/12	66/87	31	1.10	0.049	0.05/0.07 ± 0.03
	12	1.38	0.198	28/15.8	69/82	20	0.85	0.060	0.03/0.05 ± 0.01
	12	1.38	0.160	30/15.2	66/83				
	12	1.44	0.180	35/16.3	61/82				
	14	1.57	0.164	30/19	66/79				

[g] 5 mg/l alum.

[h] 7.5 mg/l alum.

effluent turbidity over a 21 hour run length was 1.35 NTU.

The use of 0.05 mg/1 of Percol 720 in combination with 20 mg/1 of alum in Run 2 produced dramatic improvements in the performance of the coarse media filters and in the filtration system as a whole (Table 11). The coarse media filter run length was approximately doubled and much more headloss, and consequently solids removal, occurred in the coarse media filters. During this run, the average rate of headloss development in the coarse media was 0.115 m per hour. The final effluent turbidity averaged less than 0.1 NTU.

The excellent results obtained in Run 2 provided a basis for comparison with Runs 4 and 9. Both of these runs were also made with 20 mg/1 of alum. In Run 4, the polymer was changed to Percol LT-20 since this polymer was approved for water treatment whereas Percol 720 was not. In Run 4, the start-up of the second coarse media filter was delayed. In Run 9, the chemical dosage was the same as in Run 2 but the second coarse media filter start-up was delayed and a new supply of kaolin was used. The performance of the coarse media filters was similar in all three filter runs. The average headloss increase and turbidity values for Runs 4 and 9 are shown in Table 11. (In all filter runs using a declining rate operational mode, the average rate of headloss increase shown in Table 11 is the average rate of headloss increase when a filter received a majority of the flow and when a filter received a reduced flow.) The average coarse media filter effluent turbidity was the least in Run 2 where a constant filtration rate was used compared to Runs 4 and 9 where a declining rate mode was used. The most noticeable

difference between Runs 2, 4, and 9 was the amount of headloss which developed in the dual media filters. In Run 2, only 0.48 m of headloss developed after 23 hours. In run 4, 1.9 m of headloss developed in 24 hours and in Run 9, 1.42 m developed after 20 hours. The final effluent turbidity in Runs 2 and 4 averaged 0.08 NTU, but in Run 9 the average final effluent turbidity was 0.13 NTU.

In Runs 5 and 6, the polymer dosage was increased to 0.1 mg/1. With the increased polymer dosage in Run 5 and an alum dosage of 20 mg/1, the coarse media filter runs were shortened due to headloss development. This was the only filter run using 100 mg/1 of kaolin in which the coarse media filter runs were terminated due to headloss limitations. The average rate of headloss development ranged from 0.263 to 0.455 m per hour. The average coarse media effluent turbidity remained low during each filter cycle and ranged from 0.81 to 2.85 NTU. The dual media filters in Run 5 developed only 0.45 m of headloss in the 24 hour run. However, the average dual media effluent turbidity was 0.15 NTU. The dual media effluent turbidity was unusually high and may be explainable by the low solids load to the dual media and the existence of a correspondingly longer ripening period.

The high coarse media turbidity results in Run 6 in which the polymer dosage was 0.1 mg/1 and the alum dosage varied from 10 to 15 mg/1 indicated that 10 mg/1 of alum was not enough for effective filtration. Fifteen mg/1 of alum proved, however, to be almost as effective as 20 mg/1 of alum without the excessive headloss development. With 15 mg/1 of alum and 0.1

mg/l of polymer, the performance of the coarse media filters was similar to their performance in Runs 2, 4 and 9. The headloss development in the dual media should have been similar to that in Runs 4 and 9 (both runs used the variable start up procedure in the coarse media filters) but in Run 6 the rate was lower. This can be explained by the overall higher dual media effluent turbidity. The dual media effluent turbidity averaged 0.43 NTU and was never lower than 0.22 NTU. The high dual media effluent turbidity in Run 6 can be attributed to the insufficient alum dosage (10 mg/l) used during the first 10 hours of the filter run. In Run 6 it was determined that a lower alum dosage could be used for particle destabilization. To evaluate the filtration system performance at lower alum dosages and 0.05 mg/l of Percol LT-20, Runs 12, 13 and 15 were performed. In Runs 12 and 13, the alum dosage was 15 mg/l. The only difference between Run 12 and Run 13 was that in Run 12 a staggered start-up of the coarse media filters was used while in Run 13 both filters were started at the same time. The average rate of headloss increase in the coarse media filters was about 0.10 m/hr during both filter runs. The average coarse media filter effluent turbidity was less in Run 13 than in Run 12 which can be explained by the longer duration of low turbidity from the coarse media filters followed by a very rapid breakthrough. When the filter operation was staggered such as in Run 12, higher coarse media filter effluent turbidities were encountered but the breakthrough was more gradual.

The rate of headloss increase shown in Table 11 for the dual media filters in Runs 12 and 13 appears to be about equal. This is a little

deceiving, however, because over 1 m of headloss developed in the dual media filters in a one hour period following the backwash of the coarse media filters in Run 13. The average final effluent turbidity was less in Run 13 than in Run 12. Part of this can be attributed to the higher average solids load to the dual media filters during Run 12 when compared to Run 13.

Run 15 was made with 10 mg/l of alum and 0.05 mg/l of polymer with a constant rate operational mode. Comparing the results of Run 15 to the results in Run 13, which differed from Run 15 only in the amount of alum used, shows that the lower alum dosage decreased the coarse media and dual media filter run lengths. The coarse media effluent turbidity was higher at the lower alum dosage while the average final effluent turbidity was slightly lower. In most cases, the terminal headloss and the rate of headloss development in the coarse media filters was less at 10 mg/l of alum. The characteristic period of low turbidity from the coarse media filters followed by rapid breakthrough was shortened with the lower alum dosage.

Runs 20 through 24 were made to evaluate filter system performance with lower kaolin concentrations. The use of 0.05 mg/l of polymer was maintained throughout these five filter runs. In Run 20 the effect of alum dosage was evaluated. Alum dosages of 5, 7.5, 10 and 15 mg/l were used. This run indicated that 7.5 mg/l was the lowest alum dosage which produced satisfactory results. The chemical combination of 7.5 mg/l of alum and 0.05 mg/l of nonionic polymer produced characteristic coarse

media filter effluent curves similar to those which resulted when 100 mg/l of kaolin was used in the raw water suspension and 15 to 20 mg/l of alum and 0.05 mg/l of polymer were the pretreatment chemicals. In Run 20 the use of 10 and 15 mg/l of alum produced higher rates of headloss in the coarse media filters compared to when 7.5 mg/l of alum was used. An alum dosage of 5 mg/l did not provide satisfactory treatment since at this alum dosage turbidity breakthrough in the coarse media filters was more severe compared to when the higher alum dosages were used.

The final effluent turbidity was below 0.1 NTU for most of Run 20 with the most notable exception being when 5 mg/l of alum was being used in the chemical pretreatment. The rate of headloss development in the dual media filters was low (0.025-0.031 m/hr) and filter run lengths of up to 45 hours were easily obtained.

The use of 7.5 mg/l of alum and 0.05 mg/l of polymer produced satisfactory results whenever 25 or 50 mg/l of kaolin was used and the flow rate was between 12.2 to 24.4 m/hr. In Runs 21 through 24, coarse media filter headloss rates varied from 0.074 to 0.188 m/hr with the highest rates occurring at the highest flow rates. The turbidity removal in the coarse media filters during Runs 21 through 24 averaged between 83 and 98 percent. The dual media filters were not always used during these runs due to headloss limitations, however, it was assumed that if the coarse media filters performed satisfactorily in comparison to previous runs the dual media filters would also perform comparatively.

In the routine analyses during the laboratory pilot filter study, measurement was made of particle zeta potential. The zeta potential results are reported in Table 12 for the raw, rapid mix and coarse media filter effluent particles. These values represent the mean values of the particle zeta potential measurements taken during a filter run. During some of the filter runs,the alum dosage was changed and the particle zeta potential values are shown accordingly. Values were not reported for final effluent particle zeta potential since most of the values were highly variable and in most cases the particle concentration was so low that only individual particle zeta potential was recorded. In general, the range of the final effluent particle zeta potential could be expected to be anywhere between the highest and lowest value found in the raw water and the rapid mix unit.

By reviewing Figure 4, the design and operation diagram for alum coagulation summarized by Amirtharajah and Mills (3), one can conclude that particle destabilization during this study occurred by either sweep coagulation or a combination of sweep coagulation and adsorption. Adjustment of the alum dosages used in this study (5, 7.5, 10, 15 and 20 mg/l) to conform to the alum dosages as reported in Figure 4, would result in alum dosages of 4.5, 6.8, 9, 13.5 and 18 mg/l, respectively. The pH of the rapid mix suspension was maintained for the most part between 7.0 and 7.5 with the desired value being 7.0. Alum dosages of 15 and 20 mg/l would be oriented more towards sweep coagulation while lower alum dosages would fall within the range of destabilization by a combination of sweep coagulation and adsorption. The probability that destabilization

Table 12. Summary of particle zeta potential measurements during the laboratory pilot filter runs

Run #	Zeta Potential, mV: Raw Water Mean	Raw Water Standard Deviation	Raw Water (n)[a]	Rapid Mix Mean	Rapid Mix Standard Deviation	Rapid Mix (n)[a]	Coarse Media Effluent Mean	Coarse Media Effluent Standard Deviation	Coarse Media Effluent (n)[a]
2	-18.31	0.66	2	+6.79	0.62	2	-7.21	-	1
3	-17.65	0.49	4	-2.75	2.25	4	-5.33	3.75	4
4	-17.93	0.87	3	-1.07	0.50	3	-7.95	3.24	4
5	-17.52	0.69	3	-0.89	0.31	3	-8.51	0.98	3
6(10 mg/l alum)	-19.78	1.33	1	-16.16	0.36	1	-	-	-
6(15 mg/l alum)	-21.05	0.77	2	-0.40	4.17	3	-10.06	0.98	2
7	-20.24	-	1	-8.38	-	1	-10.93	-	1
9	-21.94	1.22	2	-0.31	0.62	4	-5.46	4.11	4
10	-20.15	0.49	2	+0.73	0.08	2	-6.22	6.40	3
11	-18.40	2.06	2	-4.98	0.69	2	-6.87	0.56	2
12	-18.86	1.50	5	-3.22	3.69	5	-4.58	2.90	6
13	-19.05	1.19	6	-6.42	1.59	6	-8.07	2.54	6
14	-21.79	0.37	2	+0.12	3.71	3	-3.21	3.06	3
15	-20.86	0.83	6	-7.09	4.04	6	-7.37	3.34	6
17	-16.76	1.21	3	-2.81	2.45	3	-4.35	3.15	2
18	-22.32	1.03	3	-2.51	3.25	4	-0.34	0.41	4
19	-21.99	1.49	5	-3.46	2.55	5	-5.33	2.39	5
20(15 mg/l alum)	-21.74	0.65	2	-3.42	4.84	2	-8.36	0.43	2

20(7.5 mg/l alum)	-21.63	2.31	5	-10.95	0.57		-11.03	1.42	5
21	-21.05	0.93	8	-11.80	1.79	8	-10.44	2.72	8
22	-21.20	1.34	6	-13.00	1.24	6	-11.98	1.67	6
23	-21.33	0.56	4	-11.80	1.13	4	-10.82	1.62	4
24	-19.04	0.33	2	-13.02	1.13	2	-9.67	0.87	2
25A	-21.55	0.94	2	-12.17	0.95	2	-12.01	0.02	2
25B	-21.35	0.80	4	0	0	4	-5.99	1.24	4
26	-20.51	1.29	3	-12.07	1.10	2	-8.71	0.0	2
27	-19.33	1.30	4	-12.60	1.99	4	-12.88	1.25	3

[a]Number of measurements.

occurred by sweep coagulation at 20 mg/l of alum was confirmed by the jar test results and by the appearance of some flocculated particles before the coarse media filters and in the coarse media filter effluent in the latter part of the breakthrough stage. At an alum dosage of 15 mg/l, flocculation prior to the coarse media filters was not evident and floc particles in the coarse media effluent, when they occurred, were smaller and appeared more dense as compared to when 20 mg/l of alum was used. At the 15 and 20 mg/l alum dosages, the particle zeta potential was at or near zero. A low or zero zeta potential is indicative of charge neutralization. When the raw water kaolin concentration was 100 mg/l a particle zeta potential at or near the isoelectric point produced, overall, the best filtration results.

At alum dosages below 15 mg/l, no apparent flocculation occurred before the coarse media filters. Particle destabilization at the lower alum dosages was probably less influenced by sweep coagulation than by adsorption. At the lower alum dosages, the particle zeta potential was more negative than at the higher dosages. Optimum filtration was obtained when the particle zeta potential was reduced from about -20 mV to within the range of -10 to -13 mV. Since the zeta potential was reduced, some charge neutralization was obviously occurring and the predominant adsorption mechanism was likely bridging in this case.

For the various chemical pretreatment schemes and kaolin combinations evaluated in this study, there was a relationship between particle concentration and the required alum dosage. At a kaolin concentration of 100 mg/l, effective filtration required a higher degree of charge

neutralization as evidenced by the rapid mix particle zeta potential results (Table 12) obtained in Runs 2, 3, 4, 5, 9, 11, 12, 13 and 25B using 15 or 20 mg/l of alum as compared to the results obtained in Runs 6, 15 and 25A using 7.5 or 10 mg/l of alum. When using 15 or 20 mg/l of alum, the coarse media filter run lengths were longer and a more acceptable filtrate was obtained compared to when 7.5 or 10 mg/l of alum was used. The use of 7.5 or 10 mg/l of alum when filtering 25 or 50 mg/l of kaolin (Runs 20, 21, 22, 23, 23 and 26), resulted in satisfactory coarse media filter run lengths and an acceptable filtrate quality. The rapid mix particle zeta potential results (Table 12) obtained in these filter runs indicated that at the lower kaolin concentrations a rapid mix particle zeta potential of about -10 mV was required. On the other hand, at 100 mg/l of kaolin the best results were obtained by the addition of 15 or 20 mg/l of alum which resulted in a rapid mix particle zeta potential near zero.

The above observations suggest that different particle removal mechanisms were required at different influent particle concentrations. When filtering 100 mg/l of kaolin, an alum dosage sufficient to produce almost complete charge neutralization was required. When filtering 25 or 50 mg/l of kaolin,an alum dosage sufficient to reduce the particle zeta potential from about -20 mV to about -10 mV was required. The success of the filter runs when the particle zeta potential was more negative (-10 mV) indicated that particle coagulant bridging may have been an important particle removal mechanism.

By design, the purpose of the nonionic polymer was to strengthen the alum floc and enhance particle removal in the filtration system.

The polymer obviously did improve the filtration system performance as seen from the results of Runs 1 and 2. One filter run, Run 16, was made to determine the extent of polymer-particle interaction when the particles received no alum treatment. The results of the run were intended to give some indication of the role of the polymer in the destabilization process. When the particles are first destabilized with alum, it was assumed that the interaction of the polymer was with the destabilized particles but it was not known if the addition of the nonionic polymer could serve as a primary destabilizer for particles not effectively treated with the alum. The use of 0.05 mg/l of Percol LT-20 produced somewhat surprising results. The coarse media effluent turbidity was 30 NTU or less for the first 8 hours and the dual media turbidity was less than 1 NTU during this period. Since a nonionic polymer destabilizes by bridge formation, these results indicated that this particular polymer was very effective. Its use as a sole coagulant would not be recommended because of the relatively unstable system that resulted as evidenced by a rapid and severe breakthrough of the final effluent turbidity. However, from the results of Run 16 it was concluded that if particles escaped treatment in the rapid mix there was a high probability that a large portion of these particles would be destabilized by the polymer.

Further evidence of the significance of the small amount of polymer providing a great deal of destabilization was supported by comparison of the results of Run 8 where no chemical pretreatment was used with Run 16. In Run 8, the raw water turbidity averaged about 82 NTU. The coarse media effluent turbidity averaged about 67 NTU with a standard deviation of 4.0 and the final effluent turbidity averaged about 33 NTU

with a standard deviation of about 5.0.

Filter operation

One primary concern in this study was to determine how best to operate the filter system in terms of whether constant rate or declining rate may have an inherent advantage over the other. When constant rate operation was used, the coarse media filters were both started at the same time. In most of the runs using declining rate, one coarse media filter was operated for the first 3 to 4 hours of the filter run and then the second filter was brought into service.

The declining rate mode resulted in the dirtiest filter receiving a reduced flow rate. When either rate mode was being used, it was standard procedure to reduce the flow to zero over approximately a ten minute period when a filter was to be backwashed. When the constant rate mode was being used, the amount of solids in each filter was approximately equal. Doubling the flow in one filter, even after carefully shuting off the other filter, resulted in stripping of solids from the filter. This was especially true at the higher kaolin concentrations. The effect of this high solids load to the dual media was not always evident from the data. During the initial stages of a dual media filter run, the dual media filters had a high capacity to take the increased solids load with minimal affect on the headloss and effluent turbidity. During the latter stages of a dual media filter run, the effect of backwashing a coarse media filter on the dual media headloss became much more evident. The high capacity of the dual filters to accept high solids loads initially was enhanced by the relatively low coarse media filter effluent turbidity

received by the dual media filters during the initial periods of the filter run. When the coarse media filters were operated in a declining rate mode, the amount of solids stripping when one filter was backwashed was visibly reduced and less effect was shown in the dual media filter headloss data.

In Runs 3 through 6, 9, 12, 25B, and 27, declining rate operation was used in the coarse media filters. In Runs 3 through 6, 9 and 12, this mode of operation was established by a staggered start up period. In the remaining runs, the declining rate was established by starting both coarse media filters at the same time and then backwashing one of the filters slightly before it was necessary. In Runs 25B and 27 both the coarse media filters and dual media filters were operated in a declining rate mode. Comparison of the headloss increases occurring in the dual media filters when the coarse media filters were backwashed during Runs 12 and 13 gives some indication of the potential problems associated with constant rate operation with only two filters. The largest increase in dual media filter headloss in Run 12 where a declining rate mode was used in the coarse media filters was 0.3 m, (Figure 25). In Run 13 at constant rate, the largest increase was 1.1 m, (Figure 28). As can be seen from the results of Runs 20, 23 and 24, the increase in dual media headloss was much less significant at the lower raw water turbidities when the coarse media filters were backwashed and the coarse media filters were operated at a constant rate.

The declining rate mode appeared to be a more stable type of operation when only two coarse media filters were used. The breakthrough of

the coarse media filter effluent turbidity was more gradual and usually less severe. With constant rate operation, a long period of good coarse media filter effluent turbidity was usually followed by a rapid and severe breakthrough.

In Runs 25B and 27 the coarse media filters and dual media filters were both operated in a declining rate mode. This was accomplished by backwashing one of each set of filters before necessary. This means of establishing a declining rate system proved to better represent the advantages of declining rate operation. Actually, this type of operation would more closely resemble an actual plant operation. Declining rate operation in both the coarse media filters and dual media filters showed the combined advantage of the previous constant rate and declining rate operations. Minimal increases in dual media filter headloss occurred when a coarse media filter was backwashed and longer coarse media filter and dual media filter runs were experienced with overall lower average effluent turbidities compared to previous runs using a declining rate operational mode in only the coarse media filters.

Other filter runs

Several filter runs were made during this study that are not interrelated to the other filter runs. These are Runs 7, 10, 17 and 19. Run 7 was made to determine whether 20 mg/l of alum in combination with NP-10, a nonionic polymer, could produce a satisfactory filter run. It was determined from jar tests using 100 mg/l of kaolin as the raw water particle concentration NP-10 was inferior to the other nonionic polymer,

Percol 720, in terms of producing a satisfactory settled water turbidity. The results of Run 7 confirmed the inferior performance of this polymer in the filtration process. The coarse media filter effluent turbidity was extremely high and was never less than 24 NTU. The final effluent turbidity was greater than 1 NTU during all but the first hour of the 11-hour run. Headloss development in the coarse media filters was almost nonexistent. The zeta potential of the kaolin particles in the rapid mix was more negative than in other runs using 20 mg/l of alum and could have been due to the nonionic polymer imparting a negative influence.

In Run 10, alum in combination with Cat Floc T was used to destabilize the particles. The results of this run were also inferior to the results obtained with alum and a nonionic polymer, Percol 720 or Percol LT-20. The particles were destabilized by charge neutralization as evidenced by the decrease in particle zeta potential after alum addition (Table 12). Very little headloss developed in the coarse media filters and the dual media filters received most of solids removal burden. The dual media filter effluent turbidity remained below 1 NTU for the first 11 hours of the run (Figure 23). Medium molecular weight cationic polymers, such as Cat Floc T, are not noted for there ability to destabilize by bridging. This may explain the lack of success in this run using this combination of pretreatment chemicals.

Run 17 was carried out using a constant rate operational mode and the influent suspension was 100 mg/l of calcium aluminosilicate. The results of this run were excellent. Influent turbidity was less than half that which resulted when a like quantity of kaolin was used. This may

have been due to the large size of the calcium aluminosilicate particles. The chemical pretreatment was 10 mg/l of alum, the optimum dosage determined in the jar test, in combination with 0.05 mg/l of Percol LT-20. Both the coarse media and dual media filter run lengths were 33 hours and could have been extended further but the filters were backwashed to prevent the possibility of problems occurring in removing the solids from the filter during backwash. This filter run in all respects was more stable than a filter run using kaolin with an equivalent raw water turbidity. The calcium aluminosilicate had a much higher cation exchange capacity than the kaolin. The zeta potential of the rapid mix particles was slightly negative so it is likely that both charge neutralization and bridging played a role in the removal of the calcium aluminosilicate particles in the filtration system. It should be considered, however, that a proportionately large quantity of calcium aluminosilicate would be represented by 10 to 20 NTU of turbidity when compared to the same turbidity value for kaolin particles.

Run 19 (constant rate operation) used 100 mg/l each of kaolin and calcium aluminosilicate in the raw water suspension. The pretreatment chemical combination was considered optimum for the removal of the kaolin particles. Run 17 had shown that a lesser chemical dosage was necessary for satisfactory filtration results when only calcium aluminosilicate was the raw water suspension particle. The results of this run indicated that the coarse media filter run length was shortened and the characteristic period of low coarse media filter effluent turbidity

between coarse media filter backwashes was absent. The coarse media filter effluent turbidity pattern was more like that of previous filter runs using a declining rate operation in the coarse media filters. The headloss development in the coarse media filters ranged between 1 and 1.4 m and in the dual media filters the headloss was about 1.4 m. The zeta potential of the rapid mix particles was slightly negative. Overall, this run could be considered successful and was representative of the highest turbidity that was filtered effectively.

Field Study

The field study was conducted to evaluate the series filtration system performance under conditions of high turbidity in a natural surface water. Due to the weather patterns in the area, high turbidity surface waters did not develop during the study period. The raw water turbidity during the study period averaged about 28 NTU and most of the turbidity was attributable to diatoms. Traditionally, the treatment of diatom ladden waters has presented problems to water treatment plant operators. The most significant problem has been the shortening of filter runs due to the rapid clogging of the filter media, particularly in single medium rapid sand filters.

In the laboratory filter runs, the direct filtration system was evaluated using a suspension prepared with kaolin clay, which can be classified as a hydrophobic particle. The diatoms would be classified as hydrophilic particles. The nature of the difference between these two types of particles could be responsible for the difference in the

type of treatment necessary for their removal in direct filtration. Since the diatoms are organic hydrophilic particles they would be expected to be more stable due to the higher degree of hydration, a different cation exchange capacity and an overall difference in the degree in which they may interact with a destabilizing chemical.

It was determined in the field study that the chemical pretreatment used successfully in the laboratory pilot filter study was not adequate to treat the diatom laden water to a final effluent turbidity of 1 NTU or less. The lowest final effluent turbidity obtained using alum and Percol LT-20 as coagulants was between 1 and 2 NTU.

Only limited particle zeta potential data were collected during the field investigation due to the inaccessibility of the instrument and due to the possibility of changes in zeta potential during the period that would be involved in transporting the samples. The average of two particle zeta potential measurements taken of the river water particles was -6.3 mV. When 40 mg/l of alum and 0.1 mg/l of Percol LT-20 were used, the zeta potential of the coarse media filter influent particles was zero. A zero particle zeta potential was also obtained in the coarse media filter influent when 15 mg/l of alum was used in combination with 7.5 mg/l of Cat Floc T. With the alum and LT-20 combination used for chemical pretreatment, the final effluent turbidity was between 1 and 2 NTU. Using the alum and Cat Floc T combination, final effluent turbidities less than 0.5 NTU were obtained.

Monscvitz et al. (64) reported on problems associated with diatoms in Lake Mead water when direct filtration was used without flocculation. They found that filter performance in regard to diatom removal was highly variable with respect to both diatom count and species. They also reported that as the filter run progressed the efficiency of diatom removal was lowered. In pilot studies, the use of flocculation and sedimentation prior to filtration greatly improved the efficiency of diatom removal. The additional benefit from sedimentation facilities, however, did not warrant their additional cost. Due to the diatom problem at Lake Mead, flocculation facilities providing 20 to 30 minutes of flocculation time prior to filtration were added to increase the water treatment efficiency.

Dostal and Robeck (32) were able to reduce diatom levels to acceptable levels when no flocculation facilities preceded filtration. The chemical pretreatment consisted of alum and the filter contained dual media. The highest raw water diatom level was 960 cells/ml and the turbidity was around 35 T.U.

Other investigators (20, 43, 44) indicated that diatom removal could be effective if flocculation preceded direct filtration. The use of alum with a filter aid resulted in effective treatment when coarser media filters were used. The coarser media filters alleviated the problems associated with short filter runs due to clogging when diatom populations become exceedingly high.

The short-term study demonstrated that the series filtration system was able to produce a filtered water turbidity of less than 0.5 T.U.

when used to filter a natural water with a raw water turbidity of 28 NTU resulting from a diatom population of 40,000 cells/ml. Under these conditions, the coarse media filter and dual media filter run lengths approximated 10 hours. The literature suggests that these results could be improved either by providing separate short-term flocculation facilities or by increasing the depth of the coarse media filter.

Photographs of Coarse Media and Particle Deposits

During this study a pattern in particle deposition within the coarse media filters was observed. Early in the filter cycle, the suspension particles appeared to coat the filter media. As the filter run progressed, the particles appeared to deposit on top of the media grains in small mounds. Subsequent filtration resulted in further mounding of the suspension particles on top of the media which in turn either closed a pore space within the filter or led to an area of constricted flow. Channelization became quite obvious in the upper portions of the coarse media filters during the latter portion of a filter cycle. This effect was most noticeable at the 100 mg/l concentration of kaolin. At this kaolin concentration, channelization was observed in the entire upper half of the filter at run termination. When 25 mg/l of kaolin was used, only the upper portions of the coarse media filters showed evidence of channel formation. At first this pattern was thought to be due to wall effects. Observation of the surface of the coarse media filters showed a pattern of covered and uncovered media grains. The uncovered grains were at the entrance points to the channels within the filter. However,

channel entrance points were no more prominent near the wall than in the center of the filter.

Tien et al. (92) discussed particle deposition within a deep bed filter in the development of a dynamic model for filtration. They viewed particle deposition within a filter as a two stage process. Simplified, the first stage was dominated by deposition according to a smooth coating mode. The second stage was characterized by the prevalence of constriction clogging.

Raga and Kalsi (76) observed the mounding of particles on the filter media grains. They used 5 mm glass beads to filter high concentrations of flocculated and unflocculated kaolin particles. These investigators observed dislodgements of particles from the mounds of deposits when impacted by particles in suspension. Detachment of the particles became more intense as the quantity of deposit increased.

A recent study at Iowa State University which examined particle (kaolin or iron floc) deposition within a sand filter media, using a scanning electron microscope, was conducted by Stammer (85a). He found that random particle deposition on the media grains occurred during the initial stages of a filter run and was associated with the formation and growth of dendrites and clusters of particles. This type of particle deposition led him to conclude that diffusion and hydrodynamic action are the predominant filtration mechanisms during the early stages of a filter run. In addition to these two mechanisms interception and inertia may aid in the growth of chain and dendrite structures. During the latter

stages of a filter run Stammer (85a) found that dense masses of particles formed on the tops of media grains and in crevices between the media. In addition, smaller particles continued to coat the media as during the early part of a filter run. He concluded that gravity and straining mechanisms influence the growth of deposits of dense particles. As the filter run progresses particle deposits become more compact and less porous.

The initial ripening period during filtration was explained by Stammer to result from the importance of particle-particle collection and attachment in the filtration process. Particle removal during filtration increases as the filter media is coated with particles which in turn interact with particles in the suspension. The type of deposit within the filter was also found to vary with the type of suspension particle. The low density iron floc used in the experiments by Stammer exhibited the characteristic media coating during the early stages of a filter run but the mounding of these particles on the top of the media grains was less pronounced than when the higher density kaolin particles were filtered.

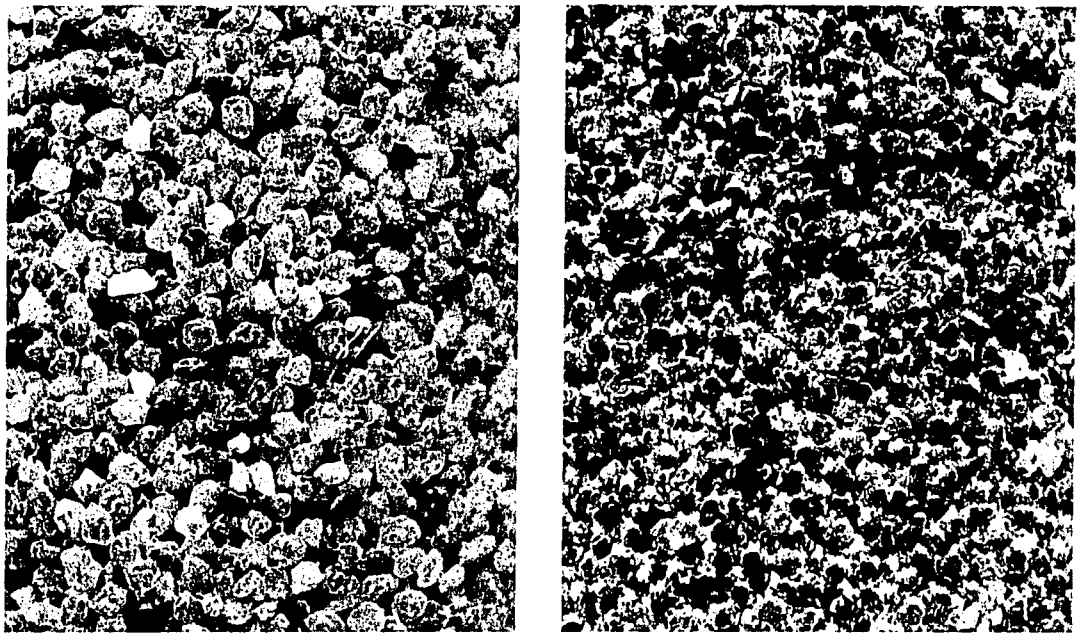
The size of the coarse media used during this study was large enough to permit visual observation of particle deposition within the filter bed. Therefore, photographs of the coarse media and particle deposits were taken at various stages during filter Runs 26 and 27. The kaolin concentration in Runs 26 and 27 were 25 and 100 mg/l, respectively. The filter media depth was 132 cm. Photographs were taken at the media surface and

at various depths beneath the surface. Photographs identified as "top" show the media surface and approximately the top 7 cm of media. Photographs identified as "upper", show the media and deposits approximately 12 cm from the surface. "Middle" photographs show the media and deposits at a depth of approximately 65 cm, while "lower" photographs show the media and deposits at a depth of about 95 cm.

The progression of particle deposit within the coarse media filters in Run 26 is shown in Figures 51 and 52. Figure 51 shows the upper portion of a coarse media filter through the 30 hours of filter run. These photographs illustrate the two stages of filtration as described by Tien et al. (92).

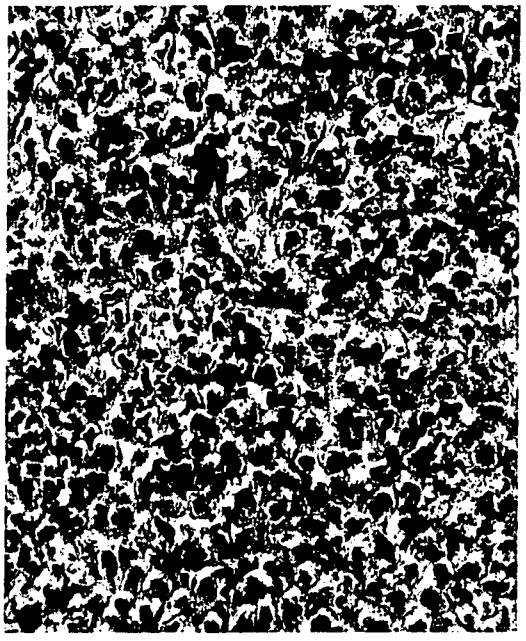
The photograph after 4 hours of filtration shows the filter media as it is being coated by particles. The photograph after 7 hours, shows the development of mounds or domes on the filter grains. The photographs after 15 hours show the constricted flow as evidenced by the development of channels within the media.

Figure 52 shows the media and deposits in the "middle" of the filter after 15 and 30 hours. After 15 hours, this portion of the filter showed the development of the mounds of deposit similar to those in the upper portion of the filter at an earlier time (after 7 hours in Figure 51). By the middle of the filter run, the deposit in the upper portion of the filter has gone from coating the media to a constricted flow pattern. The deposition in the middle portion of the filter now acts in the same way that the upper portion did earlier in the run. In other words, as

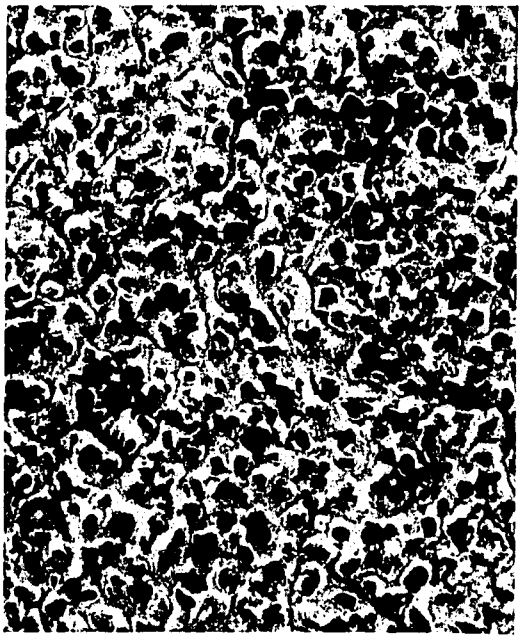


AFTER 4 HRS.

AFTER 7 HRS.

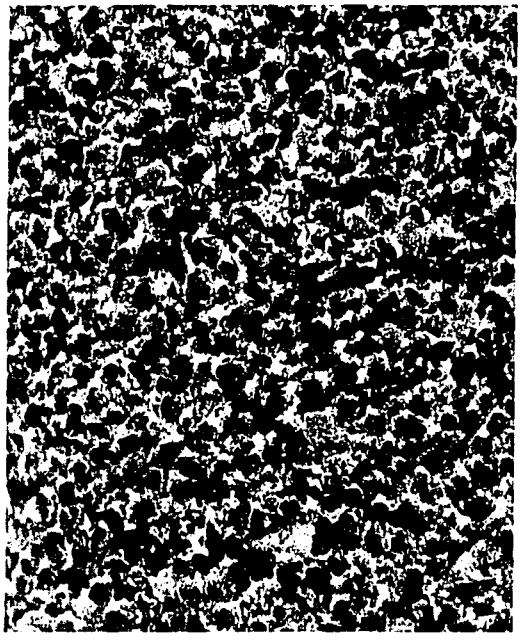


AFTER 15 HRS.

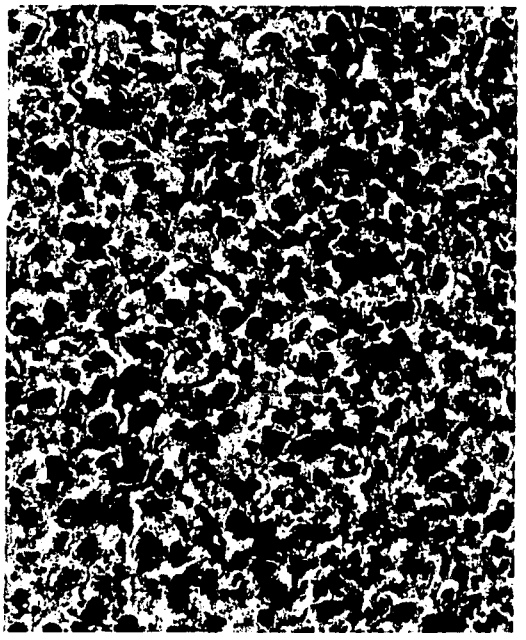


AFTER 30 HRS.

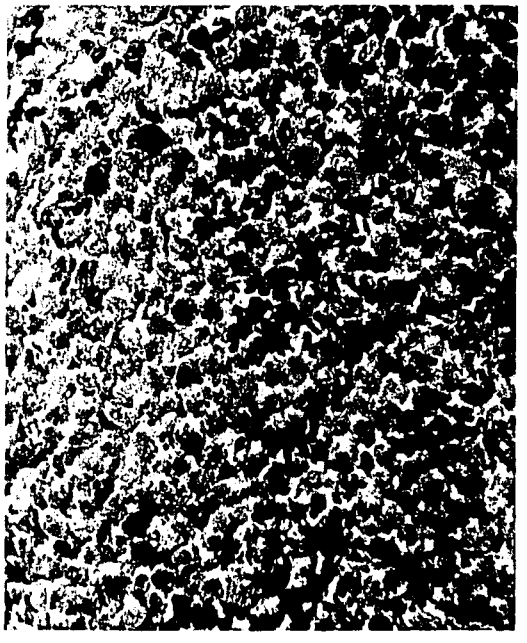
Figure 51. "Upper" photographs of coarse media in filter 2 after 4, 7, 15 and 30 hours of filtration in Run 26 (25 mg/l of kaolin, 18.3 m/hr filtration rate)



MIDDLE AFTER 15 HRS.



MIDDLE AFTER 30 HRS.



LOWER AFTER 30 HRS.

Figure 52. "Middle" and "lower" photographs of coarse media in filter 2 after 15 and/or 30 hours of filtration in Run 26 (25 mg/l of kaolin, 18.3 m/hr filtration rate)

the filter run progresses, the burden of particle removal shifts lower into the filter bed.

In Figures 51 and 52, the "upper, middle and lower" portions of the coarse media are shown after 30 hours of operation at which time the run was terminated due to headloss limitations. The development of flow channels in the upper and middle portions of the filter are clearly evident. The characteristics of the deposit shown in the middle after 30 hours of filtration are similar to those in the upper part of the filter midway through the filter run. The photographs of the lower portion of the filter at run termination shows deposits appear in mounds and no channelization is readily evident. The deposits shown are similar to those shown in the photographs of the upper media after 7 hours and of the middle media after 15 hours.

The figures clearly show the morphology of the deposits within the coarse media filter. Comparison of the photographs taken at different times and different depths illustrate the change in the deposit within the filter and indicate that the burden of removal of the suspended solids must shift deeper into the filter as the run progresses.

The photographs in Figures 53, 54 and 55 show the progression of the deposit formation from 100 mg/1 kaolin in the coarse media during Run 27. After only 4 hours, channels develop in the upper portion of the filter and mounds of deposit are evident in the middle and lower portions of the filter (Figure 53). At a kaolin concentration of 100 mg/1, the morphology of the particle deposit changes more rapidly than it does

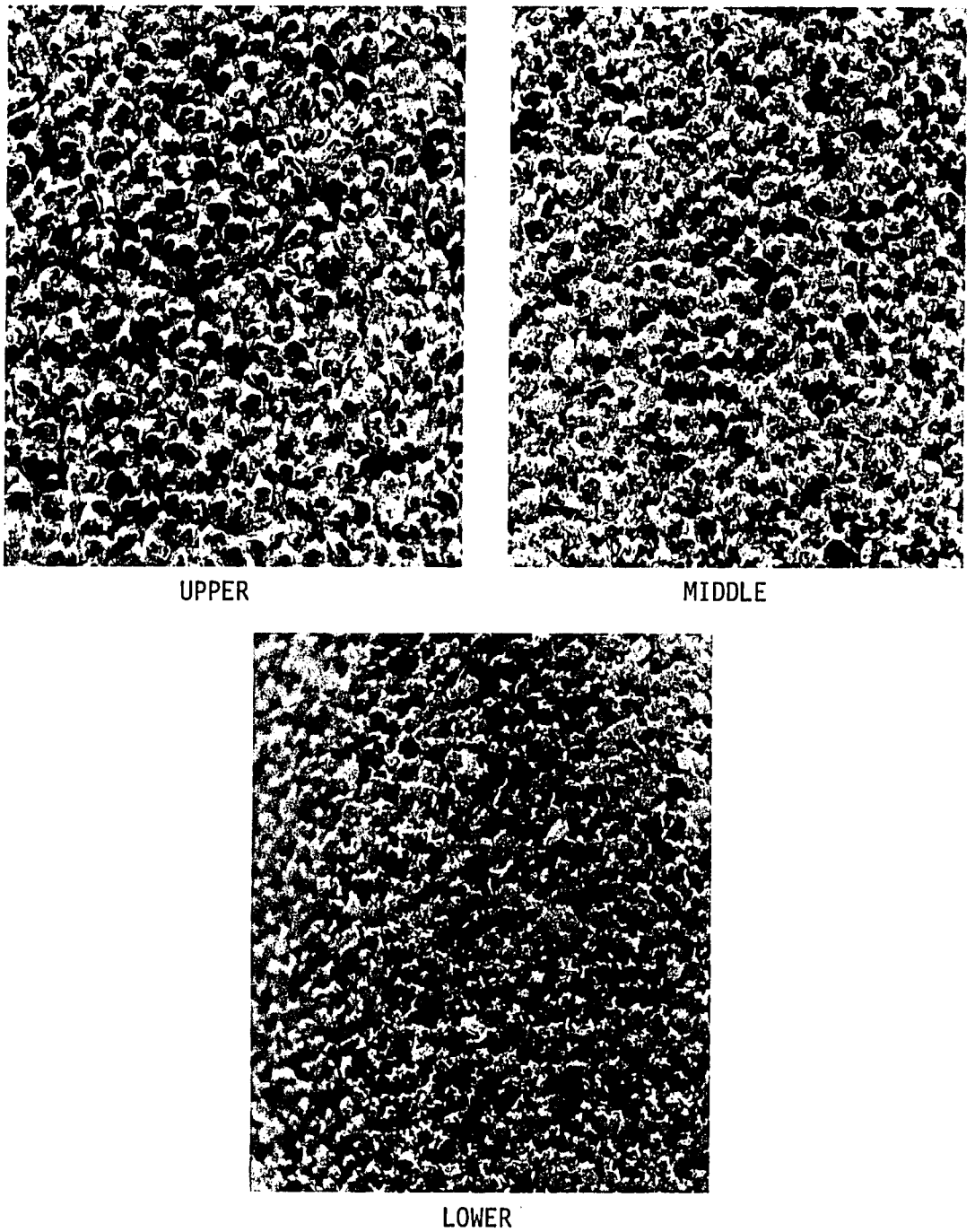


Figure 53. Photographs of coarse media in filter 2 in Run 27 after 4 hours of filtration (100 mg/l of kaolin, 12.2 m/hr filtration rate)

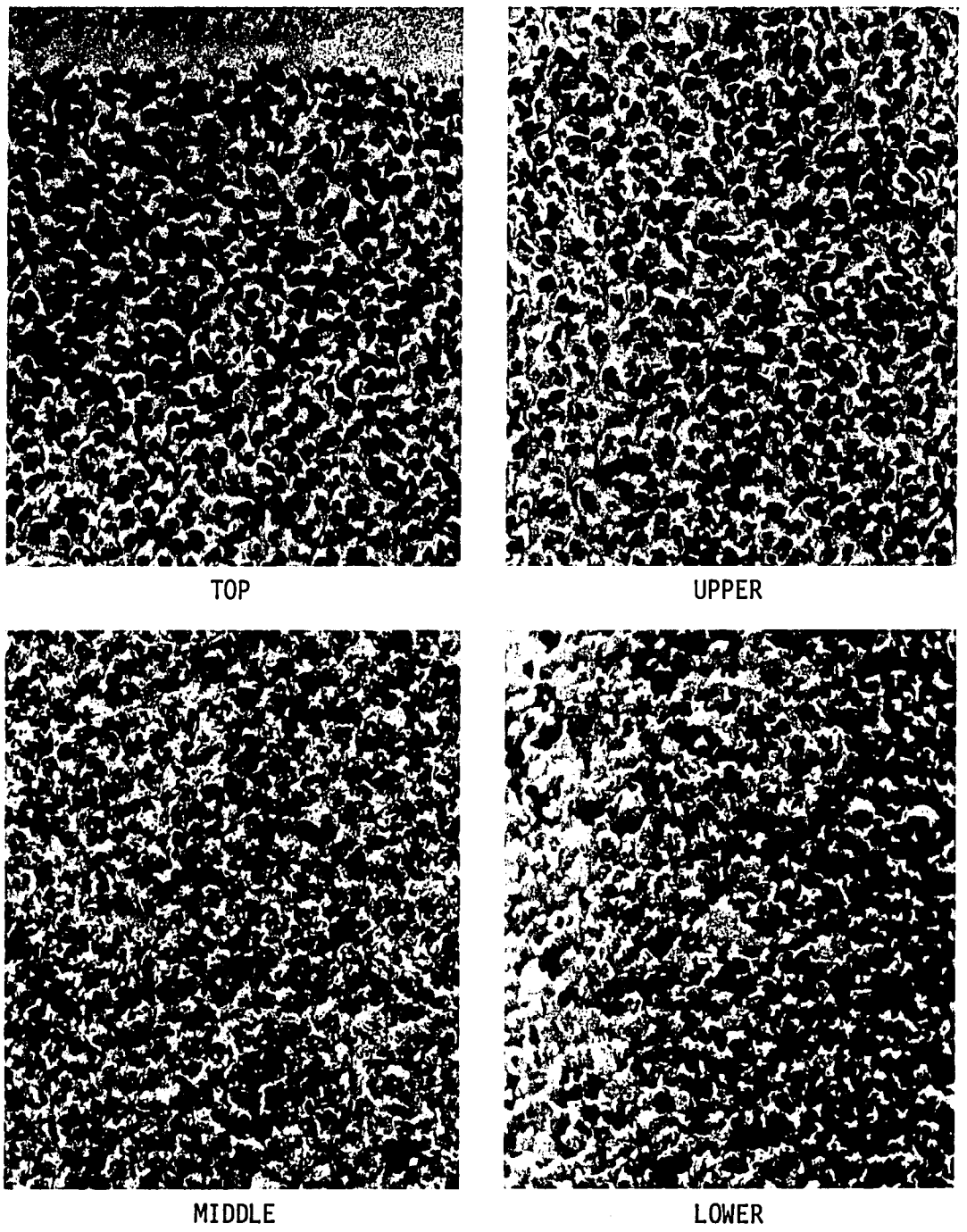


Figure 54. Photographs of coarse media in filter 2 in Run 27 after 10 hours of filtration (100 mg/1 of kaolin, 12.2 m/hr filtration rate)

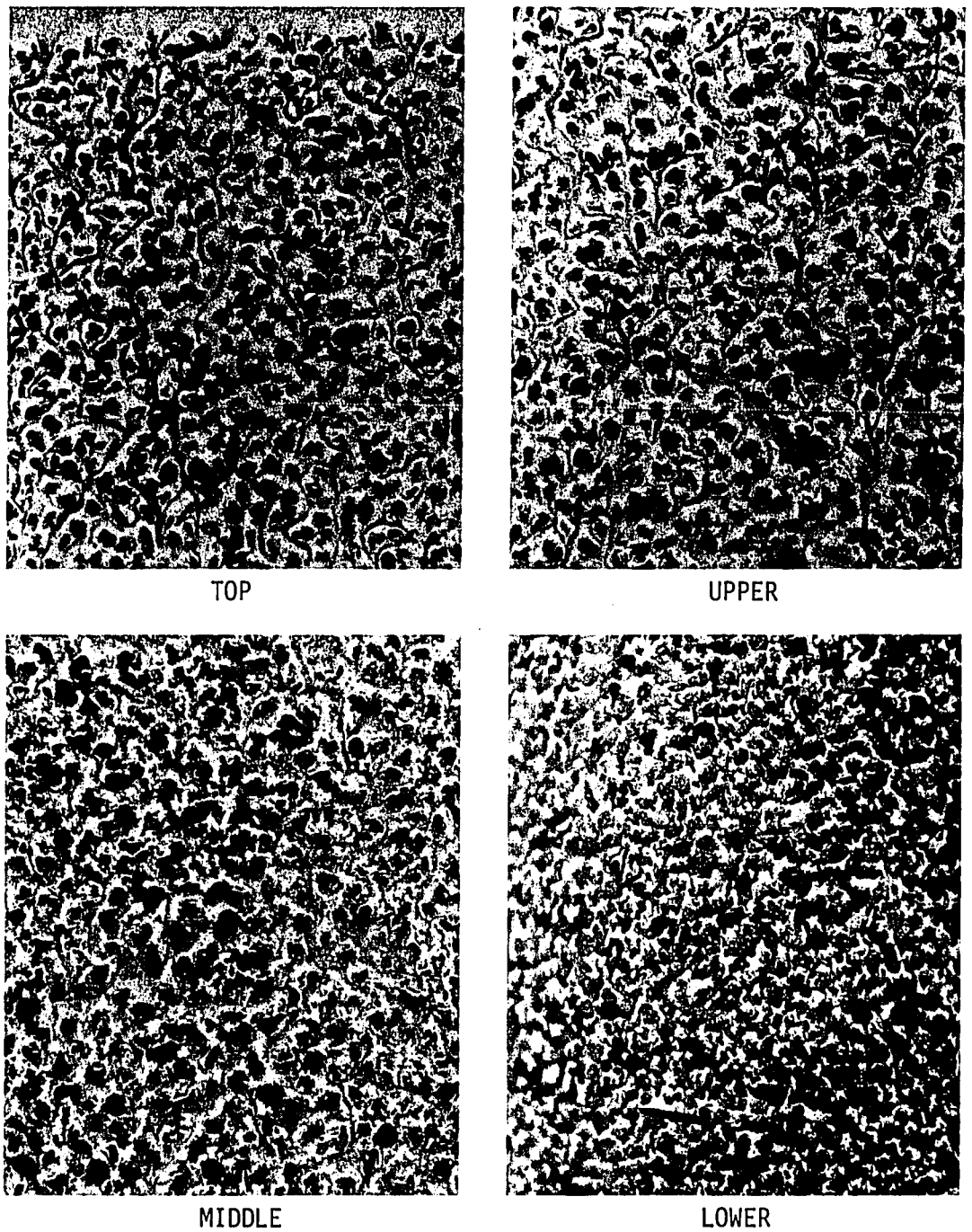


Figure 55. Photographs of coarse media in filter 1 in Run 27 after 15 hours of filtration (100 mg/l of kaolin, 12.2 m/hr filtration rate)

when only 25 mg/1 of kaolin is present. When compared to the use of the 25 mg/1 kaolin concentration, the burden of solids removal not only shifts more rapidly but extends throughout more of the filter bed depth during the early portion of the filter run. This observation supports the need for deeper filter beds of coarse media under high turbidity conditions.

The photographs in Figure 54 show the filter media and deposits after 10 hours of the filter run. The "top" photograph illustrates the channel openings which form at the surface of the media. The "upper" photograph shows that the amount of deposit has clearly increased and the channels have become more obvious. In the "middle" photograph, the amount of deposit in the filter has increased compared to the "middle" photograph in Figure 53 and channels have started to form. The "lower" photograph shows the build-up of deposit in the lower portion of the filter.

The photographs in Figure 55 show the filter media and deposits at the end of the run, terminated due to turbidity breakthrough. It should be pointed out that these photographs are not of the same filter as the previous photographs but are representative of the pattern of deposit that occurs. The other filter had very similar deposit patterns at run termination. The photographs show that significant amount of deposit and channel formation occurred in the upper and middle portions of the media. In the lower portion of the filter, the onset of channel formation can be noted. The photographs clearly show the flow lines in the coarse filter media. It is interesting to note that the channels are not only oriented

vertically but that in some instances many horizontal channels are formed.

The photograph in Figure 56 shows the surface of a coarse media filter. Some problems developed when draining the filter so that a photograph could be taken. The large areas showing complete coverage of the surface with deposit were absent just prior to draining the filter. The areas not covered by the surface deposit appeared to be the entrance points to the channels developed within the filter. Although a deposit was noticeable on the surface of the coarse media, the large media size and pore openings prevented significant particle removal by surface straining.

In summary, several observations concerning particle deposit within the coarse filter media can be cited. The pattern of deposit was consistent throughout the study in filter runs using kaolin and are represented by photographs of the coarse media and deposits taken during Runs 26 and 27. The observations are as follows:

1. During the initial periods of the filter run, the removed particles coated the media grains. The length of this period was inversely proportional to the influent kaolin concentration. When 100 mg/l of kaolin was used, the coating period was almost nonexistent in the upper portion of the filter.

2. Subsequent to the period of coating the media, mounds of deposits built up on the top of the media grains. This can be looked upon as a transitional phase between coating of the media grains and the development of constricted flow with the formation of channels.

3. At both kaolin concentrations, there were periods during the filter run where coating, mounding and channelization were simultaneously

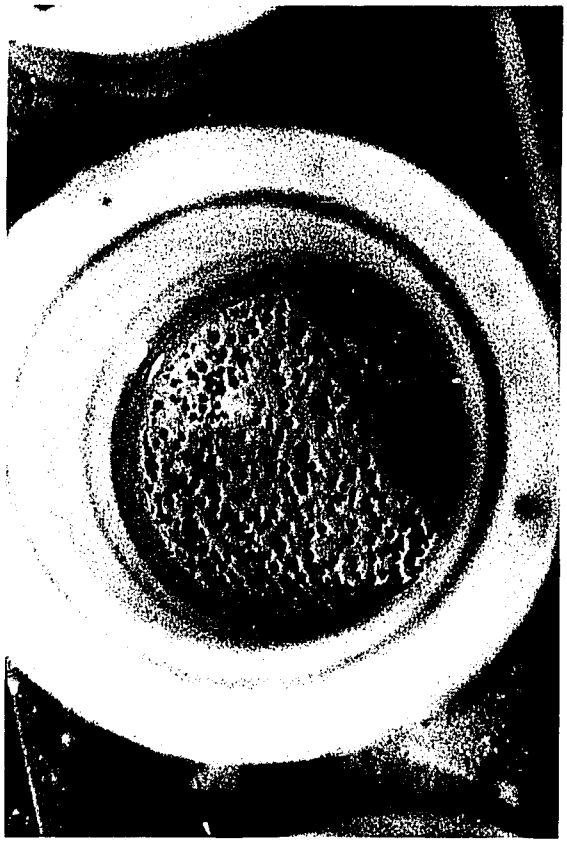


Figure 56. Top view photograph of the coarse media surface of filter 1 in Run 27 after 15 hours of filtration (100 mg/l of kaolin, 12.2 m/hr filtration rate)

occurring in the filter media.

4. In Run 26, at the low kaolin concentration when the run was terminated due to headloss limitations, channelization occurred to the mid-depth of the media and mounding occurred in the lower portions of the media.

5. In Run 27, at the high kaolin concentration when the run was terminated due to turbidity breakthrough, channelization occurred to the mid-depth of the filter and was becoming evident in the lower portion of the filter.

6. At the higher kaolin concentration, the burden of solids removal shifted deeper into the filter sooner and at a higher rate than it did with the lower kaolin concentration.

7. In both filter runs, an obvious change in the deposit morphology occurred as the run progressed and the solids removal extended deeper into the filter.

8. Channel openings could be observed on the surface of the coarse media filter.

MATHEMATICAL MODELS

Description and Application

Numerous mathematical models have been developed to describe the filtration process since Iwasaki first proposed his filtration equations in 1937 (48). The development of a mathematical model relies on the selection of a mass balance equation and a kinetic equation to describe the filtration process and the boundary conditions and assumptions which simplify the solutions to the filtration equations. Virtually every filtration model proposed to date can not model exactly the experimental operating results of pilot filters. In almost all filtration models, the major discrepancies between experimental results and theoretical predictions occur during the early stages of a filter run during the filter ripening period and during the period of constant good water quality. Recently, two filtration models have been developed by investigators at Iowa State University. The first model developed by Saatci (80) and Saatci and Oulman (81) was based on the similarities between filtration and the design of granular activated carbon columns. The bed depth service time (BDST) method based on the Bohart Adams equation for adsorption of chlorine on activated carbon was used to develop a model for deep bed filters. The BDST equation relies on the fact that the time to breakthrough for a given effluent solids concentration is dependent on the depth of the filter, the rate of filtration and the concentration of solids in the influent. Further simplification of the Bohart-Adams equation results in an equation for a logistic or "growth" curve. This type of curve is an S-shaped curve that is

symmetrical about the midpoint. In logarithmic form the equation is linearized which results in a practical method for determination of filter attachment and storage coefficients. A linear regression analysis of the logarithmic data is used in the determination of the filter coefficients. Once the regression line is determined, the theoretical breakthrough curve can be calculated. As would be expected, the theoretical and experimental results agree more closely during the turbidity breakthrough stages of a filter run. With this model, once the storage coefficient and attachment coefficient are determined, filter performance at different flowrates, media depths and filtration times can be estimated to evaluate a filtration system design. Filter coefficients can be determined from experimental data collected from filters of various depths or from data from a single depth filter. The data from a single media depth filter are considered less accurate for determination of the filter coefficients.

A filtration model proposed by Saleh (82) was based on the same mass balance equation as used by Saatci, but a different kinetic equation to describe the behavior of deep bed filtration. The most notable difference between the two kinetic equations was the incorporation in Saleh's equation of terms to describe the change in hydraulic gradient changes as particles are collected in the filter and the permeability decreases. Filtration models not incorporating the hydraulic gradient change can be used successfully in sanitary engineering applications as demonstrated by Saatci's model and by

Saatci's (80) analysis of Adin's model, described by Adin and Rebhun (2), with the hydraulic gradient term equal to unity. The reason for the success is that in sanitary engineering applications filter runs are relatively short in comparison to the time necessary to use the ultimate solids holding capacity. Filter runs are usually terminated after a specific time period, headloss development or final turbidity value.

Both Saatci's and Saleh's filtration models were applied to several filter runs made during this study. In fact, Saleh used coarse media effluent turbidity data collected during 8 filter runs of this study to confirm his filtration model. The use of either filtration model was limited to filter runs using a constant filtration rate during the filter run. The nature of the operation of the depth clarifiers used in this study proved to be almost ideal for modeling the operation of these filters. Normally, a short ripening period was encountered followed by a short period of relatively constant turbidity before the turbidity breakthrough started.

The first step in modeling a filter run by Saatci's method is to determine C/Co values at various times. The values for the filter coefficients are determined by plotting - ln (Co/C - 1) vs. filtration time and performing a linear regression analysis to determine the slope and intercept. Some discretion has to be used to delete points for the regression analysis which could be considered outliers and points representative of a long ripening period and the period of constant water quality. For the analysis in this study the selection

of points for the linear regression analysis proved to be relatively easy. From the slope of the regression line, an estimate of the attachment coefficient is found by the relationship that the product of the attachment coefficient and the influent solids concentration is equal to the slope.

$$\text{Slope} = \beta = K_1 C_o \tag{1}$$

$$F_o = \frac{-\ln Ri \; Va}{K_1 X} = \text{storage coefficient} \tag{2}$$

where

ln Ri = regression line intercept, time = 0

V_a = approach velocity

K_1 = attachment coefficient

X = filter depth.

The model prediction is then found by the following relationship.

$$C/C_o = \frac{1}{e^{K_1 (F_o x/Va - C_o t)} + 1} \tag{3}$$

The filtration model can be used to estimate the filter performance beyond the time used in experimental filter runs. Also, by rearranging equation 2 for the filter storage coefficient, the effects of varying flow rate or media depth can be estimated. Extrapolation of the data from the model should be viewed with caution since both the attachment coefficient and storage coefficient may vary with changes in flowrate and influent particle concentration (81).

The use of Saleh's model as related to this study involves the determination of C/C_o by equation 4.

$$C/C_o = \left[\frac{e^{bt} + 1}{2e^{ax} + e^{bt} - 1} \right]^2 \tag{4}$$

where

$b = 8\ K_o C_o$

K_o = attachment coefficient

C_o = influent solids concentration

t = time

x = filter media depth

$a = K_o\ \sigma_u / 2\ Vs$

σ_u = ultimate specific deposit

V_s = superficial flowrate.

Initial values of "a" and "b" are determined by equations 5 and 6.

$$a = \ln \sqrt{C_o/C_i} \tag{5}$$

$$b = \ln \frac{2\ C_o/C_i - 1 + C_o/C_f}{C_o/C_f - 1} \div t_f \tag{6}$$

where

C_i = the initial effluent solids concentration

C_f = the final effluent solids concentration

t_f = the filter run length.

If the "a" and "b" values determined from equations 5 and 6 do not produce a breakthrough curve which reasonably fits the experimental data, the values can be adjusted by trial and error for a better fit. The value of "a" predominates at the beginning of a run and raising the "a" value will decrease the initial effluent concentration. The value of "b" predominates at the end of a run and increasing "b" will increase the effluent concentration and therefore shorten the filter run.

Saleh used the trial and error method on the data from Runs 13, 14, and 18 of this study to model the filter effluent characteristics. He also used the trial and error method on the data from Run 20 and extrapolated the results to predict filter performance at different kaolin concentrations and filtration rates in Runs 21, 22, 23, and 24. The operating conditions for all the filter runs can be found in Table 9. In Runs 20 through 24, the same chemical pretreatment was used but the influent solids concentrations and flowrates were not always the same. The method of extrapolation involved adjusting the value of "a" in proportion to the $(\text{flowrate})^{-1}$ and the value of "b" in proportion to the influent solids concentration. Table 13 gives the values of "a" and "b" as determined by Saleh for filter runs 13, 14, 18, and 20, and Table 14 gives the values of "a" and "b" for runs 20, 21, 22, 23, 24, and 25A.

Run 25A was made to test the validity of extrapolating the results of a filter run to another filter run with a higher influent solids concentration. Prior to Run 20 it had been determined that approximately

Table 13. Values of "a" and "b" as determined by Saleh (82) for filter runs 13, 14, 18 and 20

Run #	$\frac{C_i}{C_o}$	Initial Value of "a" ($inch^{-1}$)	t_f (hr)	$\frac{C_f}{C_o}$	Initial Value of "b" (hr^{-1})	Final Value of "a" (hr^{-1})	Final Value of "b" (hr^{-1})
13	0.01	0.04428	14.5	0.48	0.25371	0.04428	0.25491
14	0.31	0.01126	3.5	0.73	0.6061	0.010956	0.59451
18	0.12	0.02039	3.5	0.41	0.4982	0.020475	0.50458
20	0.005	0.051	20	0.33	0.177	0.0684	0.20262

Table 14. Values of "a" and "b" as determined by Saleh (82) for filter runs 20, 21, 22, 23 and 24 and by Saleh's method for 25A

Run #	(gpm/ft^2)	Kaolin, (mg/l)	"a" ($inch^{-1}$)	"b" (hr^{-1})
20	5	50	0.0684	0.20262
21	5	25	0.0684	$0.20262(\frac{25}{50}) = 0.10131$
22	7.5	25	$0.0684\ (\frac{5}{7.5}) = 0.0456$	$0.20262\ (\frac{25}{50}) = 0.10131$
23	7.5	50	$0.0684\ (\frac{5}{7.5}) = 0.0456$	0.20262
24	10	25	$0.0684\ (\frac{5}{10}) = 0.0342$	$0.20262\ (\frac{25}{50}) = 0.10131$
25A	5	100	0.0684	$0.20262\ (\frac{100}{50}) = 0.40524$

Table 15. Results of the analysis of filter runs 13, 14, 18, 20, 23 and 25A by Saatci and Oulman's (81) linearized logistic curve method

Run #	C/C_o	n	Slope hr^{-1}	Intercept	r^2	k_1 $(NTU)(t)^{-1}$	Fo NTU	Filtration Rate m/hr	Kaolin Concentration mg/l
13	0.01	8	0.342	-4.70	0.99	0.0042	10403	12.2	100
14	0.35	4	0.472	-6.03	0.97	0.0021	2575	12.2	300
18	0.14	4	0.529	-1.87	0.99	0.0035	4938	12.2	200
20	0.006	10	0.204	-5.21	0.98	0.0055	8778	12.2	50
23	0.01	14	0.272	-4.54	0.97	0.0074	8519	18.3	50
25A	0.15	11	0.214	-1.77	0.98	0.0028	5776	12.2	100

15 mg/l of alum and 0.05 mg/l of Percol LT-20 would produce a reasonable filter run length when the influent solids concentration was 100 mg/l of kaolin. Reducing the alum concentration to 10 mg/l shortened the filter run. Run 25A was made with 100 mg/l of kaolin as the influent solids concentration and the chemical dosages were 7.5 mg/l of alum and 0.05 mg/l of Percol LT-20, the same chemical dosages as used in Run 20.

The results of Saleh's model predictions for Runs 13, 14, and 18 are shown in Figure 57. The model predictions compare favorably with the experimental results. The comparisons between the experimental results and the model predictions for Runs 20, 21, 22, 23, 24, and 25A using Saleh's model are shown in Figure 58. The model predictions fit the experimental results for Runs 20, 21, 22, 23, and 24. The model predictions for Runs 21, 22, 23, and 24 are from the extrapolation of the model prediction for Run 20 as determined by the trial and error method. The comparison between the theoretical prediction and the experimental results for Run 25A is also shown in Figure 58. The results predicted for Run 25A by extrapolation of the data of Run 20 by Saleh's method poorly describe the actual filter performance. Although the conditions of Run 25A were not repeated in another filter run, it was felt the experimental results are reliable and can be substantiated by the higher chemical requirements of preceding filter runs using 100 mg/l of kaolin. The poor fit of the extrapolated results supports the warning by Saatci and Oulman (81) that extreme care must be exercised in

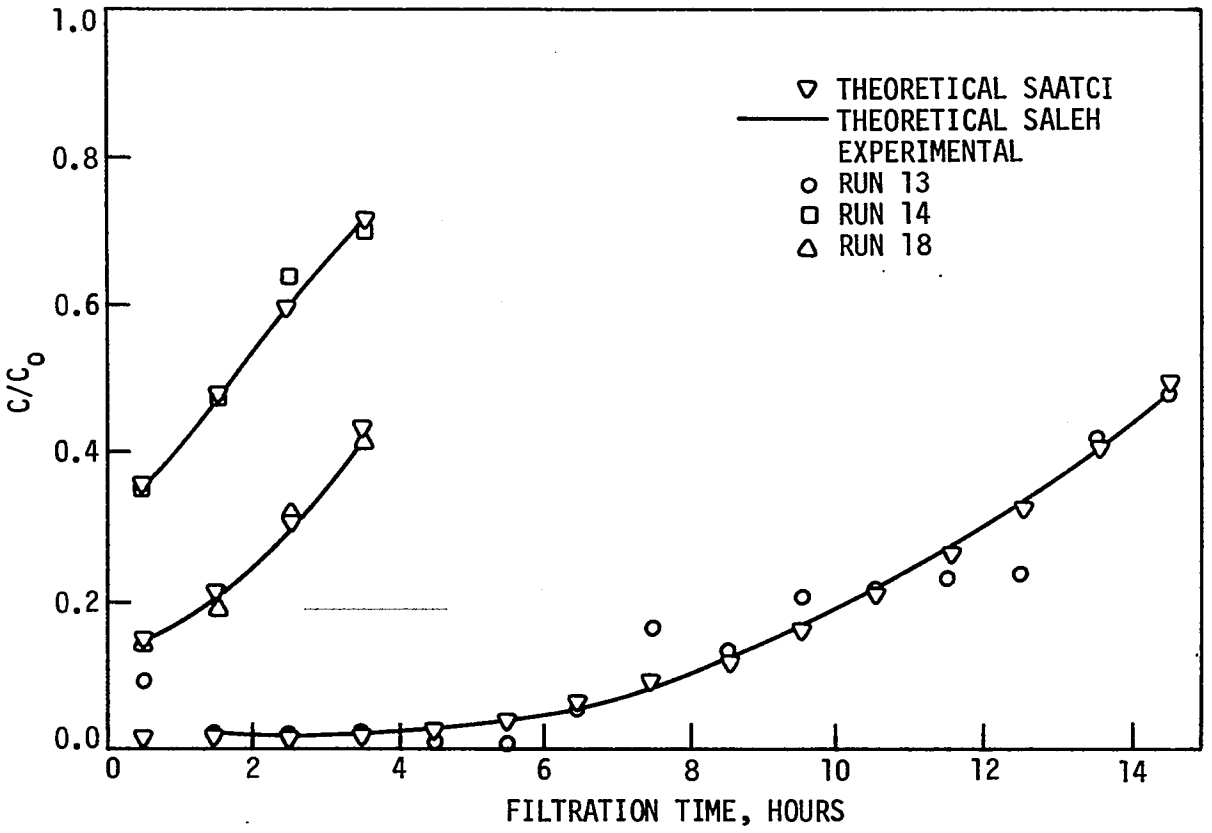


Figure 57. Comparison between the experimental and theoretical breakthrough curves for Runs 13, 14 and 18

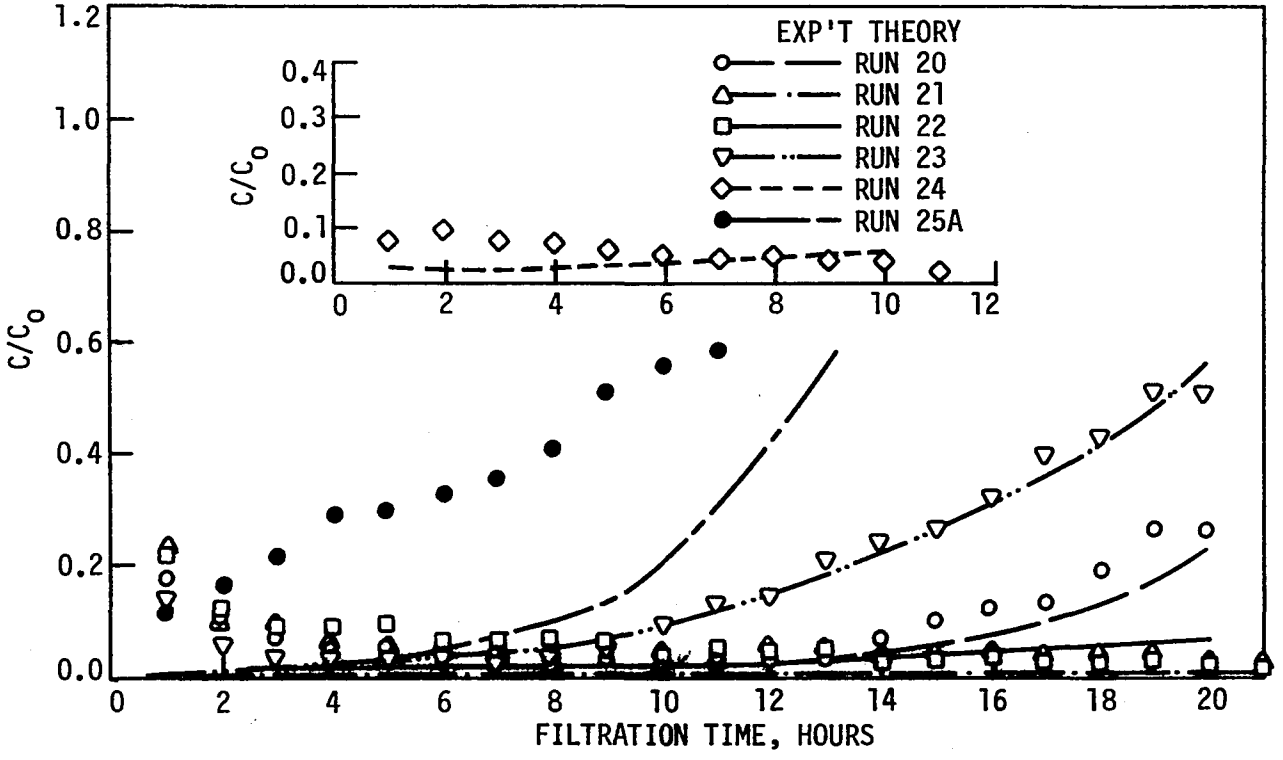


Figure 58. Comparison between the experimental and theoretical breakthrough curves for Runs 20, 21, 22, 23, 24 and 25A

applying the results of one filter run to another. Both the attachment coefficient and storage coefficient may vary with filtration rate and influent solids concentration. Further evidence of the shortcomings of the extrapolation procedure is evident from the results of the evaluation of Saatci's filter runs by Saleh (82).

Saatci and Oulman's (81) filtration model was used to predict the filter performance in Runs 13, 14, 18, 20, 23, and 25A. This model could not be used to predict the filter performance in Runs 21 and 22 since turbidity breakthrough did not occur and since the model predictions can not be extrapolated from one filter run to another filter run having a different influent solids concentration. The results of the regression analysis for Runs 13, 14, 18, 20, 23, and 25A using Saatci and Oulman's (81) filtration model are given in Table 15. The comparisons between the model predictions and the experimental results in Runs 13, 14, and 18 are shown in Figure 57. It can be seen in Figure 57 that the model fits the experimental data and gives essentially the same theoretical results as Saleh's model.

The theoretical predictions using Saatci and Oulman's (81) filtration model for Runs 20, 23, and 25A are shown in Figure 59. The theoretical predictions compare favorably to the experimental data. Also, shown in Figure 59 is the theoretical prediction for Run 23 determined from the extrapolation of the regression analysis of Run 20. In Run 20 the filtration rate was 12.2 m/hr and in Run 23 the filtration rate was 18.3 m/hr. The extrapolation of results using Saatci and

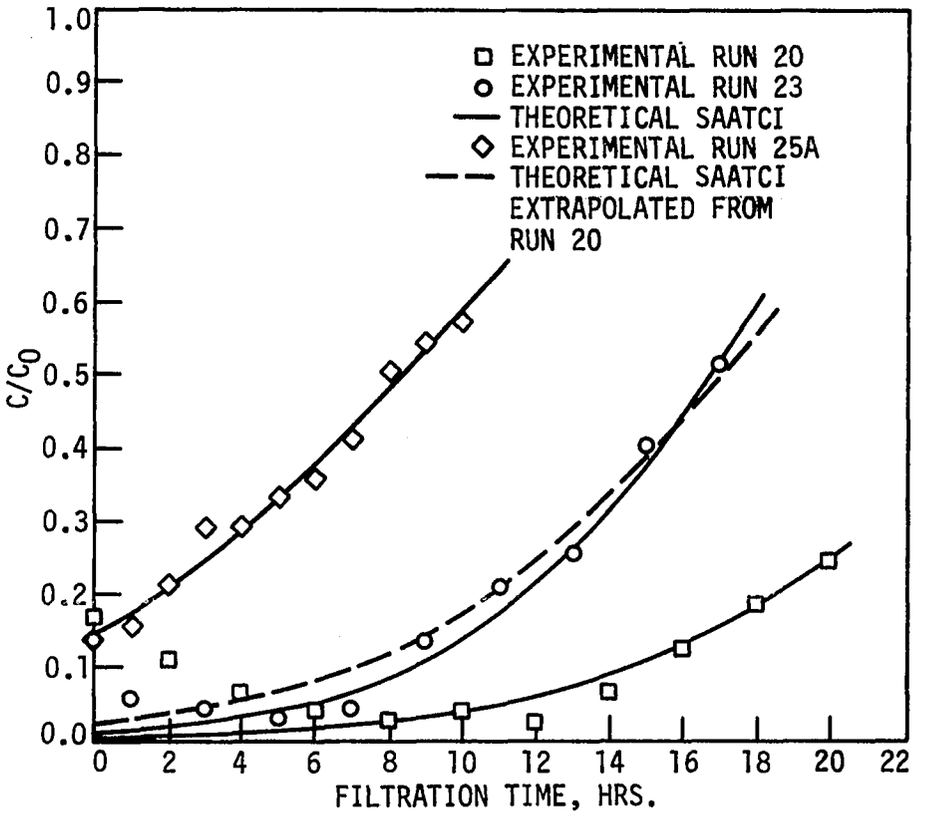


Figure 59. Comparison between the experimental and theoretical breakthrough curves for Runs 20, 23 and 25A by Saatci's method

Oulman's model involves using the attachment and storage coefficients for one run and substituting the appropriate approach velocity into equation 3 to determine the breakthrough curve. The extrapolated results shown in Figure 59 for Run 23 compare favorably to the experimental results.

Summary

The use of either Saatci's filtration model or Saleh's filtration model with the trial and error procedure for determining "a" and "b" produced good agreement with several filter runs performed during this study. The best agreement of the models with the experimental data occurred during the latter stage of a filter run and in filter runs exhibiting no extended ripening period or constant quality effluent. The use of the filtration models to extrapolate filtration results to different operating conditions is a risky process. Extrapolation of results should be limited to estimating filter operating conditions similar to those from which the attachment and storage coefficients were determined. Both filtration models adequately described the experimental results when the model predictions were determined from another filter run at a different filtration rate but all other conditions the same. The results from Run 25A showed that Saleh's extrapolation procedure did not fit the experimental results when the chemical pretreatment was not changed and a significantly higher kaolin concentration was filtered.

PLANT DESIGN

General

The selection of unit processes to be used for water treatment depends on a number of factors pertaining to the availability and quality of the raw water source, the treatment plant capacity, the desired water quality and the overall cost of equivalent alternatives. As a minimum, most filtration systems include screening, rapid mixing, flocculation and filtration. When raw water turbidity is either consistently or seasonally high (>30 NTU), sedimentation facilities are incorporated into the treatment scheme to reduce the solids loading onto the filters. In this study, coarse media filters were evaluated as an alternative to sedimentation to reduce the solids load to the dual media filters when the raw water turbidity was near 20, 40 and 100 NTU.

The treatment plant capacity affects the selection of treatment alternatives in two ways. First, there is an economy of scale factor involved with treatment plant sizes. Smaller treatment plants tend to be prefabricated package plants while larger plants are designed specifically for each application. Both types of treatment plants use the same basic unit processes; however, there is more of a tendency to use pressure filtration with the package plants as opposed to gravity filtration with the larger plants. Second, smaller treatment plants are more automated to reduce the need for continuous operator attention. This reduces the plant operating costs for labor and helps to ensure treated water quality since many small communities can not afford full-time trained operators. As a general rule, water plants with capacities

less than 1 MGD tend to be package plants.

The 1977 Safe Drinking Water Act requires that water treatment plants produce a finished water with a turbidity of 1 NTU or less. In some cases, an even lower turbidity is required to produce an acceptable finished water. An example would be the treatment of Great Lakes waters where a finished water turbidity of about 0.1 NTU is necessary to insure asbestos fiber removal.

The overall cost of a water treatment plant includes both the capital cost and the operation and maintenance costs. Most often the treatment system that costs the least and produces the desired effluent quality is the one selected for use. The reliability and flexibility of the selected alternatives should also be considered as a potential secondary cost when selecting the final alternative.

Design and Operation of Series Filtration System

The design of a series filtration system requires consideration of the influent solids concentrations to both sets of filters. The influent solids concentration to the coarse media filters is uncontrollable but the influent solids concentration to the dual media filters can be controlled by the design and operation of the coarse media filters. A standard dual media or multi media design can be used for the second set of filters in a series filtration design. The influent turbidity to the dual media filters should not be allowed to exceed for long durations, the turbidity levels commonly considered maximum levels acceptable for direct filtration. Generally, turbidity levels above 30

NTU are considered too high for a standard direct filtration system. The turbidity range acceptable for direct filtration, however, will vary with the type of turbidity causing particle, the efficiency of chemical pretreatment, the size of filter media, the depth of the media and the filtration rate.

The design of the coarse media filters and the pretreatment system depends on a number of factors which were considered during this study. These factors are:

1) Media size and type. A large media size (3.08 mm) was selected for the pilot plant to provide a large suspended solids storage capacity and lower rate of head-loss development during periods of high turbidity. Sand was selected as the media type since it was felt that it would be easier to clean than a media such as anthracite.

2) Filtration rate. Filtration rates of 12.2, 18.3 and 24.4 m/hr were used in the operation of the pilot plant. When the influent solids concentration was 100 mg/l, the maximum practical flow-rate was 12.2 m/hr. Higher flow-rates were successful at lower kaolin concentrations. In the design of a series filtration system for removal of kaolin in concentrations that may reach 100 mg/l, the maximum design filtration rate should be limited to 12.2 m/hr to avoid excessive breakthrough under high solids loading conditions. Higher filtration rates can be used when the maximum level of kaolin concentration expected is less than 100 mg/l.

3) Pretreatment chemicals. Alum and a nonionic polymer were used successfully in the laboratory pilot filter runs using kaolin as the influent suspended solid. Alum dosages in excess of 20 mg/l (water pH of about 7.0), produced a noticeable degree of flocculation prior to the coarse media filters and increased the rate of headloss development. Polymer dosages of 0.05 to 0.1 mg/l were adequate for successful treatment. Excess polymer increased the rate of headloss development and did not aid the treatment process. In the field study, where diatoms were the predominant turbidity causing particles, alum in combination with a cationic polymer provided the most desirable results. The design of a series filtration system should consider alum in combination with either nonionic or cationic polymer. Alum dosages less than 20 mg/l will be adequate.

During this study alum dosages between 7.5 and 15 mg/l were considered the optimum alum dosages when the influent kaolin concentration was between 25 and 100 mg/l. This study was performed using a raw water pH between 7.0 and 7.5. By maintaining the raw water pH between these values the pH of the raw water suspension, after alum addition, was about 7.1 ± 0.1. Examination of the design and operation diagram for alum coagulation proposed by Amirtharajah and Mills (3), Figure 4, shows that during this study particle destabilization occurred within either the range of sweep coagulation or the range of sweep coagulation in combination with adsorption. According to Figure 4, particle destabilization with alum, by these mechanisms, will occur with the alum dosages used in this study up to a water pH of about 8. In situations where the water pH is above 8 after alum addition, it may be necessary to provide a means of

lowering the pH value to 8 or less to obtain a satisfactory filtered water quality with direct filtration. Flocculation facilities should be considered on an individual basis especially if diatoms are a frequent problem.

4) Operational mode. When the coarse media filters are operated in a declining rate mode the filter operation is more stable. The turbidity breakthrough is more gradual and less severe compared to constant rate operation. Declining rate filtration also produced less of an effect on the system when a filter was removed from service for backwashing. Declining rate filtration is recommended for series filtration design.

5) Backwashing. A water wash followed by a combined air and water wash proved to be a satisfactory combination for cleaning the coarse media filters. The coarse media filters were backwashed until clean and no attempt was made to determine the optimum rates of filter backwash water and air. Optimum backwash requirements should be considered in future investigations of coarse media filtration. Normal backwash procedures can be used with the dual media filters. Effective coarse media backwash in wastewater filtration applications has been reported using a simultaneous air-water backwash at rates of 10-12 gpm/ft^2 and 8-10 CFM/ft^2 respectively.

6) Number of filters. Two filters were used in parallel during this study but for design purposes a minimum of three filters is recommended. Three filters provide a safety factor in addition to minimizing the effect of filtration rate change during backwashing.

The series filtration system provides a viable alternative to conventional treatment when the raw water source is a turbid or seasonally turbid surface water. The most promising application appears to be in small treatment systems using a turbid water supply. For such systems the advantages of series filtration over conventional treatment are:

1) The series filtration system can be easily automated (head loss and/or turbidity control) to provide a minimum of operator attention.

2) The coarse media filters in a series direct filtration system operate effectively over a broad raw water turbidity range with little or no change in chemical pretreatment.

3) In many instances, small conventional treatment plants must be overdesigned hydraulically to provide a larger sedimentation basin to reduce the effects of short circuiting.

4) Chemical requirements and the amount of sludge production are less in series direct filtration than in complete conventional treatment.

5) Almost without exception flocculation facilities, as a separate unit process, are not required when series direct filtration is used.

6) The number of times water must be pumped may be reduced in series direct filtration systems by using a completely pressurized filtration system.

Design examples

There are three basic design options which can be considered for a series filtration system. The first design option, single train of filters, is the use of a single coarse media filter in series with a single dual

media filter. This system would have to operate at a constant rate and would result in no water production during a filter backwash. A filtration system of this type would not normally be used since there is no back-up system for water production in case of equipment failure.

A second design option, a double train of filters, would consist of two coarse media filters in series with two dual media filters. With a two train system, a variable declining rate operational mode could be used when all the filters are in operation. During a filter backwash of one filter in a set, the other filter would receive the entire flow, therefore, each filter must be designed to operate at the maximum filtration rate that gives satisfactory service. This is true also when considering one filter acting as a backup to another filter in case of equipment failure.

The third design option is the use of three or more trains of filters. With a filtration system consisting of three or more trains of filters, a variable declining rate mode can be used consistently under normal operating conditions. During a filter backwash or when one filter is out of service, the filtration rate increase in the other filters is split between the filters.

A schematic diagram of a three train series filtration system is shown in Figure 60. The system has provisions for a rapid mix unit, chemical addition, optional flocculation, three coarse media filters, three dual media filters and an effluent rate control device. The rapid mix requirements could be met by utilizing either pump suction or

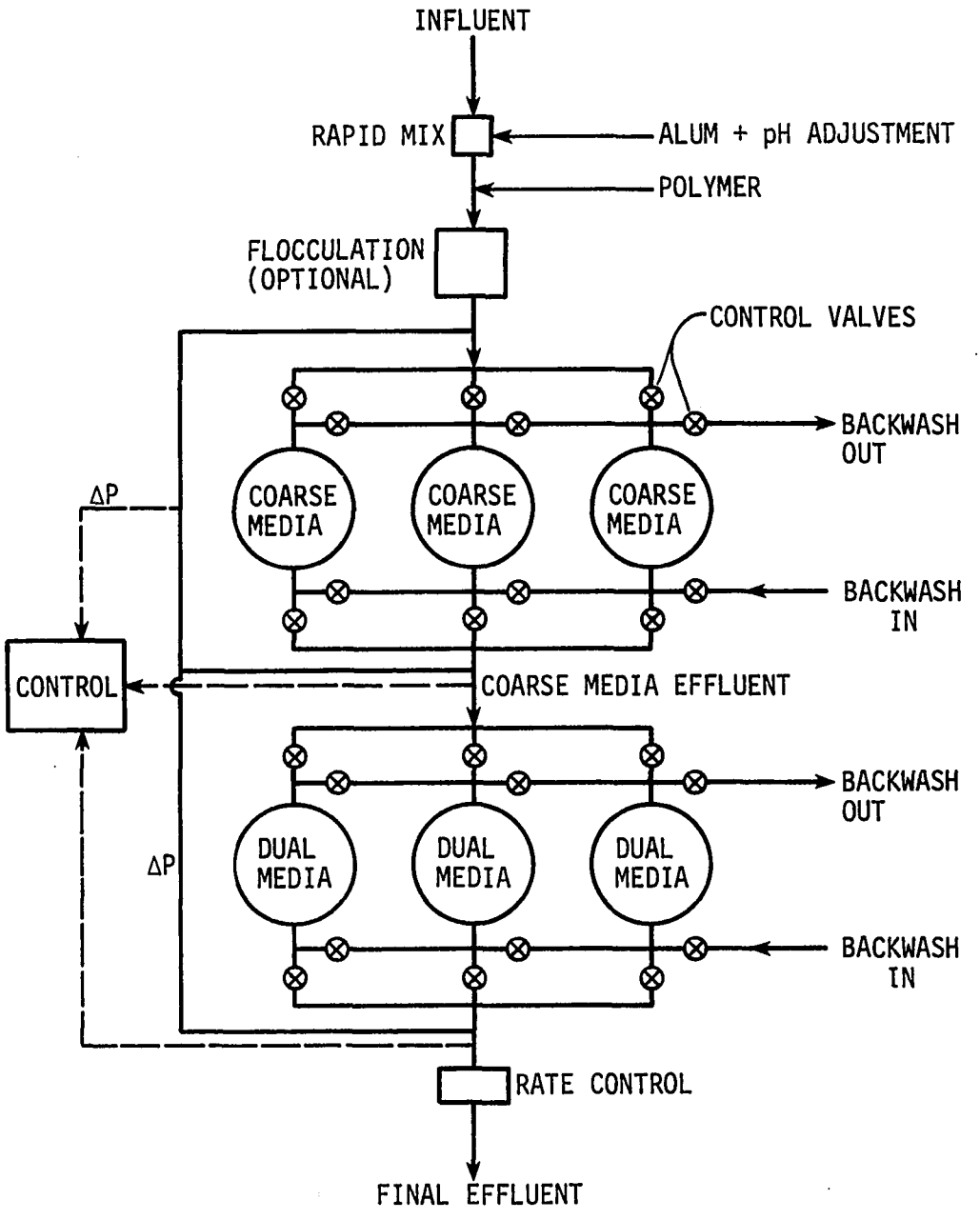


Figure 60. Schematic diagram of a series direct filtration plant for water treatment

commercially available in-line mixers. The addition point of alum and chemicals to adjust the pH of the influent suspension should be just prior to the rapid mix unit. The polymer (nonionic or cationic) is normally added between the rapid mix unit and the coarse media filters. In direct filtration utilizing the series filtration design standard flocculation facilities are normally not used. The amount of flocculation that occurs within the piping between the rapid mix unit and the influent to the filters and within the filter media is normally adequate to produce satisfactory filtration results. Additional study needs to be made to determine whether conventional flocculation facilities or if additional contact time between chemical addition and filtration will enhance the performance of series filtration systems. There is evidence referred to in the literature review section that indicates that flocculation times less than 10 minutes are needed in direct filtration plants using only dual media or multimedia filters (18, 29, 53, 86, 94, 101). O'Melia (65) determined that flocculation was not essential in coarse media direct filtration of high turbidity wastewater effluents. With the coarse media filter serving as a contact flocculation unit, it is doubtful that any additional flocculation time is required in a series direct filtration system.

Figure 60 shows the flow to and from each filter and the backwash flow to and from each filter controlled by a total of 4 values. As an alternative, two 3-way valves could be used to control the flows to and from each filter. The control of the filtration system flow rate can be

accomplished by the use of an effluent control valve.

In package water treatment plants, chlorine for disinfection is usually added in the influent line to the water treatment system. The chlorine residual in the filtered water is monitored automatically and an alarm system sounds if the chlorine residual falls below the desired level.

The control of a series filtration system could be either manual or automatic. In each set of filters, the automatic controls can be designed such that reaching either the maximum effluent turbidity or maximum pressure drop will initiate the backwashing of the dirtiest filter in a set.

The objective of filtration is to provide a turbidity in the water leaving the dual media filters of less than about 0.5 NTU. Therefore, the automatic control should be set to backwash a dual media filter whenever the filtered water turbidity reaches a level of 0.5 NTU or when the headloss across the dual media filter exceeds about 2-3 meters.

The purpose of the coarse media filters is to reduce the solids load on the dual media filters to a maximum dual media filter influent turbidity of about 30 NTU. Therefore, the coarse media automatic backwash control should be set to backwash a coarse media filter whenever the turbidity of the coarse media effluent exceeds 30 NTU or when the headloss across the coarse media filter exceeds about 2-3 meters.

During this study a coarse media filter was backwashed when the effluent turbidity exceeded 33% of the influent turbidity or when the headloss exceeded 2 m through the filter media. The dual media filters

were also backwashed when the headloss exceeded 2 m through the filter media. In filter runs using optimum operating conditions, the final effluent turbidity from the dual media filters rarely exceeded 0.5 NTU and was usually less than 0.2 NTU. In the design of a series filtration system, a maximum pressure drop through the filter media and underdrains in each set of filters should be approximately 2.5 to 3.0 m.

Presently, one water equipment manufacturer, Culligan U.S.A., is marketing a direct filtration system (Multi Tech System) utilizing coarse media filters in series with dual media or multi-media filters. Filter sizes available range from 1.67 feet to 7 feet in diameter with a total three train capacity ranging from 0.065 MGD to 1.15 MGD. The capacity of the Culligan filter system is based on a filtration rate of 17 m/hr (7 gpm/ft^2) assuming all filters are in service and 25 m/hr (10.5 gpm/ft^2) when one train of the 3 train system is in backwash. The backwash water requirement for a Culligan coarse media filter is 140 gal/ft^2(14 gpm/ft^2 for 10 minutes). Air scour, at the rate of 5-6 cfm/ft^2, is also used in the coarse media filter backwash. The model number, daily capacity and filter size of the Culligan series filtration system are given in Table 16.

As stated previously the Culligan system design capacity is based on a filtration rate of 17 m/hr with all the filters in service. The selection of this filtration rate assumes that the filtration system will produce satisfactory results regardless of the levels of raw water turbidity.

Table 16. Culligan "Multi-Tech" clarification and filtration systems details[a]

Model	Maximum Daily Capacity[b] (3 trains @ 24 hrs) MGD	m^3/day	Filter Size Diameter ft	m	Area (1 filter) ft^2	m^2
MT-20	0.065	245	1.67	0.51	2.18	0.20
MT-24	0.095	360	2.0	0.61	3.14	0.29
MT-30	0.150	570	2.5	0.76	4.91	0.46
MT-36	0.215	815	3.0	0.91	7.07	0.66
MT-42	0.280	1060	3.5	1.07	9.62	0.89
MT-48	0.367	1390	4.0	1.22	12.57	1.17
MT-54	0.475	1800	4.5	1.37	15.90	1.48
MT-60	0.580	2209	5.0	1.52	19.63	1.82
MT-72	0.842	3190	6.0	1.83	28.27	2.63
MT-84	1.150	4337	7.0	2.13	38.48	3.58

[a]Culligan Product Bulletin 8125-15.

[b]Filtration rate of 7 gpm/ft^2 or 17 m/hr.

To aid in the design of a series filtration system, Figures 61 and 62 were developed. Figure 61 is a combination of Figure 57 and the application of the mathematical extrapolation of the filtration data as proposed by Saatci and Oulman (81). Also, for comparison, several points calculated by Saleh's (82) model extrapolation procedure are shown in Figure 61. The constants used for both model predictions were obtained from Tables 13, 14 and 15 for the run indicated on Figure 61. The mathematical extrapolation was applied to pilot plant data obtained at kaolin concentrations of 50, 100

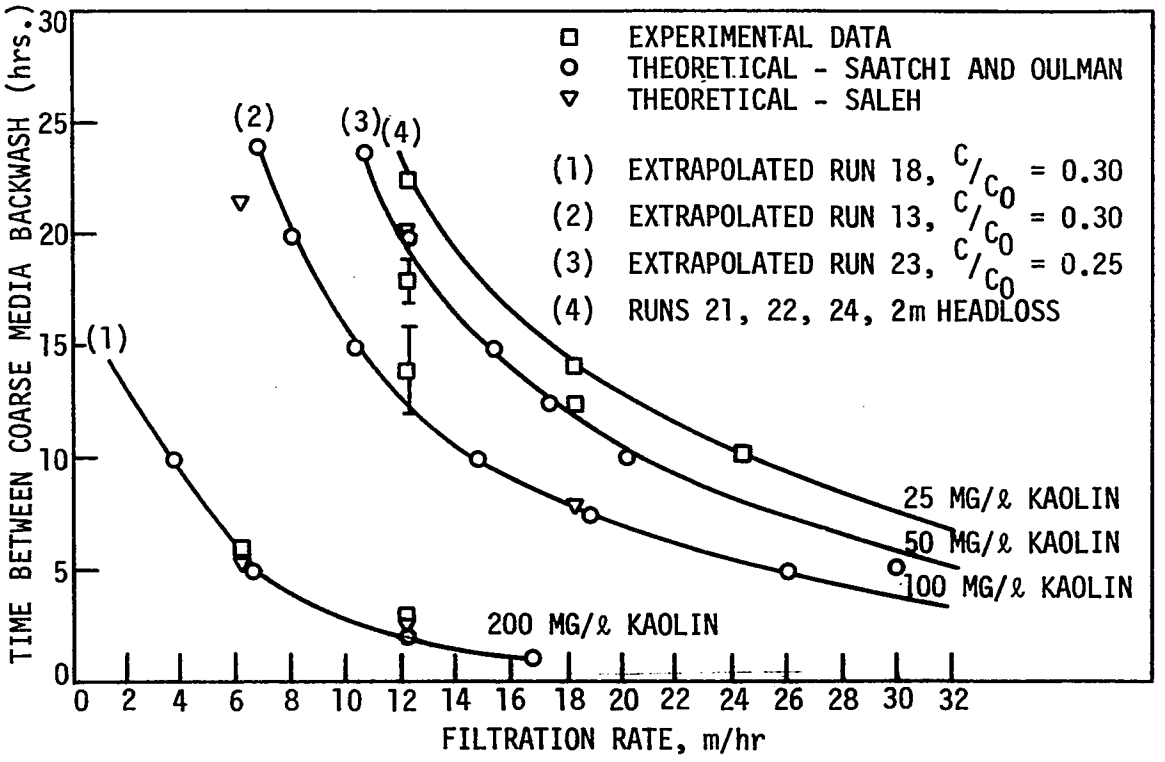


Figure 61. Time between coarse media backwash vs. filtration rate at various kaolin concentrations

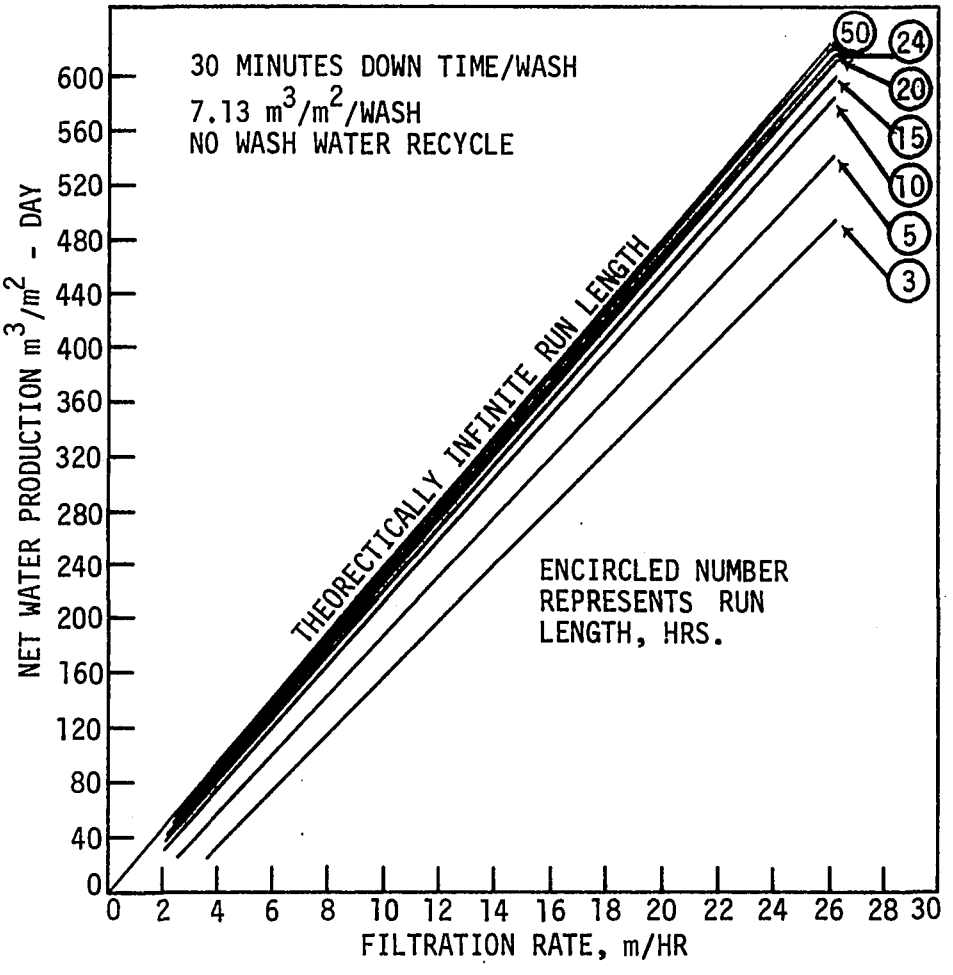


Figure 62. Net water production vs. filtration rate at various run lengths

and 200 mg/1. The curve shown for 25 mg/1 was developed from actual performance data. Since at 25 mg/1 the coarse media filter terminal head-loss was reached before turbidity breakthrough occurred. Figure 62 shows the net water production for a 24 hour period as a function of filtration rate and run length. In developing this figure, the filter backwash was assumed to take 30 minutes and the amount of backwash water consumed was 7.13 m^3/m^2. A detailed procedure for developing a figure such as Figure 62 is given by Huang (42).

The use of Figures 61 and 62 in a series filtration system design is illustrated in the following design examples:

Example 1: Design a two train series filtration system with a maximum capacity of 304 m^3/day (80,000 gal/day). The maximum anticipated influent solids concentration is equivalent to 50 mg/1 of kaolin.

Step 1. In the design of a two train filtration system one filter should be capable of handling the entire flow when the other filter is being backwashed or is out of service indefinitely. The minimum filter run length should be 10 hours. From Figure 61, the maximum filtration rate that can be used at a kaolin concentration of 50 mg/1 and still maintain a 10 hour filter run length is about 21 m/hr. The average filtration rate would therefore be 10.5 m/hr and the expected run length would be 24 hours.

Step 2. From Figure 62, the net water production at a filtration rate of 10.5 m/hr and a 24 hour filter run is about 240 m^3/m^2-day. The net water production for a 10 hour filter run at a filtration rate of 21 m/hr is about 465 m^3/m^2-day.

Step 3. The required total filter area is determined by dividing

the maximum required capacity by the net water production.

$$\text{Required filter area (2 train system)} = \frac{304\ m^3/day}{2(240)m^3/m^2\text{-day}} = 0.63\ m^2$$

$$\text{Required filter area (1 train system)} = \frac{304\ m^3/day}{465\ m^3/m^2\text{-day}} = 0.65\ m^2$$

Step 4. The design requirements can be satisfied by selecting a Culligan two train filtration system composed of filters having an area of 0.66 m^2 each (Table 16). An average filtration rate less than 10.5 m/hr would result from the use of filters with a total area slightly greater than required.

A design approach similar to the above could be used to design a series filtration system with any number of filter trains up to a capacity where package water treatment plants are economically competitive with water treatment plants designed on an individual basis. A larger capacity series filtration treatment plant design is given in Example 2.

Example 2. Design a four train series filtration system with a maximum capacity of 3406 m^3/day (0.9 MGD). The maximum anticipated influent solids concentration is equivalent to 100 mg/l of kaolin.

Step 1. From Figure 61 a 10 hour run length can be obtained at this kaolin concentration at a filtration rate of about 15 m/hr. This filtration rate would be the maximum filtration rate when one filter out of four is out of service. The average filtration rate would be 11.25 m/hr and the expected run length would be 13.5 hours.

Step 2. From Figure 62, the net water production at a filtration rate of 11.25 m/hr and a run length of 13.5 hours is about 250 m^3/m^2-day. The net water production for a 10 hour filter run at a filtration rate of 15 m/hr is about 325 m^3/m^2-day.

Step 3. $$\text{Required filter area (4 train system)} = \frac{3406\ m^3/day}{4(250)\ m^3/m^2\text{-day}} = 3.41\ m^2$$

$$\text{Required filter area (3 train system)} = \frac{3406\ m^3/day}{3(325)\ m^3/m^2\text{-day}} = 3.49\ m^2$$

Step 4. The design requirements can be satisfied by selecting a Culligan four train filtration system composed of filters having an area of 3.58 m^2 each (Table 16). The average filtration rate would be less than 11.25 m/hr.

$$\text{Average filtration rate} = \frac{3.41}{3.58}\ (11.25\ m/hr) = 10.7\ m/hr.$$

Another design option is to increase the depth of the coarse media filters to extend the filter run length when filter runs are shorter than about 10 hours. Pilot studies are the best method to determine the required filter depth for a desired increase in filter run length. However, a filtration model can be used to determine an approximate depth as a starting point or to determine whether the required depth would be practical. The filtration model developed by Saatci and Oulman (81) was based on the results of pilot filtration experiments using filters

of various depths. With the model proposed by Saatci and Oulman (81). the required depth for a specified effluent turbidity and desired run length can easily be approximated by the following equation:

$$x = \frac{V_a \, C_o(-\ln Ri)}{F_o \, \beta} \tag{1}$$

The approach velocity, V_a, and influent turbidity, C_o, are both known. The values of F_o and β are determined from the data of a pilot plant run. The value of F_o represents the storage coefficient and the value of β, the slope of the logistic curve, is proportional to the attachment coefficient. The value of ln Ri is determined by equations 2 and 3.

$$\ln R = \frac{C}{C_o - C} \tag{2}$$

$$\ln Ri = \ln R - \beta t \tag{3}$$

The value used for filtration run length, t, is the desired filtration run time. The values of β and F_o determined in Run 13 were used to estimate a filter depth for a 24-hour filter run with a final turbidity equivalent to $C/C_o = 0.33$. By this method it was determined that increasing the filter media depth from 1.32 m to 2.5 m should result in a 24-hour run length.

Equations 1, 2 and 3 can be used to determine an approximate relationship between filter media depth and the time between coarse media backwash for a specific kaolin concentration. Figure 63 shows the relationship

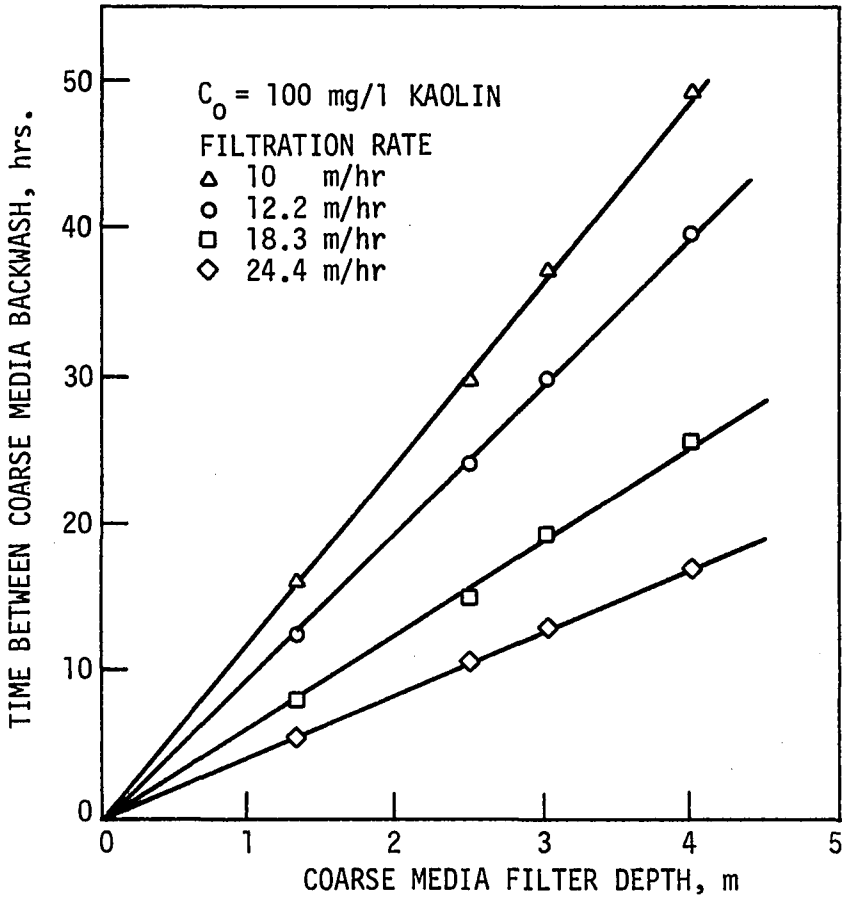


Figure 63. Time between coarse media filter backwash vs. coarse media filter depth

between run length and media depth at various filtration rates when 100 mg/l of kaolin is used as the influent suspended solids concentration. Figure 63 was developed with the values obtained from the analysis of Run 13 using Saatci and Oulman's modeling technique. The run length time, t, was calculated by first selecting values of media depth and filtration rate and solving for -ln Ri in equation 1. Second, the run length was determined using equations 2 and 3 by substituting in the appropriate values for C, C_o and β. In Figure 63, a filter run length of 10 hours is shown for a filter media depth of 2.5 m and a filtration rate of 24.4 m/hr. From Figure 62, a net water production of 540 m^3/m^2-day can be expected under these conditions. If the series filtration system in Example 2 utilized a media depth of 2.5 m instead of the 1.32 m media depth used to determine the design curves in Figure 61, the average filtration rate in the system would be 18.3 m/hr (24.4 m/hr x 0.75). The expected run length, as shown in Figure 63, would be about 16 hours. Under these conditions a net water production of 415 m^3/m^2-day would be expected. Using the 2.5 m filter media depth, the average plant capacity would be 5940 m^3/day (1.57 MGD) and the maximum capacity with one filter out of service and still maintaining a 10 hour filter run length would be 5800 m^3/day (1.53 MGD).

Design of larger plants

Series filtration could be used on larger scale plants to treat high turbidity water. With plant capacities greater than about 1 MGD, gravity filters would likely be used. The previously mentioned

advantages of series filtration in terms of reliability and operational flexibility in small treatment plants have less of an influence on the selection of series filtration in preference to sedimentation followed by filtration for large treatment plants. In larger facilities sedimentation facilities are large enough to function effectively. Assuming both types of systems could be used to treat a raw water to the desired final turbidity goal, the alternative selection would most likely be based on the cost of treatment. Each case would need to be considered on an individual basis. Factors influencing alternative selection are land costs and the need for flocculation facilities in series filtration. Sedimentation facilities are generally more costly in terms of capital cost while filtration facilities are more costly in terms of operation costs. The high operation costs for filtration are due to the energy demands. The selection of alternatives based on costs may be decided on the prevailing interest rate versus energy costs.

The Environmental Protection Agency has published cost estimating curves for water treatment facilities (36). This publication was used to develop an example cost comparison for a 40 MGD water plant, as shown in Table 17. The costs shown in Table 11 are for conventional flocculation facilities, sedimentation facilities and coarse media filters. The costs associated with raw water pumping, rapid mix facilities, chemical feed equipment, dual media filtration, disinfection, storage and distribution are assumed to be the same regardless of the process selected. The assumptions made to develop Table 17 from the cost curves are:

1. The capital cost and operation and maintenance costs given for multimedia deep bed gravity filters could be used for an estimate of the coarse media filter costs.

2. Common backwash facilities could be incorporated in a series filtration design for use with both the coarse media filters and dual media filters. Therefore, no additional allowance was made for backwash facilities in the coarse media filter estimate.

3. Coarse media filters and sedimentation facilities would be operated twelve months out of the year.

4. The construction cost index used to develop the cost curves was 265.38 for October, 1978. An average construction cost index of 334.14 was used to update the construction costs. This index was the average for 20 U.S. cities reported in the August 13, 1981, Engineering News Record (24).

In the example cost estimate shown in Table 17, the use of coarse media filtration facilities instead of conventional flocculation and sedimentation would result in an annual savings of $134,100. This represents a considerable savings and indicates that coarse media filtration as a substitute for conventional flocculation and sedimentation should be given careful consideration when selecting water treatment alternatives for large treatment facilities. Furthermore, the cost comparison shown in Table 17 does not reflect the lower alum requirements (one third to one half that required in conventional treatment) and lower sludge production in the coarse media filter alternative. Consideration of

these two factors would give an even larger economic advantage to coarse media filtration in comparison to conventional flocculation and sedimentation.

Table 17. Comparison of estimated costs of treatment alternatives for a 40 MGD water treatment plant

	Conventional Flocculation 35 min, G = 50	Sedimentation 40.73 m^3/m^2-day	Coarse Media Filtration 12.2 m/hr
Capital Cost, $			
Construction	562,900	2,829,600	2,387,200
Sitework (5%)	28,100	141,500	119,400
Contractor Overhead + Profit @ 10%	59,100	297,100	250,700
Subtotal	650,100	3,268,200	2,757,300
Engineering @ 10%	65,000	326,800	275,700
Subtotal	715,100	3,595,000	3,033,000
Land @ $30,000/acre	9,000	27,500	3,800
Total, $	724,100	3,622,500	3,036,800
Annual Cost, $			
Amortized Capital cost @ 12%, 20 yr.	96,900	485,000	406,600
Energy @ 5.5¢/kw-hr.	8,500	3,900	61,100
Maintenance Material	6,000	12,700	15,300
Labor @ $15/hr	5,900	65,200	67,000
Total, $	117,300	566,800	550,000

SUMMARY AND CONCLUSIONS

There are two basic types of water treatment processes presently used to treat a surface water supply. These two processes are direct filtration and complete conventional treatment. The major differences between the two processes are that in complete conventional treatment, conventional flocculation and sedimentation facilities are included in the process whereas in direct filtration, conventional flocculation facilities may or may not be included and sedimentation facilities are eliminated.

The selection of a water treatment process is usually based on the quality (turbidity) of the raw water supply. Generally, direct filtration can be used when the raw water turbidity remains below about 30 NTU. When the raw water turbidity exceeds about 30 NTU and remains high for extended periods complete conventional treatment is normally selected. Standard water treatment filters (dual or multimedia) used in direct filtration of low turbidity waters do not have the necessary solids holding capacity to provide a reasonable filter run length without excessive headloss or poor effluent turbidity when the raw water turbidity is consistently high. The key to the extension of the direct filtration process to the treatment of high turbidity waters is to increase the solids storage capacity of the direct filtration system. Baumann[1]

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[1]Personal communication, Dr. E.R. Baumann, Department of Civil Engineering, Iowa State University, Ames, Iowa.

suggested that the use of coarse media filters in series with standard dual media filters would increase the solids storage capacity of a direct filtration system and make the process applicable to the treatment of high turbidity waters. Baumann's suggestion of using coarse media filters in series with dual media filters was partially based on the results of a study conducted by Breland and Cleasby (14). These investigators evaluated direct filtration of high turbidity surface waters using single stage dual media filters and small dosages (<20 mg/1) of alum in the chemical pretreatment. The filter runs were unreasonably short due to the early solids breakthrough at low headloss.

As suggested by Baumann, one equipment manufacturer, Culligan USA, designed a package direct filtration system comprised of coarse media filters in series with multimedia filters for small scale applications requiring about 1 MG or less of treated water daily. The Culligan direct filtration system uses a constant filtration rate of 17 m/hr (7 gpm/ft), a single cationic polymer in the chemical pretreatment, no conventional flocculation and at run termination both the coarse media and multimedia filters are backwashed.

Cleasby and Saleh (18) conducted field studies using the Culligan series filtration system. The Culligan system was operated at a constant filtration rate of 9.3 m/hr (3.8 gpm/ft^2) or 15.4 m/hr (6.3 gpm/ft^2), the pretreatment chemical was alum or a cationic polymer, no conventional flocculation facilities were used and both the coarse media and multimedia filters were backwashed at run termination. The use of alum alone

in the chemical pretreatment was not evaluated at a filtration rate of 9.3 m/hr. At a filtration rate of 15.4 m/hr, alum alone in the chemical pretreatment was found to be ineffective in producing an average filtered water turbidity below 1.0 TU. The use of only the cationic polymer in the chemical pretreatment resulted in satisfactory filter water turbidity (<1.0 TU) for a reasonable run length (>10 hours) at both filtration rates evaluated. However, Cleasby and Saleh (18) recommended that when using the cationic polymer in the chemical pretreatment that the maximum filtration rate be limited to 12.2 m/hr. An average raw water turbidity of 123 TU was the highest average raw water turbidity encountered by Cleasby and Saleh during their study. At this raw water turbidity the filtration system run length was 19 hours, the filtration rate was 9.3 m/hr and the cationic polymer was used in the chemical pretreatment. Cleasby and Saleh (18) also compared the performance of a series filtration system using a dual media filter as the second filter in the series to the Culligan series filtration system which used a multimedia filter as the second filter in the series. They found that essentially equivalent finished water turbidities were produced with either system.

Several operation and design considerations relevant to the operation and design of a series filtration system were not included in the Culligan system design and the field evaluation of the Culligan system conducted by Cleasby and Saleh (18). These operation and design parameters are:

1. The effect of backwashing the coarse media filters separately from the dual media filters by establishing independent effluent turbidity and filter headloss criteria for each filter type.

2. The effect on the filtration system performance when declining rate filtration is used compared to when constant rate filtration is used.

3. The use of alum and polymer in the chemical pretreatment.

The study reported herein, emphasized the evaluation of the effects of the above operation and design considerations on the performance of a series filtration system. The study was conducted in three phases. The first phase involved the use of jar tests to determine optimum chemical dosages for destabilization of kaolin and aluminosilicate. The second phase utilized the results of the jar tests to evaluate direct filtration of high turbidity water using coarse media filters followed in series by dual media filters. Kaolin was used in the majority of the filtration experiments as the influent suspended solid. The third phase involved a limited field investigation of the series filtration system using Des Moines River water as the raw water source.

From the results of this study, the following conclusions can be made:

1. The jar test procedure proved to be useful in determining the combination of coagulants best suited for use in the series filtration experiments. The optimum alum dosage determined in the jar tests for particle removal by settling was higher than the optimum alum dosage for filtration.

2. The results of the laboratory filter runs show that coarse media filters in series with dual media filters are a viable treatment alternative for direct filtration of high turbidity waters. When the raw water turbidity was as high as 100 NTU and the filtration rate was 12.2 m/hr, the coarse media filter run lengths were above the minimum desired run length of 10 hours, the dual media filter run lengths were between 21 and 23 hours and the final effluent turbidity averaged less than 0.1 NTU.

3. In direct filtration of high turbidity waters the use of coarse media filters in series with dual media filters increases the solids storage capacity of the filtration system and increases the dual media filter run length while maintaining a low turbidity final effluent. When 25 to 100 mg/l of kaolin was used in the raw water suspension, an average of 80 to 98% of the influent turbidity was removed in the coarse media filters.

4. The use of 200 and 300 mg/l of kaolin in the raw water suspension resulted in coarse media filter run lengths of less than 10 hours due to turbidity breakthrough in the coarse media filter effluent. The depth of the coarse media filters should be increased when such high turbidities are expected so that a minimum run length of 10 hours can be achieved.

5. The results of the laboratory filtration experiments indicated that an equivalent final water quality was obtained when the coarse media filters were operated in either a constant rate or declining rate mode. A declining rate mode of operation in the coarse media filters was preferred since the rate of turbidity breakthrough

was gradual and less severe at run termination compared to when a constant rate mode was used. Operating the coarse media filters at a constant rate resulted in the stripping of solids from an operating filter when the other filter was taken out of service for backwashing. This caused a rapid increase in the headloss development in the dual media filters and at times poor final effluent turbidity. The use of a declining rate mode in both the coarse media filters and dual media filters resulted in excellent finished water turbidity results. The final effluent turbidity was consistently less than 0.1 NTU and often as low as 0.03 NTU. After the backwash of one of the dual media filters, the final effluent turbidity remained below 0.1 NTU when the filters were operated in the declining rate mode. However, when a constant rate mode was used, the final effluent turbidity was often above 0.1 NTU for one or two hours.

6. The use of alum and a nonionic polymer in the chemical pretreatment resulted in finished water turbidity results superior to those when only alum was used in the chemical pretreatment. The use of alum in the chemical pretreatment aids in shortening the filter ripening period while the polymer strengthens the deposits retained in the filters.

7. Depending on the raw water kaolin concentration (25 to 200 mg/l) the optimum alum dosage range was 7.5 to 20 mg/l. Alum dosages in excess of 20 mg/l produced an undesirable amount of large flocs prior to the coarse media filter. A nonionic polymer dosage of 0.05 to 0.1 mg/l in combination with alum was necessary for the production of a high quality final effluent. Alum alone produced unsatisfactory results.

8. During the laboratory filter runs, a consistent pattern of particle deposition within the coarse media filters was observed. This was an on-going process that began near the top of the filters and progressed downward. Initially, the suspension particles coated the filter media. Subsequent to the coating of the filter media the filter pores were filled with the suspension particles. Finally, narrow channels developed within the filter media which allowed the suspension particles to travel deeper into the filter for subsequent removal.

9. The coarse media effluent turbidity results from selected laboratory filtration experiments were used to evaluate the usefulness of two mathematical models in predicting coarse media filter performance. By design, the coarse media filters operate for the most part in a stage of controlled breakthrough. This type of operation is ideal for mathematical model applications. Both filtration models could be applied reasonably well to the coarse media filtration data when the influent kaolin concentration was 50 mg/l or more. The validity of the filtration models was not tested with filter runs using 25 mg/l of kaolin as the influent solids concentration since significant solids breakthrough did not occur in the coarse media filters at this kaolin concentration.

10. The two mathematical models were used to extrapolate the results of the laboratory filter runs to estimate the effects of different operating conditions for the development of design curves. The extrapolation of filtration data should be done cautiously since some filtration

parameters will change depending on filtration rate, chemical treatment and influent solids concentration.

11. The results of the field study indicated that the pretreatment chemical combination (alum and a nonionic polymer) used in the laboratory filter runs was inadequate when used in the filtration of the Des Moines river water. Alum combined with a cationic polymer produced promising results. The nature of the turbidity causing particles was most likely the reason for a change in chemical requirements. The laboratory studies were conducted with kaolin clay while the turbidity in the Des Moines River consisted mainly of diatoms.

12. Series direct filtration should be given serious consideration in the evaluation of potable surface water treatment alternatives for small scale applications. The operation of a series direct filtration system can be almost totally automated and the system can be operated over a broad range of raw water turbidities with little or no change in chemical pretreatment requirements. In large scale applications, where operation requirements are generally less critical since better skilled operators are employed, series direct filtration may be economically advantageous when compared to the use of conventional flocculation followed by sedimentation.

RECOMMENDATIONS FOR FUTURE WORK

Additional study of the series direct filtration system concept should be performed to further refine the system. The major emphasis of further study is recommended in the following areas.

1. Laboratory studies to optimize coarse media grain size, shape and depth would be of primary interest. In this study, a uniform sand was used. The sand size was 3.08 mm and the depth of the media was about 1.32 m. To increase filter run lengths, consideration should be given to the evaluation of a more angular sand media with perhaps a smaller grain size and an increased media depth.

2. The use of a short flocculation period prior to the coarse media filters should be evaluated. At high clay concentrations it is likely that flocculation is not needed; however, at high diatom concentrations flocculation may prove beneficial to the process.

3. Once the previous studies have been completed an extensive field study should be undertaken to test the laboratory results. This study should be designed so that the filtration system could be operated at various periods throughout the year when major changes in raw water quality occur. A major concern of this study should be the evaluation of the chemical treatment combinations. As a minimum the chemical pre-treatment should consist of alum, alum in combination with a nonionic polymer, alum in combination with a cationic polymer and cationic polymer alone.

4. Any additional studies should incorporate the use of mathematical models, particularly the two models evaluated in this study. The use of mathematical models in future studies would aid in selecting design parameters for model varification which in turn may be useful in further model refinement. The determination of the relationship between the filtration operating parameters and the filter attachment and storage coefficients would be of considerable value in model applications to design.

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