FLOCCULATION, SEDIMENTATION AND FILTRATION
A technical and economic analysis of water treatment

by
Torsten Hedberg

Göteborg 1976

PREFACE

Research and development of water treatment processes have been carried out since 1962 at the Division of Water Supply and Sewerage, Chalmers University of Technology. To a large extent the work has consisted of applied research ranging from laboratory experiments to full scale operations. The research has been carried out in cooperation with the Water and Sewage Works in Göteborg, and the activities have mainly been located in the Water Treatment Plants at Alelyckan and Lackarebäck, where the raw water from the river Göta Älv is treated. In connection with the long distance delivery of water to the south of Sweden, the South-Water Project, investigations of raw water from lake Bolmen have been performed during two years. The two raw waters were very soft waters with low buffer capacities.

The intention of this thesis is to give a concise description of water treatment techniques, especially for soft surface water, from a technical and economic point of view and to give guidelines for future investigations.

The experimental investigation was partly carried out by Ö. Andersson, R. Hammarberg, and C.Hernebring under my supervision.

I wish to express my sincere thanks to Professor Gunnar Weijman-Hane, Head of the Division, for stimulating discussions, to the personnel at the Water Treatment Plants; and to all my colleagues for their invaluable assistance and encouragement.

Göteborg 1976
Torsten Hedberg

TABLE OF CONTENTS

1. INTRODUCTION

1.1 Introductory remarks

In the future, demands on water quality are likely to increase, and therefore, effective and safe water treatment techniques will be needed. During recent years, many new unit processes and unit operations and combinations of them have been developed. The treatment techniques differ from country to country, a certain type being predominant in each country. Certain variations in the process are necessary because of differences in the raw water quality. However, in most cases, the variations depend on the particular technique developed in each country.

In Sweden as in many other countries, the unit processes and unit operations - flocculation (coagulation) and separation (sedimentation, flotation, filtration) - are, as a rule, connected with each other, and thus their functions are interdependent. Because of the fast technical development in the field of water treatment, a survey of the water treatment process from a technical, as well as an economic point of view, is important. This fact has also been pointed out by one American group that in 1974 planned the need for research activities in connection with water treatment. (A Committee Report, 1974).

The yearly investment cost for surface water treatment plants in Sweden was estimated at 60-70 mill. Sw. Cr. in 1971-1972 (Hilmer and Andersson, 1973).

1.2 Objective of study

The objective of this study has been to complete and combine existing theories for the different unit processes and unit operations - coagulation, sedimentation, and filtration - in order to get an overall technical and economic model of the treatment process. Such a model is valuable for the planning of research and development work, but its main use is in the planning of new water treatment plants and analyses of existing plants where increased capacity is required. It is therefore important not only to develop and explain the different submodels that are included in the overall model, but above all to reach a close agreement between the overall model and reality. The sub-models are suitable for further investigations, provided that the overall model shows economic advantages when the unit process or unit operation in question is used. In the following overall model, some submodels are of a general type, and some are related to certain prevailing conditions.

The advantage of full scale testing is that the results obtained are reliable; with this method it is, however, difficult to obtain variations of the investigated parameters and to perform a sufficient number of tests for a valid statistical evaluation.

In this report investigations carried out during 1963-1975 are described and analyzed. Numerous tests have been done, and some of the results are presented as examples in graphs and tables.

1.3 Other investigations

The different unit processes and unit operations in water treatment techniques have been described in a vast number of articles and theses. Some of these are referred to in connection with the discussion of the unit operations in question. Investigations of optimization of the water treatment process are, however, very rarely described in the literature.

1.4 Unit operations and their interrelation

1.4.1 Definitions

The concept of the unit operations is described by Fair, Geyer, and Okun in Waste Water Engineering, 1967:

The concept of unit operations first found expression in the analysis of common procedures in chemical engineering. As suggested by A.D. Little: "Any chemical process, conducted on whatever scale, may be resolved into a coordinated series of what may be called 'unit actions'...... The number of these basic unit operations is not very large, and relatively few of them are involved in any particular process". The kind of thinking that emerged from this early pronouncement of principle has contributed greatly to the development of chemical engineering and, in the course of time, also to the advancement of water and wastewater treatment. Among its principal contributions are (1) a better understanding of inherent processes and capabilities in water and wastewater treatment; (2) the development of mathematical and simple physical models or analogs of treatment mechanisms and their use in identifying the basic components of treatment plant design; and (3) the coordination of effective treatment procedures to attain required plant performance and effluence.

Within the context of this book, an essential element in the concept of unit operations is acceptance of the principle that there is no fundamental difference between the unit actions by which the treatment objectives of water and wastewater are reached, but merely a difference in the nature of the raw waters subjected to treatment and in the quality of the finished waters or effluents produced. In both instances unwanted substances are either removed from the incoming waters or transformed into acceptable substances, or both. Moreover, unwanted behaviour is either suppressed or transformed into acceptable behaviour. Nevertheless, there are some operations that, for economic and technological reasons, find routine application only in water treatment or wastewater treatment.

In comparison with the unit operations of chemical engineering, there is a similar sharing of many of the procedural purposes and principles. However, the terminology, units of measurement, and many of the factors governing important unit operations or processes of water and wastewater treatment are often different from those of chemical engineering. This is not reprehensible, nor need it be a hindrance to fruitful communication between the two branches of engineering. Both are entitled to the preservation of their technological vocabulary, their systems of measurement, their integrity, and their culture. Chemical engineers generally apply the term unit operations to procedures in which the changes produced are essentially physical. In water and wastewater treatment, however, chemical and biological operations are often the governing components of treatment processes; thus sanitary engineers are justfied in lifting the chemical engineer's restriction of the term unit operations to physical phenomena in order to expand its concepts into chemical and biological phenomena as well.

The unit operations in a conventional water treatment plant are shown in FIG. 1-1.

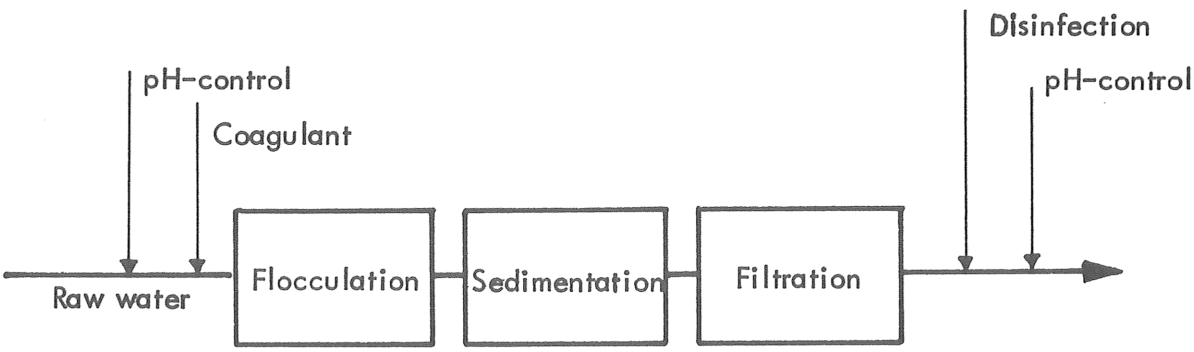


FIG. 1-1 Flow sheet of a conventional water treatment plant

In order to obtain high efficiency, we need thorough knowledge of the different factors involved in the unit operations as well as knowledge of the interaction of the different unit operations. In general, it is believed that an increased knowledge of each unit operation means a possibility to attain a higher efficiency without a significant increase in costs. Further improvement is related to a change in construction which often implies higher costs. This increase must be in proportion to the result obtained, and it is therefore important to analyze the total water treatment operation. Calculation of the total costs under varying conditions will give information on the optimal design of a water treatment plant.

1.4.2 Raw water quality

Raw water guality is more or less changing during the year. The temperature varies from 0 to 25^{0} C in the surface waters. These temperature variations are very important for all the unit operations, not only because of changes in the viscosity, but also because of changes in the biological activity and the chemical properties with changing temperature. The water quality also differs from one year to another, and the investigations described in this report are affected by this fact. The deviations in the results are often very large, but these deviations may be compensated for by an increase in the number of tests and by studies of some surface waters within the category of soft surface waters in order to establish certain limits.

1.4.3 Choice of chemicals

The choice of chemicals is of great importance for the function of the various unit operations. In Sweden aluminium sulfate (alum) is the predominating coagulant, and activated silica is normally used as a coagulant aid. For the pH-control, which is always necessary when treating soft waters, lime (calcium hydroxide) is often used in larger treatment plants, while soda (sodium carbonate or sodium hydroxide) is often used in smaller treatment plants. Investigations (Hedberg, 1969 and 1974) have shown operational advantages with a simultaneous increase in the water hardness, by the use of lime and carbon dioxide, a technique alredy applied at some water treatment plants. The chemicals used affect the separation characteristics, structure, size, density, and strength of the flocs formed in the flocculation unit. The main part of this investigation has been carried out with aluminium sulfate as a coagulant and lime for pH-control. Activated silica has been used as a coagulant aid in various dosages.

1.4.4 Flocculation - sedimentation - filtration

When alum is added to the water containing colloids, a destabilization of the suspension is achieved, and microflocs are formed already in the mixing device. In order to increase the floc size, one can use different systems of agitation. Mixing is a brief operation, seeking a quick response, whereas agitation and stirring are more protracted operations, aiming at the conjunction of suspended particles or flocs. Aggregation of particles is referred to as floc growth. A combination of mixing and stirring or agitation that produces aggregation is called flocculation.

In optimization studies, it is important to describe adequately the properties of the flocs. This is, however, difficult as no simple standard methods are available for this purpose. In addition, the flocs are affected in different ways depending on the separation operation used. In this investigation sedimentation analysis equipment has been used for a general characterization of the settling properties of the flocs (see FIG. 1-2). From the analysis the settling velocities of the flocs can be calculated.

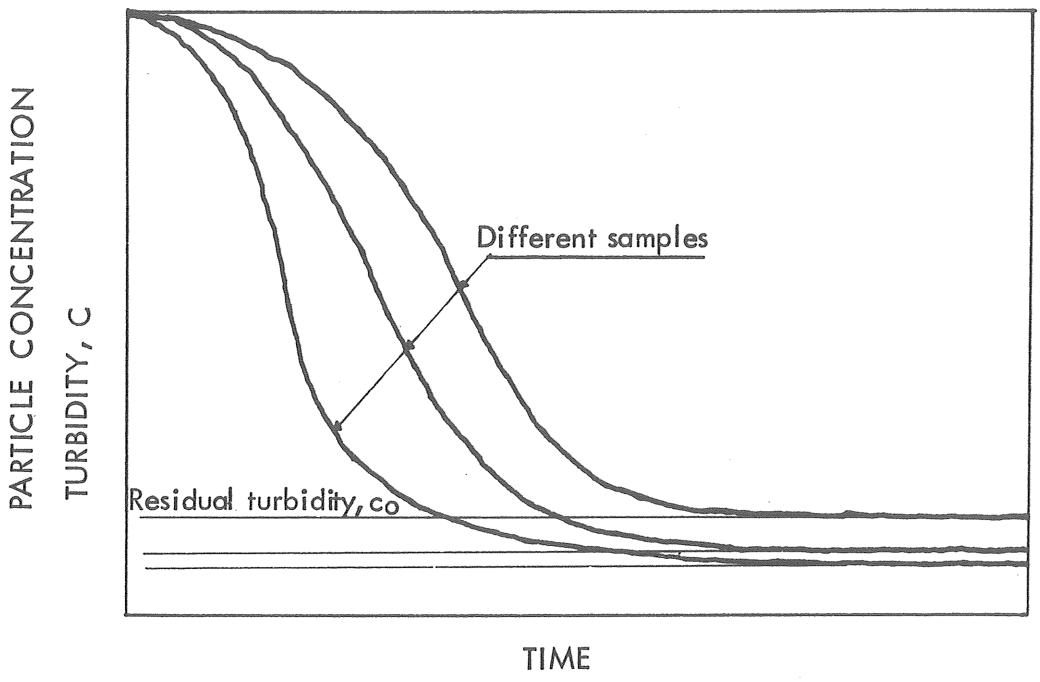


FIG. 1-2 Sedimentation analyses.

The optimization procedure of flocculation, sedimentation, and filtration is briefly summarized in the following figures.

With the sedimentation analysis as a basis, it is possible to obtain a relationship between the residual turbidity (or other concentration values) and the flocculation time (FIG. 1-3). Design and operation parameters affect the relationship.

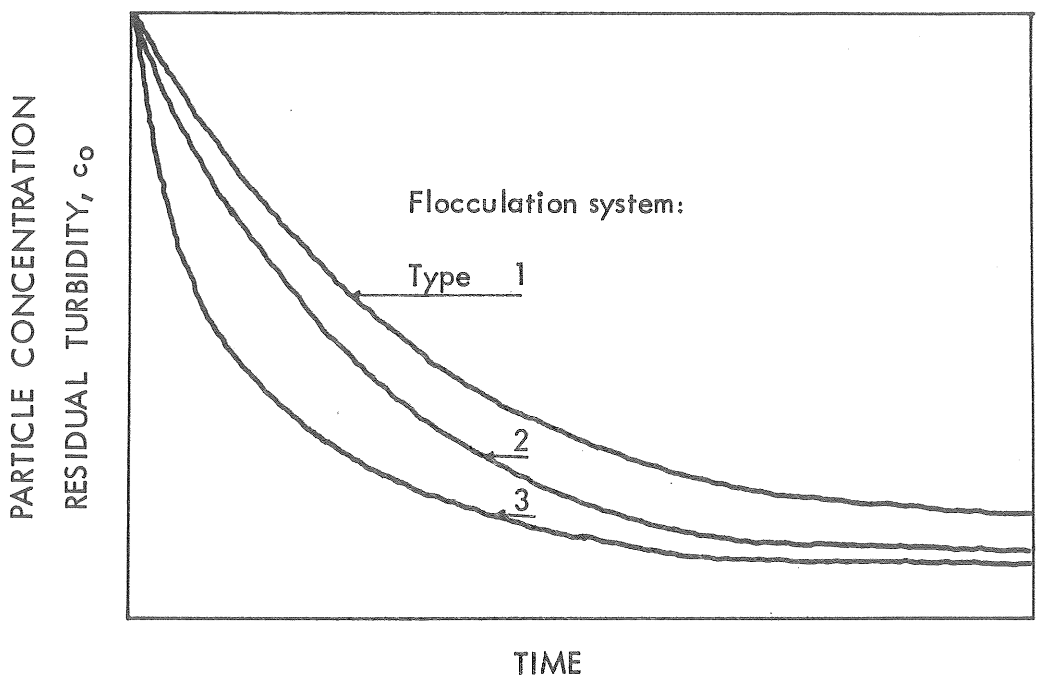


FIG. 1-3 Residual turbidity as a function of residence time in the flocculation unit.

The sedimentation unit design (overflow rate, surface load) can be calculated from the sedimentation analysis, and some typical results are illustrated in FIG. 1-4.

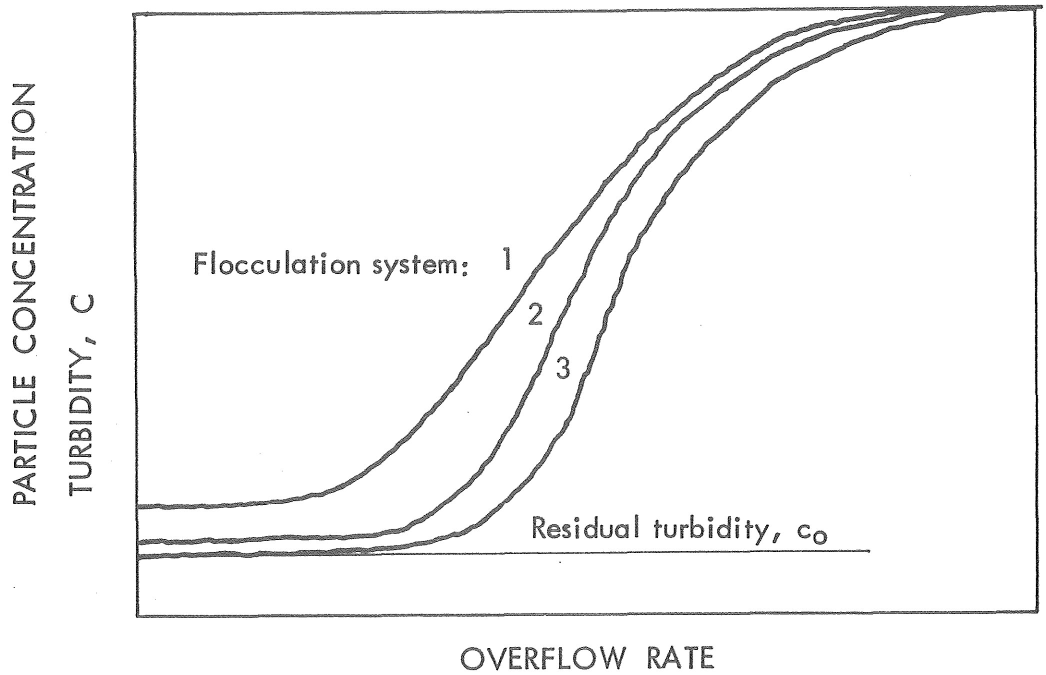


FIG. 1-4 Turbidity of effluent from the sedimentation unit as a function of overflow rate.

The turbidity steadily grows with increasing overflow rate, and the particles causing the turbidity have to be removed in the subsequent unit operation. The concentration of suspended particles in the water to be filtered is of great importance for the filter operation. If a certain run time is to be maintained for different filter designs, relationships concerning influent concentration and filtration rate may be of the type presented in FIG. 1-5.

Many parameters affect the unit operations, and as already pointed out, it is important to find the optimal design of the total water treatment procedure from a technical as well as an economic point of view, rather than an optimal design for each unit operation. The relationship between the performance of the individual unit operations in terms of particle concentration and the effective process volume is of fundamental importance. This particle-volume relationship of the unit operations is briefly shown in FIG. 1-6.

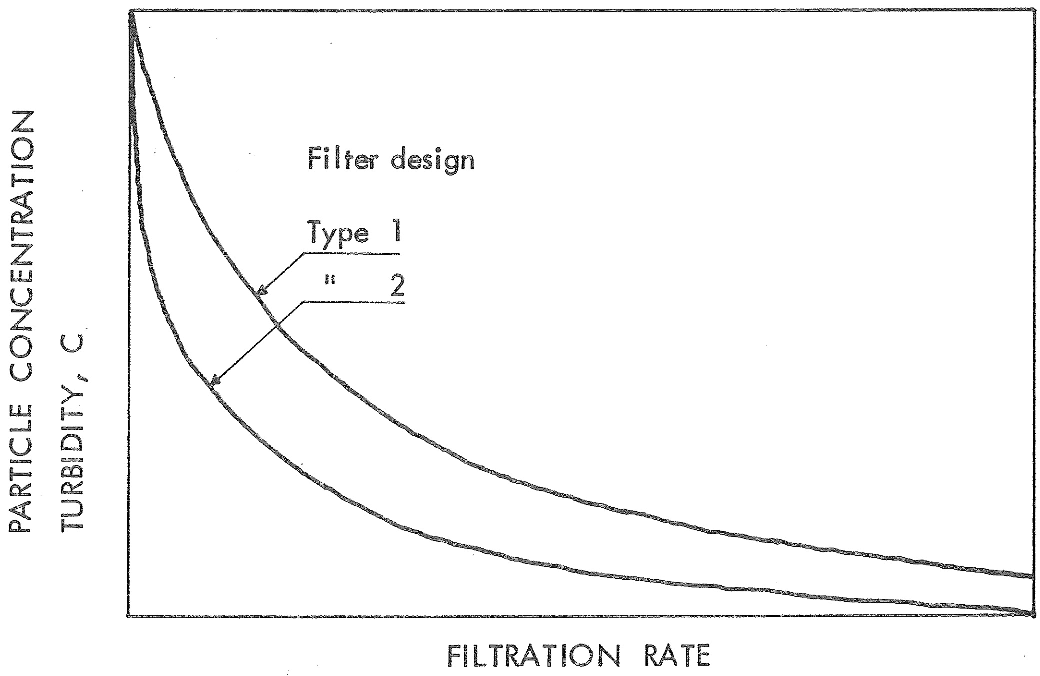


FIG. 1-5 Significance of the particle concentration of the influent for the magnitude of the filtration rate provided that a certain filter run time is to be maintained.

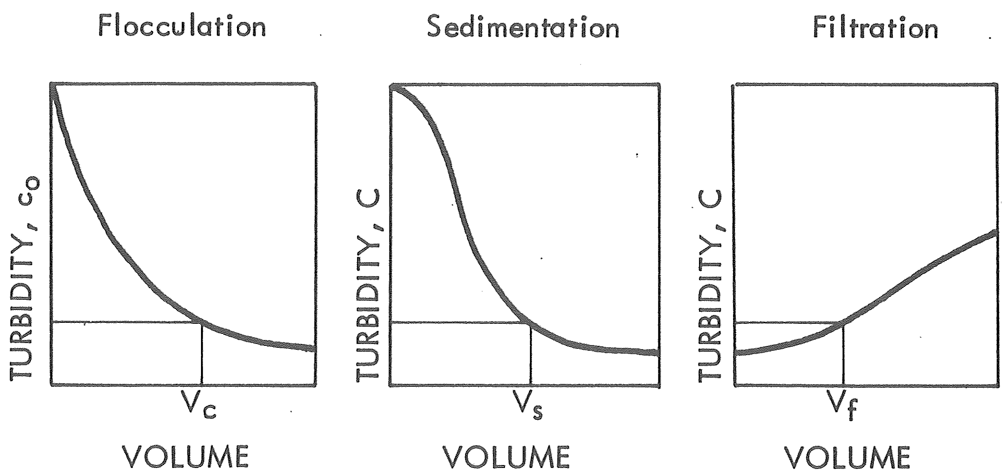


FIG. 1-6 The particle-volume relationship of the unit operations.

The optimization objective is to minimize the total cost of a water treatment plant, the investment cost as well as the running cost.

The total cost can be expressed as

$$\text{Total cost} = C_c \cdot V_c + C_s \cdot V_s + C_f \cdot V_f \qquad (1\text{-}1)$$

in which C_c is the cost of the flocculation unit per unit volume
C_s is the cost of the sedimentation unit per unit volume
C_f is the cost of the filtration unit per unit volume
V_c is the volume of the flocculation unit
V_s is the volume of the sedimentation unit
V_f is the volume of the filter unit

This thesis will focus on the relationship expressed in Eq. (1-1).

2 COAGULATION-FLOCCULATION, BACKGROUND

2.1 General

An important process in the treatment of water is aggregation of the colloids in the water, and formation of flocs, which can be effectively removed from the liquid by settling or filtration. This process is called coagulation. The various steps, both chemical and physical, are presented in FIG. 2-1 and TABLE 2-1.

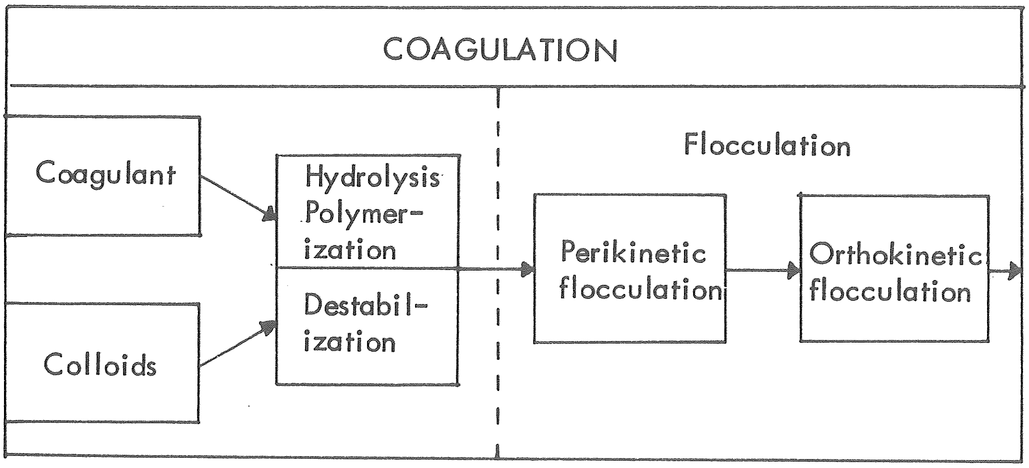


FIG. 2-1 Various steps in coagulation

TABLE 2-1 Schematic representation of the different steps in coagulation (after Hahn and Stumm, 1968).

Coagulant formation	Destabilization	Particle transport
Hydrolysis and polymerization reactions, when hydrolyzing metal ions are used as coagulant.	1. *Reduction of potential energy of interaction between particles.* (a) Compression of double layer by counter ions (Schultze Hardy) (b) Decrease in surface potential due to specifically adsorbed counter ions or due to surface reactions.	1. *Brownian motion* causes primarily collision of small colloids. (Perikinetic flocculation)
	2. *Bridge formation.* Specifically adsorbed polymeric species form bridges between colloids.	2. *Velocity gradients* transport large colloids, by imposing different velocities upon neighboring particles. (Orthokinetic flocculation)

Rate of Agglomeration = Collision efficiency factor x Collision frequency

2.2 Destabilization of colloids

Coagulants, coagulant aids, and the mixing conditions are of fundamental importance in the destabilization of colloids and determines the characteristics of the floc particles formed. The mechanisms of the destabilization have not been investigated in this study.

2.3 Perikinetic flocculation

The present knowledge of transport mechanisms is based upon the early (1917) work of Smoluchowski. The rate of change in the number of particles due to thermal motion or Brownian motion (perikinetic flocculation) can be represented as follows:

$$Jp = 4\,\pi \cdot D_{ij} \cdot r_{ij} \cdot n_i \cdot n_j$$

or (2-1)

$$Jp = \frac{2k \cdot T}{3\mu}\,(r_i + r_j)\,\frac{1}{r_i} + \frac{1}{r_j})\; n_i \cdot n_j$$

in which suffixes i and j indicate the number of primary particles comprising aggregates, called i-fold and j-fold particles.

J_p is the rate of collision of i-fold and j-fold particles per unit volume

n_i, n_j are the number of i-fold and j-fold particles per unit volume of fluid

r_i, r_j are radius of i-fold and j-fold particles

D_{ij} is the individual diffusion coefficient of i-fold and j-fold particles. $D_{ij} = \frac{k \cdot T}{6\mu}\,\frac{(r_i + r_j)}{r_i \cdot r_j}$

k is Boltzmann's constant

T is the absolute temperature

μ is the fluid viscosity (dynamic)

Equation (2-1) has been simplified for a monodisperse system by Swift and Friedlander (1964):

$$J_p = \frac{d\,n_t}{d\,t} = \frac{4\eta}{3\mu}\, k \cdot T\, n_t^2 \qquad (2\text{-}2)$$

in which n_t is the total concentration of particles at time t

η is the collision factor representing the fraction of the total number of collisions which are successful in producing aggregates

According to Hahn and Stumm (1968), Eq. (2-2) describes with sufficient accuracy the perikinetic flocculation even in heterogeneous suspensions. The rate of perikinetic flocculation is second order with respect to the number of particles and is independent of particle size within certain limits. The perikinetic flocculation is dependent upon the temperature.

When the particles have aggregated to be larger than 1 μ, perikinetic flocculation ceases to be significant, and particle collision must be induced by hydrodynamic movements, causing relative motion of the particles. This process is called orthokinetic flocculation.

2.4 Orthokinetic flocculation

For the particle collision due to hydrodynamic movements, Smoluchowski has developed a theory. According to him the rate of collision of two sizes of particles is proportional to the product of the particle concentrations, to the velocity gradient, and to the cube of the sum of the particle radii. Thus,

$$J_o = \frac{4}{3}\, n_i \cdot n_j\, (r_{ij})^3\, \frac{dv_x}{dz}$$

or (2-3)

$$J_o = \frac{4}{3}\, n_i \cdot n_j\, (r_i + r_j)^3\, \frac{dv_x}{dz}$$

in which J_o is the rate of collision of i-fold and j-fold particles per unit volume

$\frac{dv_x}{dz}$ is the velocity gradient (G)

Equation (2-3) is considered to be the basic equation describing orthokinetic flocculation and predicts the collision frequency of a particle system of uniform size or of a particle system of two sizes. It has, with certain assumptions, been simplified by Hudson (1965) as follows:

$$-\frac{dn_1}{dt} = \frac{\eta \, \phi \, \bar{G} \, n_1}{\pi} \tag{2-4}$$

and by O'Melia (1972) as follows:

$$-\frac{dn_t}{dt} = \frac{4 \, \Omega \, \eta \, \bar{G} \, n_t}{\pi} \tag{2-5}$$

in which n_1 is the number of primary particles per unit volume
n_t is the total number of particles per unit volume
η is the collision factor
ϕ is the volume fraction of flocs
Ω is the volume fraction of the total number of particles
$\bar{G}$ is the velocity gradient, mean value

The difference between the two equations is briefly that Hudson neglects the size of the primary particles in comparison with the floc particles, while O'Melia considers the suspension to be monodisperse.

As mentioned, the theories presented describe the flocculation in a system of particles of two sizes only. Harries *et al.*, (1966) have tried to overcome this shortcoming and have developed various rate equations for the flocculation process.

The preceding equations were developed for the case of laminar flow which is stable below G values of $32s^{-1}$. In turbulent conditions the particles are transported at random by local fluctuations in the over all fluid movement. Robinson (1964) suggested for turbulent conditions, an equation analogous with Eq. (2-1) and (2-3):

$$J_o = \frac{K^1 E}{\lambda_E^2} \, f\,(d)\; n_i \cdot n_j \tag{2-6}$$

in which E is the eddy kinematic viscosity

λ_E is the mixing length in the turbulent system, λ_E is defined as the mean distance which an eddy travels at right angles to the direction of flow before it has lost its original fluid content

K^1 is a constant

$f(d)$ is some function of the particle diameter

Another model for the flocculation process has been presented by Argaman and Kaufman (1970). They suggest that the principal objection to Eq. (2-3) stems from the fact that when calculating the mean velocity gradient, one does not take into account the length and time scales over which the turbulent fluctuations occur. Velocity gradients of a given length scale will not contribute significantly to the collision of particles that are either much larger or much smaller than this scale. Thus the mechanistic model developed by Smoluchowski for laminar flow conditions is not entirely applicable to flocculation under turbulent conditions. Argaman and Kaufman (1970) presented a model for flocculation based on the hypothesis that particles which are suspended in a turbulent fluid experience a random motion resembling that of gas molecules. The motion can be characterized by an appropriate diffusion coefficient that can be expressed in terms of the measurable hydrodynamic properties of the fluid motion. The authors suggest a rate of collision between primary particles, radii r_o, and flocs, radii $r_f (r_F >> r_o)$, which gives

$$-\frac{dn_o}{dt} = 4 \pi K_S \cdot r_F^3 \cdot n_o \cdot n_F \cdot \bar{u}^2 \qquad (2\text{-}7)$$

in which K_S is a proportionality coefficient expressing the effect of the turbulent energy spectrum on the diffusion coefficient. For a particular turbulence field and particle size, K_S is constant

$\bar{u}^2$ is the mean square velocity fluctuations

n_o is the number of primary particles per unit volume

n_F is the number of flocs per unit volume

The relationship between $\bar{u}^2$ and $\bar{G}$ is expressed as

$$\bar{u}^2 = K_P \cdot \bar{G} \qquad (2\text{-}8)$$

in which K_P is a performance parameter characterizing the stirring arrangement

Argaman and Kaufman's model for the rate of collision may be summarized as

$$-\frac{dn}{dt} = K_p \cdot K_S \cdot K_F \cdot \bar{G} \cdot n \qquad (2\text{-}9)$$

in which K_F is a constant expressing the floc volume per unit volume and the fraction of collisions which result in aggregation. (Compare Eq. (2-4) and (2-5).

According to Argaman and Kaufman (1970), the difference between Eq. (2-3) and the simplified Eq. (2-4) and (2-5) and Eq. (2-9) is that in Eq (2-9) a relation to the stirring arrangement is suggested, as it is regarded insufficient to characterize the flocculation process by the G-value only.

The rate of collision in a certain suspension increases in direct proportion to the velocity gradient. In turbulent flow rapid changes in velocity and inertia of the floc particles cause disruptions. The disintegrating force on aggregated particles in a shear field increases with increasing floc size at a higher rate than the attractive force. Therefore, large flocs are more likely to be broken than small flocs (Hannah *et al.*, 1967).

The exact breakup mechanism and the sizes of the fragments are not known. Argaman and Kaufman (1970) assumed, as also noted by Tomas (1964), that flocs in laminar flow have rotational motions which result in a rearrangement of particles within the floc with no significant breakup. Among the possible breakup modes, the stripping of individual primary particles from a floc surface is one of the most credible ones. The rate at which primary particles are released depends on the surface shear and the size of the primary particles. The above mentioned investigators have expressed this relationship as follows:

$$\left(\frac{dn_o}{dt}\right)_{Breakup} = K_B \, r_F^2 \, \frac{n_F}{r_o^2} \, \bar{u}^2 \qquad (2\text{-}10)$$

in which K_B is the breakup constant.

Lagvankar and Gemmel (1968) suggested multiple (at least three) levels of floc aggregation: (1) a disperse level, (2) flocs comprising primary particles, and (3) a more advanced level. The particles at the third, or floc-aggregate, level are thought to be loose aggregates in which the flocs are held together by mainly mechanical entanglement of the projections or tentacles extending randomly outward from the main floc body. If this is true, these floc aggregates are quite likely to be fragile and would have a tendency to break apart at the tentacle junctions. The resultant debris would, according to Lagvankar and Gemmel (1968), consist of flocs or floc aggregates of various numbers and sizes of particles.

Important in floc aggregation is the strength of the flocs in the different steps of the flocculation process. This strength is deemed to be dependent on the mode of floc formation. Flocculation at high velocity gradients produces a dense floc with good settling properties (Hudson and Wolfner, 1967). Lagvankar and Gemmel (1968) have noted, however, that the intensity of agitation provided during flocculation did not significantly affect the density of the floc of any given size. But they have shown, both theoretically and in laboratory studies, that the floc density generally decreases with the floc size. The explanation of the higher densities of small flocs may be found in the fact that large flocs either do not form or quickly break up under intense agitation. As small flocs are in fact denser than large flocs, different intensities of agitation simply produce different size-frequency distributions rather than directly affect the floc density.

A broken floc usually does not re-form to the same degree of flocculation because of similar internal neutralization of bonding forces. Apparent floc strength, as deduced from particle size measurement, may therefore decrease with age and agitation (Hannah *et al.*, 1967). Hydroxide floc alone is relatively weak and is apparently strengthened by incorporating solids into the floc. Coagulant aids have a pronounced effect upon the formation of large, strong flocs, but limited dosages must be employed as floc strength must be tailored to fit the overall treatment operation. This includes filtration which may be adversely affected by a very high floc strength.

To summarize, it can be concluded that there is a limiting relationship between particle size and velocity gradient. Weijman -Hane (1962) has schematically expressed flocculation as a reversible process:

$$F_1 + F_2 \rightleftarrows F_{12} \tag{2-11}$$

where the flocs F_1 and F_2 form the floc F_{12}. The reaction goes to the right when the floc strength is not exceeded.

Ritchie (1956), Camp (1955), Lagvankar and Gemmel (1968), Argaman and Kaufman (1970), and Ives (1973)consider the maximum floc size (radius r max) inversely proportional to the velocity gradient which can be expressed

$$r_{max} = \frac{K}{G} \tag{2-12}$$

in which K is a proportional constant

The equation is valid within certain limits.

The above mentioned clearly shows the importance of an even distribution of the energy input. Otherwise, particles formed at low velocity gradients will be broken when passing areas with high velocity gradients. The relationship expressed by Eq.(2-12) is probably too simple and is also difficult to apply practically due to the difficulty of determining the constant K.

Several investigators have suggested the product $G \cdot T$ as a characterization of the flocculation process. However, this product — which has been shown experimentally (Ives, 1973) — is probably too simple a characterization.

Ives concluded that tapered flocculation attains better clarification compared to uniform flocculation, whatever the mode of breakup of flocs. It is also more efficient when the initial size distribution is polydispersed. The relative efficiency of the tapered flocculation increases when the value of the velocity gradient, or initial concentration of particles, or both, is increased.

One of the objections to the basic equation describing orthokinetic flocculation has been that the particles do not follow those streamlines which would occur without particles in the water and on which the equation is based. The divergence is particularly noticeable when the difference in size of the particles is great. The small particles are deflected from the bigger ones. If this is considered in the model – only particles fairly alike in diameter are aggregated – the relative advantage of the tapered flocculation is decreased. However, the tapered flocculation attains a better clarification under all conditions, (Ives, 1973).

The rate of collision of particles is described by the general equation in which the velocity gradient must be adapted to the floc growth. In order to obtain information about the detention times necessary for a given reduction of the number of particles, one must know the reaction rate. Hahn and Stumm (1968) have carried out investigations in order to examine the reaction rates of the different steps in the process of successful aggregation. The result of this investigation shows that the reaction rates are very low for the perikinetic flocculation. The half-time of the reaction is of the magnitude of 10^{-3}s while the half-time of the reaction in orthokinetic flocculation is of the magnitude of some minutes. These conditions emphasize the importance of a rapid and effective mixing of chemicals if the goal is to decrease the flocculation time.

2.5 Reactors

In practice flocculation is carried out under conditions which more or less satisfy the theoretical models. The floc formation takes place in reactors consisting of basins equipped with different stirrers to create turbulence in the water. The design and the number of reactors are of great importance for the flocculation process.

For continuous flow systems there are two types of ideal reactors. A Completely Mixed Flow Reactor (CMF), and a Plug Flow Reactor (PF).

In a continuous, completely mixed flow reactor, the time distribution for the water is obtained by means of a material balance study, and the flocculation process for this type of reactor may be expressed as follows if Eq. (2-5) describing the collision rate is used.

$$Q\,n_o - Q\,n_1 - \frac{4}{\pi}\,\eta\,\Omega\,\bar{G}\,n_1 \cdot V = 0 \qquad (2\text{-}13)$$

in which $Q\,n_o$ is the input number of particles per unit of time

$Q\,n_1$ is the effluent number of particles per unit of time

$\frac{4}{\pi}\,\eta\,\Omega\,\bar{G}\,n_1 \cdot V$ is the number of successful particle collisions within the reactor

n_o is the number of particles in the influent per unit volume

n_1 is the number of particles in the effluent per unit volume

Q is the flow rate

V is the volume of the reactor

The equation can be rearranged as follows:

$$\frac{n_o}{n_1} = \frac{4}{\pi}\,\eta\,\Omega\,G \cdot T + 1 \qquad (2\text{-}14)$$

in which T is the mean residence time in the CMF reactor $(\frac{V}{Q})$

For a number m of CMF reactors in series, Eq. (2-14) can be written if the G-value is constant:

$$\frac{n_o}{n_m} = \frac{4}{\pi}\,\eta\,\Omega\,G \cdot \frac{T}{m} + 1 \qquad (2\text{-}15)$$

and for varying G-value

$$\frac{n_o}{n_m} = (\frac{4}{\pi}\,\eta\,\Omega\,G_1\,\frac{T}{m} + 1)(\frac{4}{\pi}\,\eta\,\Omega\,G_2\,\frac{T}{m} + 1)\ldots.(\frac{4}{\pi}\,\eta\,\Omega\,G_m\,\frac{T}{m} + 1) \qquad (2\text{-}16)$$

In an ideal PF reactor, bulk flow proceeds through the reactor in an orderly, uniform manner. The contents are uniform in the radial direction, and there is no mixing due to concentration gradients in the longitudinal direction. When the number of reactors m $\to \infty$ in the Eq. (2-15), an expression of the flocculation in a plug flow reactor is obtained ($n_\infty = n_1$):

$$\ln \frac{n_o}{n_1} = \frac{4}{\pi} \eta \, \Omega \, G \, T \qquad (2\text{-}17)$$

Equation (2-17) is valid for a constant G-value. If the G-value varies in the flow direction (x-direction), the equation can be expressed as follows:

$$\ln \frac{n_o}{n_1} = \frac{4}{\pi} \eta \, \Omega \, \frac{1}{v_x} \int_0^L G\,(x)\, dx \qquad (2\text{-}18)$$

in which v_x is the flow velocity in the x-direction

2.6 Comparison between different reactors

An analysis of the number of CMF reactors in series which are necessary in order to simulate a plug flow is presented in FIG.2-2. The value of $\pi/4\ \eta\ \Omega\ G$ in Eq. (2-14) has been chosen to T. The volume has been divided into a number of compartments. Compartmentalization of a continuous-flow system exerts a considerable influence on the flocculation performance, but there is no need for further compartmentalization of the six reactors in series.

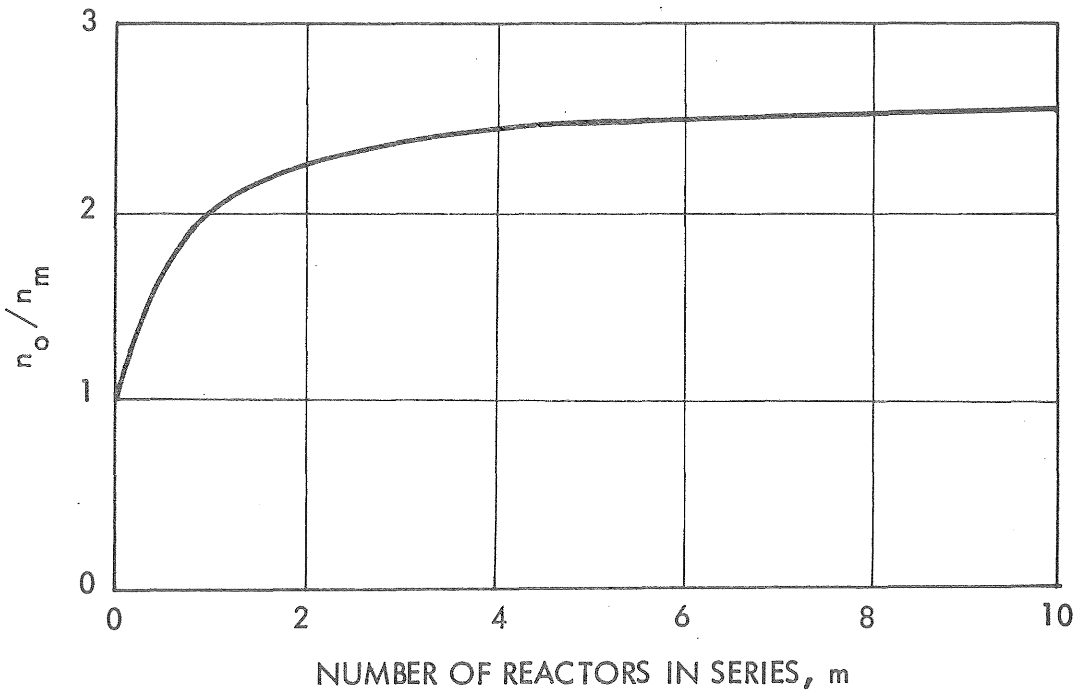


FIG. 2-2 Particle reduction as a function of the number of compartments

Apart from the number of compartments, the design of the single reactors is important. The extremes represented by the CMF reactor and the PF reactor are never fully realized in most full-scale process applications, although many designs closely approximate these ideals. Deviations can be caused by short-circuiting, by recycling, or by the presence of stagnant zones within the reactor; thus, it is necessary to determine experimentally the flow and mixing characteristics of the reactor. This information is most easily obtained by a stimulus – response technique using step or pulse inputs of a readily detectable tracer. Studies have usually dealt with time distributions in a sedimentation basin where a plug flow is the ideal. Flocculation basins normally have a time distribution similar to that of a completely mixed flow for the individual reactor; but, if the flocculation unit is compartmentalized by means of baffles, the time distribution differs from that of a completely mixed reactor.

The tracer-response curve for an ideal, completely mixed reactor can be expressed as

$$\frac{c}{c_o} = e^{\frac{t}{T}} \qquad (2\text{-}19)$$

in which c is the concentration of tracer at the time t.
c_o is the initial concentration of tracer
t is the actual time
T is the theoretical residence time

Several investigators have proposed methods for the analysis of the tracer curves or time distribution functions of a real system, Danckwerts (1953); Sailenmacher (1965); Villemonte *et al.*,(1967); Wolf and Resnick (1963); Levenspiel (1962).

Stein (Camp, 1955) has developed the equation for an instantaneous dispersion curve for m tanks in series as follows

$$\frac{c}{c_o} = \frac{m}{(m-1)!} \cdot \left(\frac{t}{T}\right)^{m-1} \cdot e^{-m\frac{t}{T}} \qquad (2\text{-}20)$$

This function is graphically presented in FIG. 2-3 for some values of m.

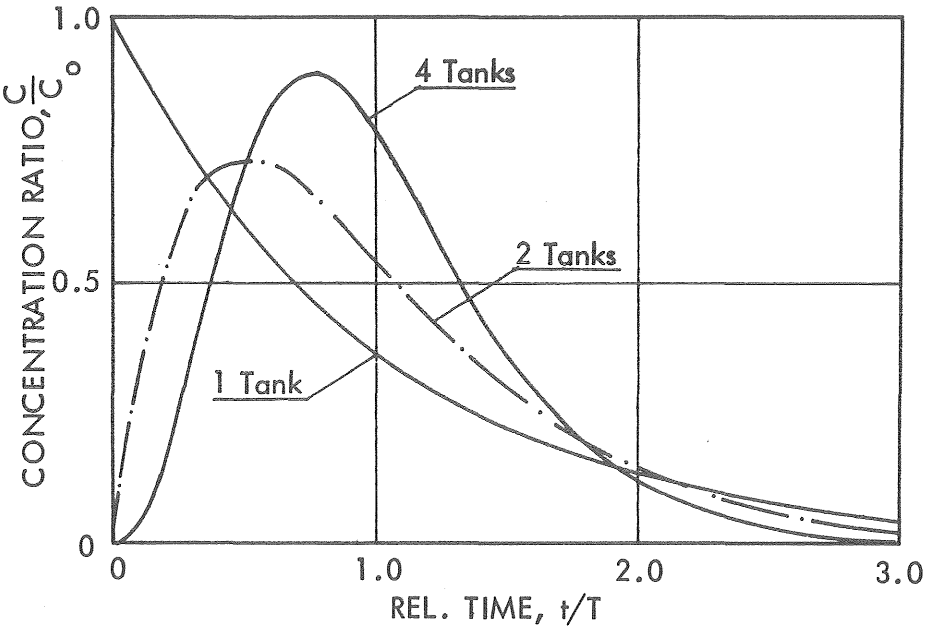


FIG. 2-3 Dispersion curves for tanks in series.

By comparing the tracer response and the theoretical curves, one obtains information of the flow distribution in a reactor. In practice, a somewhat smaller number of reactors in series may be used compared to the calculated number based on CMF reactors, since a reactor normally has a smaller amount of dispersion than an ideal CMF reactor.

2.7 Optimum design of flocculation units

FIGURE 2-4 shows the important factors in the flocculation process:

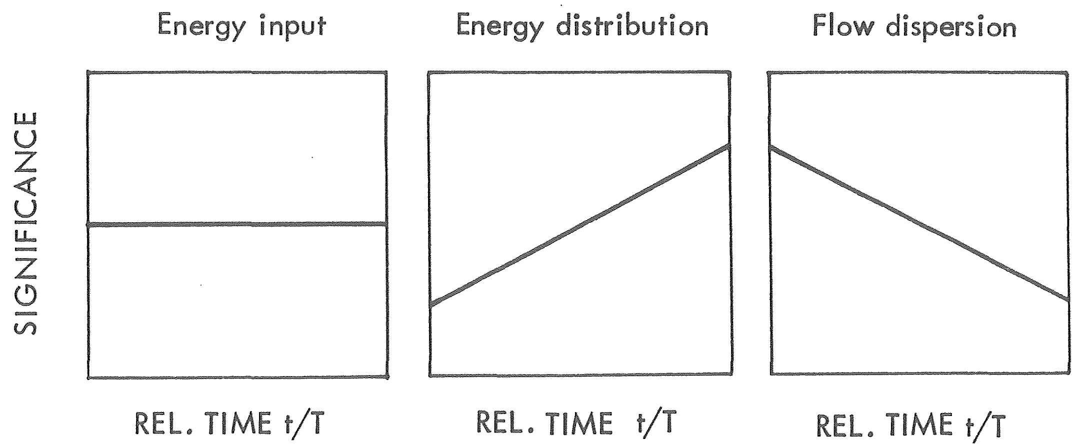


FIG. 2-4 Illustration of the mutual importance of different factors in the flocculation process.

It is important to have high velocity gradients in the initial stage of orthokinetic flocculation. A reactor with a high energy input is characterized as an almost completely mixed reactor, and the unfavorable flow distribution must be compensated with an increased compartmentalization. This can be done by a system of baffles. Later, the velocity gradient G must successively decrease in order to attain floc of increasing size without causing floc breakup. At low G-values (of course depending on the actual design of the stirring equipment), the flow distribution is nearer the plug flow. At this stage, the difficulties mainly consist of getting an even energy distribution in the reactor, and the design of the paddle also becomes important.

A theoretical consideration based upon simplified assumptions and the flocculation equations gives an idea of the suitability of varying the residence time in relation to the actual velocity gradient G during the flocculation process. This means that in this report, instead of a fixed value of the product G·T of the whole system, there is a suggested division of the product G·T in such a way that G·t is almost equal in each of the tanks in series. This implies that the residence time during the flocculation successively increases from one reactor to the following one.

Under idealized and controlled conditions, which in this case means plug flow and even energy distribution, this flocculation system can be exemplified by the following equations:

$$\ln \frac{n_o}{n_t} = 3 \ln \frac{r_t}{r_o} = \frac{4}{\pi} \Omega \, \eta \, \bar{G} \cdot t \qquad (2\text{-}21)$$

which yields

$$\log \frac{r_t}{r_o} = K_0 \, \bar{G} \cdot t \qquad (2\text{-}22)$$

in which r_t is the radii of the floc particles at time, t
r_o is the initial radius of the floc particle
K_0 is a constant

If a limiting relationship between maximum floc radii r and G-value is assumed valid (Eq. 2-12), the floc grows according to Eq. (2-22) as long as the energy input or velocity gradient is kept less than the level corresponding to r_{max}. When the velocity gradient is decreased, the floc starts to grow again but at a lower rate. This concept is illustrated in FIG. 2-5.

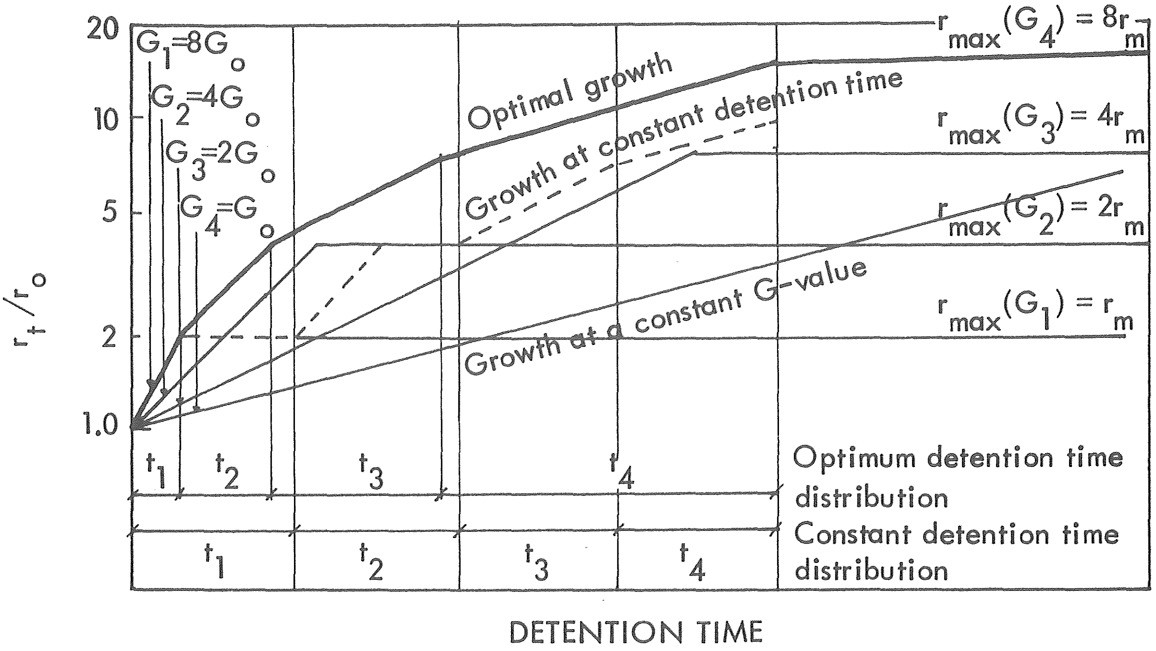


FIG. 2-5 Floc growth at different velocity gradients

In FIG. 2-5 the floc growth at four different G-values has been marked ($8G_o$, $4G_o$, $2G_o$, G_o). The maximum particle size will then be inversely proportional to the respective G-values (r_m, $2r_m$, $4r_m$ resp. $8r_m$). The value of the maximum floc radii at the velocity gradient of $8G_o$ has arbitrarily been chosen to be $2r_o$ which establishes the levels for the remaining r_{max}-values.

The chosen values imply an increase in floc radius of 16 times during the flocculation. This can be regarded as a reasonable assumption. In practice, it corresponds to an increase from a radius of 0.1 mm (just visible) to a radius of 1 mm. Calculations show that each floc contains about 4000 primary particles at this stage.

By combining the floc growth curves in the most favorable way, one obtains an optimum relationship between floc radii and flocculation time. In a conventional flocculation system, the duration at the different G-values is constant. If the flocculation time for the optimum growth (the upper curve) is divided into four equal time intervals, a growth according to the dashed curve is obtained. When the floc radius has increased 16 times under the most favorable conditions, the corresponding value of the floc radius increases 9 times, for the conventional flocculation system.

At the same time the floc radius under a constant velocity gradient, G_o, has only tripled.

In addition to the tank design, the design of the stirring equipment is important. As the G-value is only a measure of the mean energy input in the system, it is necessary to combine this value with some sort of measurement of the energy distribution. This can be done by a parameter K_j which is dependent on the paddle design. The following equation is suggested

$$K_j = \frac{Lp}{f(V)} \qquad (2\text{-}23)$$

in which Lp is the sum of the edge length of the paddles, at right angles to the direction of movement

f(V) is a function of volume

Other parameters in the paddle design have been suggested by other investigators. A literature review has been done by Hedberg (1974).

2.8 Mathematical model

2.8.1 General

In order to obtain information on the different factors affecting the flocculation process, one can use a mathematical model. A model cannot substitute practical experiments, but on the other hand it is necessary to show the most important factors in the flocculation process.

2.8.2 Basic model

Different opinions of the equation describing the rate of floc growth can be summarized as follows:

$$\frac{dn}{dt} = - K_1 \cdot n \cdot \bar{G} \qquad (2\text{-}24)$$

in which n is the number of particles per unit volume

t is the flocculation time

$\bar{G}$ is the mean velocity gradient

K_1 is a constant

Various compositions of the constant K_1 have been suggested in Eq. (2-4); (2-5); (2-9).

Equation (2-24) does not consider the simultaneous breakup of flocs. In the literature, two modes of floc breakup have been discussed. Ham and Christman (1969) state that the breakup is due to collision effects. Argaman and Kaufman (1968) and Parker *et al.* (1971) have assumed that breakup is due to surface erosion of flocs.

The floc breakup rate equation due to collision can be based upon the turbulence collision frequency equation in the same way as the floc growth, Eq. (2-24)

$$\frac{dn}{dt} = \beta \; K_1 \; n \; \bar{G} \qquad (2\text{-}25)$$

in which β is the fraction of collision between primary particles and flocs that lead to breakup

Several investigators have demonstrated that the floc size is inversely proportional to the velocity gradient:

$$d_{max} = \frac{K}{\bar{G}} \qquad (2\text{-}26)$$

in which d_{max} is the maximum stable floc size diameter limited by floc breakup

K is the floc strength constant

The fraction of collisions that lead to breakup is assumed proportional to the velocity gradient (Ødegaard, 1975) as follows:

$$\beta = K_b \cdot \bar{G} \qquad (2\text{-}27)$$

and the collision floc breakup rate equation can be written

$$\frac{dn}{dt} = K_b \; K_1 \; n \; \bar{G}^2 \qquad (2\text{-}28)$$

in which K_b is the collision floc breakup coefficient.

Parker *et al.* (1971) have assumed that the floc breakup is caused by surface erosion. The maximum stable floc size and the kinetics of the release of primary particles resulting from floc breakup, have been analyzed for two alternative modes and two turbulence regimes. The result for the maximum stable floc size limited by floc breakup has been proposed to be (see also Eq. 2-26).

$$d_{max} = \frac{K}{\bar{G}^s} \qquad (2\text{-}29)$$

in which K is the floc strength constant
s is the stable floc size exponent

They found that the expression for the primary particle erosion rate could be generalized as:

$$\frac{dn}{dt} = K_B \, \Omega \, \bar{G}^p \qquad (2\text{-}30)$$

in which K_B is the floc breakup rate coefficient
Ω is the floc concentration per unit volume
p is the floc breakup exponent

According to Parker *et al.*, the floc breakup is supposed to be caused either by erosion of primary particles from the floc surface or by breakage of flocs into floc fragments.

The theoretical values for the size exponent s and the floc breakup rate exponent p are shown in TABLE 2-2.

TABLE 2-2 Floc breakup exponents (after Parker *et al.*, 1971)

Floc breakup mode and hydraulic regime	s Stable floc size exponent	p Floc breakup rate exponent
Surface shearing		
Inertial convection subrange	2	4
Viscous dissipation subrange	1	2
Filament fracture		
Inertial convection subrange	1/2	-
Viscous dissipation subrange	1/2	-

It must be pointed out that it is quite uncertain whether the given constants and exponents in a model of the flocculation process really can be assumed to be constant. It is, however, necessary to describe the process with some sort of mathematical model in order to optimize the process.

To summarize, two different overall flocculation rate equations can be expressed depending on the assumed mode of breakup.

Collision breakup:

$$\frac{dn}{dt} = -K_1\ n\ G + K_2^1 \cdot K_1\ nG^2 = K_1\ n\ G\ (K_2^1\ G - 1) \qquad (2\text{-}31)$$

Surface erosion breakup:

$$\frac{dn}{dt} = -\ K_1\ n\ G\ +\ K_2\ G^p \qquad (2\text{-}32)$$

in which K_1 is a constant for floc growth
K_2 is a constant for floc breakup
K_2^1 is a constant for floc breakup

If the flocculator consists of a completely mixed identical reactor (CMF reactor) with a residence time t a material balance equation for the surface breakup equation would be:

$$Q\ n_o - Q\ n_1 - (K_1 \cdot n_1\ G - K_2\ \bar{G}^p)\ V = 0$$

or (2-33)

$$\frac{n_1}{n_o} = \frac{1 + K_2\ \frac{\bar{G}^p}{n_o} \cdot t}{1 + K_1\ G \cdot t}$$

in which Q is the flow rate
n_o, n_1 are the number concentration of primary particles entering the reactor and the concentration of particles leaving the reactor
V is the reactor volume
t is the residence time in the single reactor

If the flocculator consists of m CMF reactors with a total residence time of T, the following expression can be derived:

$$n_i = n_{i-1} \left[\frac{1 + K_2 \frac{\bar{G}_i{}^p}{n_{i-1}} \frac{T}{m}}{1 + K_1 \, \bar{G}_i \cdot \frac{T}{m}} \right] \qquad (2\text{-}34)$$

This equation may be modified with regard to the proposed ideas of successively increasing the flocculation time from one reactor to another and the hydraulic conditions in the individual reactors:

$$n_i = n_{i-1} \left[\frac{1 + K_2 \frac{\bar{G}_i{}^p}{n_{i-1}} R_i \frac{T}{F_i}}{1 + K_1 \, \bar{G}_i R_i \frac{T}{F_i}} \right]^{F_i} \qquad (2\text{-}35)$$

in which n_i is the number concentration of the particles leaving the reactor i

K_1 is a coefficient for floc growth

K_2 is a coefficient for floc breakup

$\bar{G}_i$ is the mean velocity gradient in the reactor i

R_i is the relation between the residence time in reactor i and the total residence time

T is the total residence time

F_i is a coefficient describing the hydraulic conditions in the reactor ($F=1$ for a CMF reactor, $F = \infty$ for PF reactor)

p is the floc breakup exponent

The above mentioned equations only give information on the particle reduction during the flocculation. It is also necessary to have information on the floc size distribution in order to describe the following separation operation. Thus, it is necessary to combine the flocculation rate equation with an equation which gives information about the size distribution of flocs.

2.9 Settling velocity of particles

A particle or aggregate can be considered to be separated from the water to be treated when it adheres to a surface under a statically irreversible condition or alternatively when entering a collecting zone. Therefore, separation must be considered under dynamic con-

b) increase the density of the particles

c) have a spherical-shaped particle of regular dimensions

In practice it is impossible to calculate the settling velocity of particles; therefore, the settling velocity must be analyzed by some sort of standardized procedure.

Evaluation of the settling characteristics of a suspension can be accomplished in a column where the suspension is allowed to settle

ditions. A first force field is due to gravity. A given particle can move in the direction of this field only if the resultant of the gravitational forces is sufficiently high compared to opposing forces: friction, turbulence, electrostatic repellent force, convection, etc.

When particles are spheres of small size, viscosity is a predominant factor and Stoke's law can be applied:

$$v_f = \frac{d^2 g}{18 \mu} \{\rho_s - \rho\} \qquad (2\text{-}36)$$

in which d is the particle diameter

μ is the viscosity (dynamic)

ρ_s is the density of the particle

under quiescent conditions. The concentration of the particles is determined by samples withdrawn from several depths at different time intervals. The particle concentration as a function of time and the corresponding calculated settling velocity can be illustrated as in FIG. 2-6, in which some definitions describing the curves are suggested.

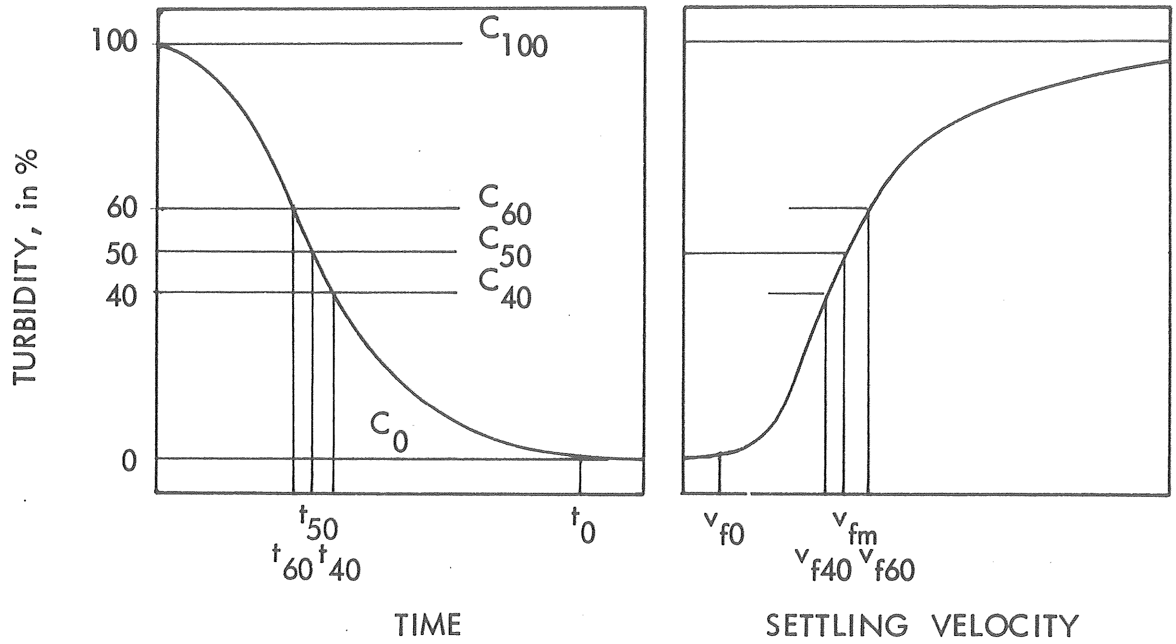


FIG. 2-6 Settling velocity analysis and settling velocity distribution curve

Earlier investigations (Rosén, 1967) have shown that the settling distribution curve is approximately a Gaussian curve, if the "tails" of the distribution curve are truncated. The floc properties can therefore be characterized by the mean settling velocity v_{fm} the standard deviation σ and the residual concentration c_0. The frequency equation of the settling velocity of the flocs is:

$$f(v_f) = \frac{1}{\sqrt{2\pi}} \exp - \frac{(v_f - v_m)^2}{2\sigma^2} \qquad (2\text{-}38)$$

in which v_f is the settling velocity of the flocs
v_m is the mean settling velocity of the flocs
σ is the standard deviation of the settling velocity of the flocs

The calculation of the standard deviation is normally done by transposing the time turbidity curve and plotting the result on a Gaussian distribution diagram. Rosén (1969) has developed a simplified method by using the inclination of the tangent to the points representing 60 % and 40 %, respectively, of the suspension concentration in the settling velocity distribution curve. The difference between the settling velocities v_{f60} and v_{f40} is then multiplied by 2, which gives a value of the standard deviation in agreement with the result obtained from Gaussian distribution diagrams. It is, however, very important when the standard deviation is calculated for suspensions with varying mean settling velocities, to adjust the time scale. This complicates the use of this method for suspensions with different mean settling velocities. The standard deviation may instead be calculated directly from the time – concentration relationship curve.

As alternatives to an approximate Gaussian distribution of the settling velocity, other distribution functions may be possible. In the studies of the settling properties of the floc, the following typical examples have been found, FIG. 2-7.

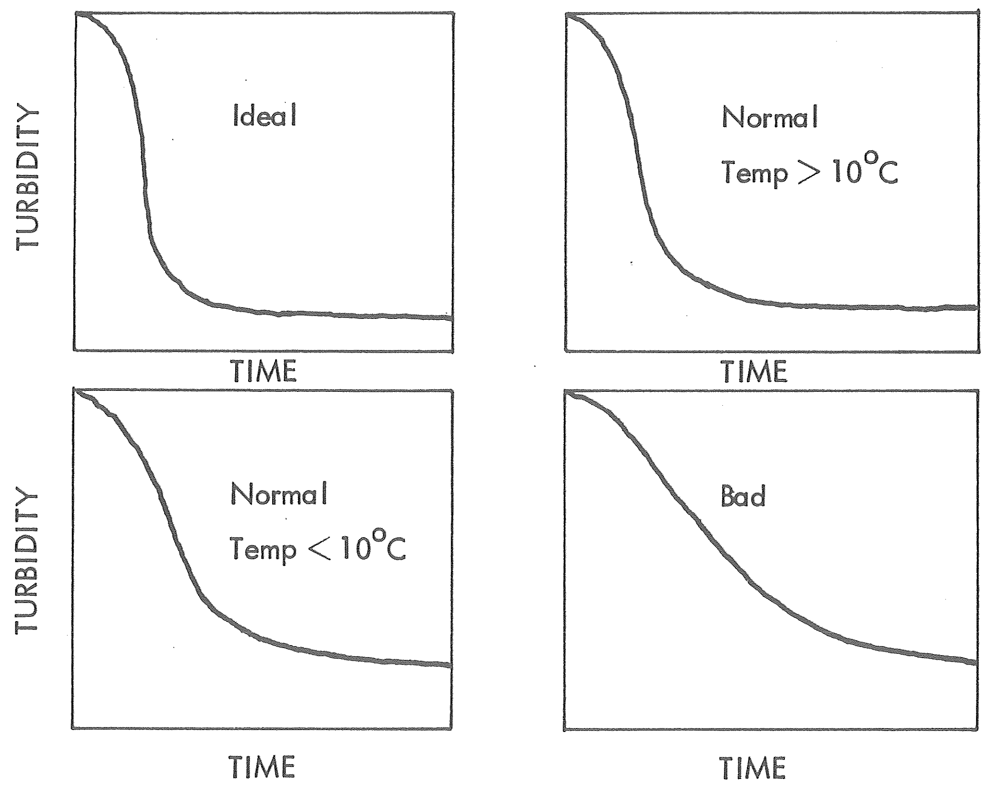


FIG. 2-7 Typical examples of settling analysis curves

From the result obtained the following hypothesis of the relationship between the residual number of particles leaving the flocculation unit and the settling characteristics of the floc can be written:

$$v_{fo} = f(n_i)$$
$$v_{fm} = f(n_i) \qquad (2\text{-}39)$$
$$\sigma = f(n_i)$$

in which n_i is the number concentration of particles leaving the flocculation reactor i

Normally the concentration of particles is measured in turbidity units and the Eq. (2-39) can be written

$$v_{fo} = f(c_{o,i})$$
$$f_{fm} = f(c_{o,i}) \qquad (2\text{-}40)$$
$$\sigma = f(c_{o,i})$$

in which $c_{o,i}$ is the residual turbidity of the suspension after a relatively long sedimentation time. Suffix i indicates the number of the reactors

A relationship according to Eq. (2-39) or (2-40) would, in combination with Eq. (2-35), give complete information of the flocculation-sedimentation operation.

3 COAGULATION — FLOCCULATION, EXPERIMENTAL INVESTIGATION

3.1 Objectives of the experimental investigation

The main objective of the experimental investigation was to quantify at least some of the factors affecting the flocculation process in order to be able to describe mathematically the process for optimization purposes.

3.2 Scope of the investigation

The following variables have been studied:

- Temperature
- Optimum pH-value
- Design of reactors
  - Geometrical design
  - Energy input
  - Rate of flow through the reactor
  - Direction of flow through the reactor
  - Paddle design
- Principle of the distribution of the total reactor volume

These variables have been studied on different scales ranging from laboratory experiments to full scale operation, and in the following the results from these experiments are exemplified by graphs. For further information see Hedberg (1969; 1974; 1975) and Appendix 3.

3.3 Effect of temperature on flocculation

Other investigations (Mohtadi and Rao, 1973) have shown that, when alum is used as a flocculant, the rate of flocculation is temperature dependent. The flocculation is, however, virtually independent of temperature when a polymer (Purifloc) is used as a flocculant. In those experiments a constant velocity gradient was used throughout the whole flocculation period, which is a doubtful experimental procedure. In the investigations carried out here, the experimental technique has, as far as possible, been similar to practice. The result of the flocculation was measured indirectly by turbidity analyses of the suspension after a certain sedimentation time. The disadvantage of this method is that both the effect of the flocculation and the effect of the sedimentation are simultaneously obtained. The laboratory result is shown in FIG. 3-1.

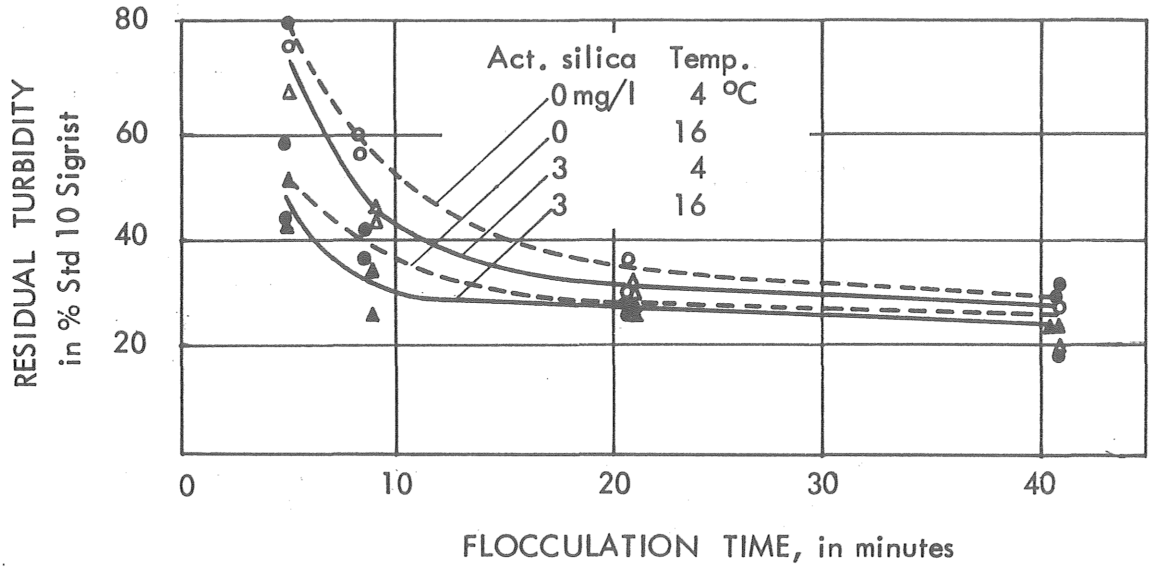


FIG. 3-1 Residual turbidity as a function of flocculation time. Temperature 4 and 16°C, activated silica 0 and 3 mg/l. Alum 45 mg/l, pH-value 6.4-6.5. Sedimentation time 30 minutes. Sedimentation depth 0.10.

The result is difficult to interpret due to the simultaneous effect of flocculation and sedimentation, but there is a temperature effect especially at shorter residence times. Poor flocculation in winter is often mentioned in the literature. The high viscosity of the water at low temperatures descreases the G-value and thus the rate of floc formation. Also, due to the high viscosity, the internal shearing stresses in the floc will be greater in the winter than in the summer so that the floc formed will be small and settle slowly at low temperatures. In practice, this is normally compensated for by adding a coagulant aid. However, due to the different seasons, the raw water characteristics may also vary, and if experiments are carried out with natural waters, a more marked temperature effect on the rate of the flocculation has been noticed (FIG. 3-2).

Tests performed in full scale have shown similar results to those obtained in laboratory experiments and in pilot plants. The results from the full-scale experiments are shown in FIG. 3-3.

FIGURE 3-3 shows the results from flocculation units with a

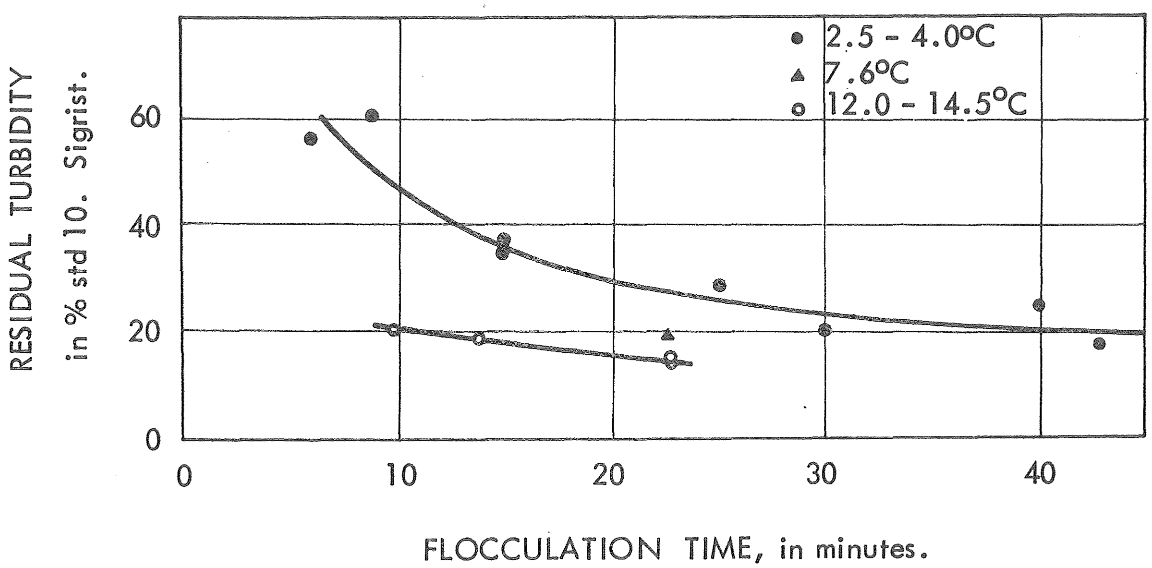


FIG. 3-2 Residual turbidity as a function of flocculation time. Alum 40 mg/l. Activated silica 4-6 mg/l, pH-value 6.0-6.5. Pilot plant 5. Appendix 1.

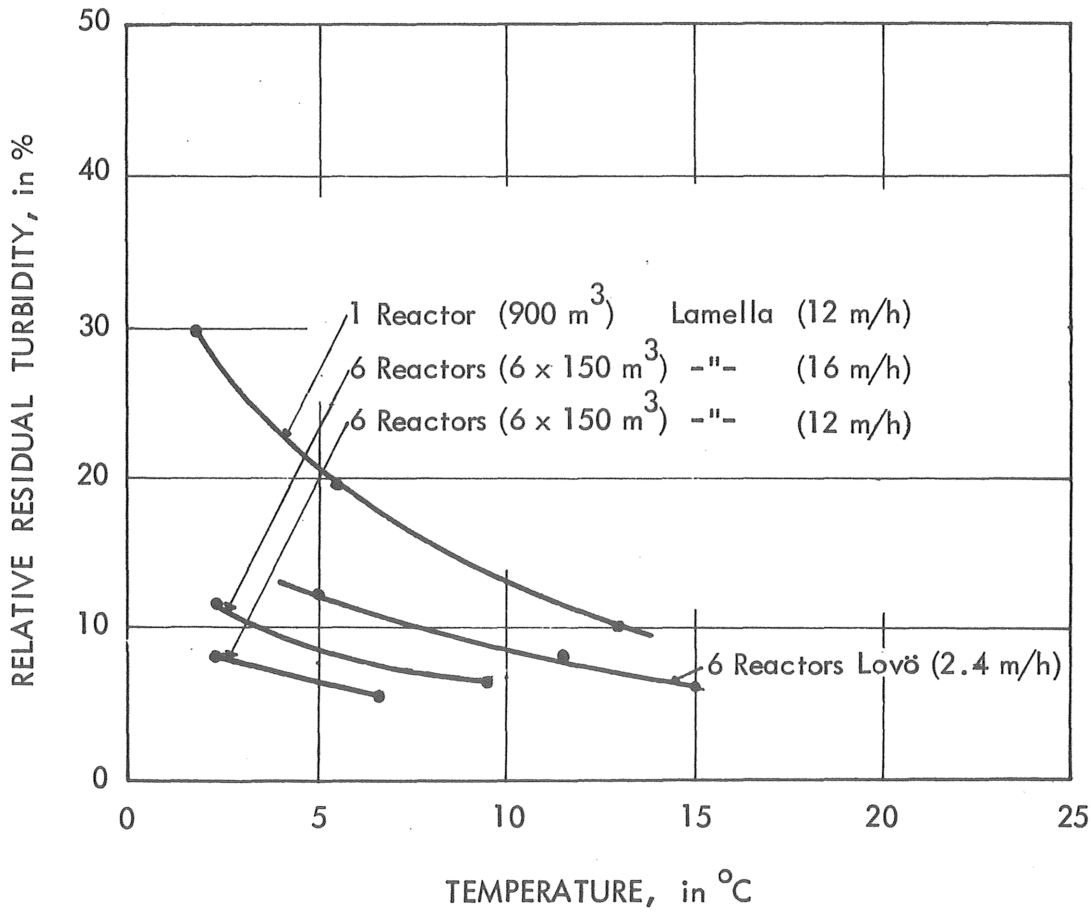


FIG. 3-3 Residual turbidity as a function of temperature. Mean values, Full scale 7, 6, and 8. Appendix 1.

different number of reactors, and it can be seen that in the summer almost the same turbidity in the effluent from the sedimentation basins is obtained for the two flocculation systems, while in the winter the turbidity is greater in the system with only one reactor. The detention time distribution in the system with only one reactor is very close to that of a completely mixed reactor, while the detention time distribution in the system with six reactors in series has a relatively small amount of dispersion.

However, even during flocculation times above 20-30 minutes, the effect of temperature upon the floc formation can be observed. The temperature effect measured as residual turbidity is not changed in proportion to the change in water viscosity according to Stoke's law, which would be the case if the floc size was unchanged.

Tesarik (1967) has investigated sludge-blanket clarifiers and has shown that the upflow velocity in a clarifier is inversely proportional to the cube root of the kinematic viscosity, provided that the particle size and density are invariable with temperature. This means that the influence of temperature on the settling velocity is lower than predicted by Stoke's law.

In the present investigation the opposite has been observed, but the concentration of suspended solids is normally higher in a sludge-blanket clarifier than in a conventional flocculation-sedimentation system.

Reed and Murphy (1969) have shown that the influence of temperature on activated sludge settling decreases as the concentration increases.

In the present investigation some preliminary tests have been carried out where sludge from the sedimentation unit was recycled to the flocculation unit. The tests were performed in a pilot plant (Pilot plant No. 3, Appendix 1), and sludge was added in the following concentrations: 0 %, 2 %, and 4 % of the flow. The dry solid content in the sludge added was between 0.61 to 0.78 %. The tests were carried out, at temperatures ranging from 8-12^{o}C. The result is shown in FIG. 3-4.

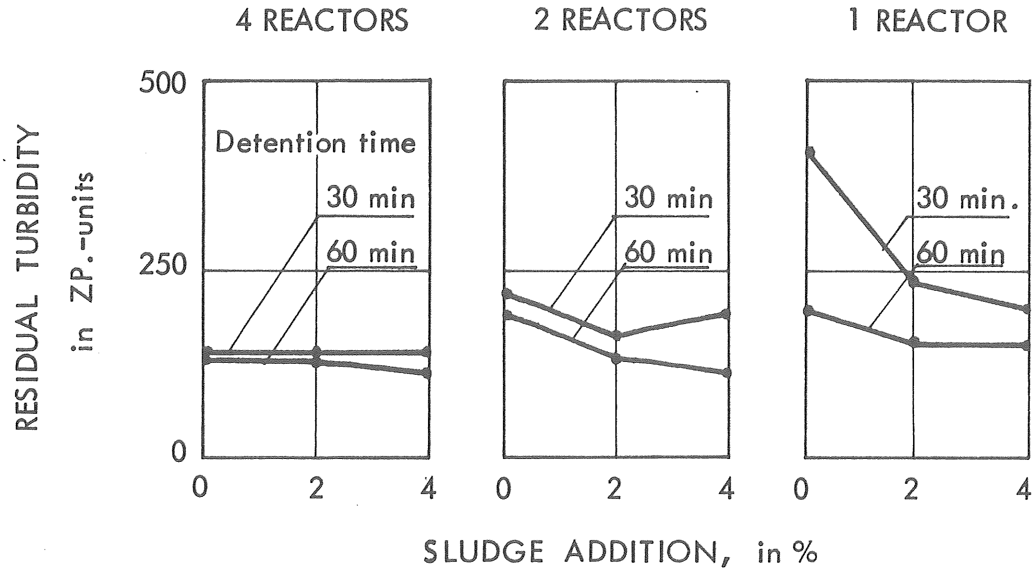


FIG. 3-4 Residual turbidity as a function of particle concentration

The result showed an increase in removal efficiency when sludge recycling increased. The tests are, however, too few to give any quantative results. (Appendix 4.)

The chemicals added have a significant effect on the temperature dependence of the flocculation. Thus, the results shown in FIG.3-2 and FIG. 3-3 are valid when a coagulant aid is added (activated silica 5 to 10 mg/l). Some tests carried out in conjunction with the filtration tests (see chapter 7) at an extremely low temperature (1 to 2°C) and with varying dosages of activated silica, showed a marked temperature dependence of the flocculation, especially when the activated silica addition was very low or completely omitted. The result is shown in FIG. 3-5 (Pilot plant No. 3).

From the studies carried out at varying temperatures, it can be concluded that the influence of temperature, measured in turbidity units after sedimentation, decreases with increasing particle concentration and activated silica addition. It can be assumed that the particle size and also the particle shape are influenced by temperature.

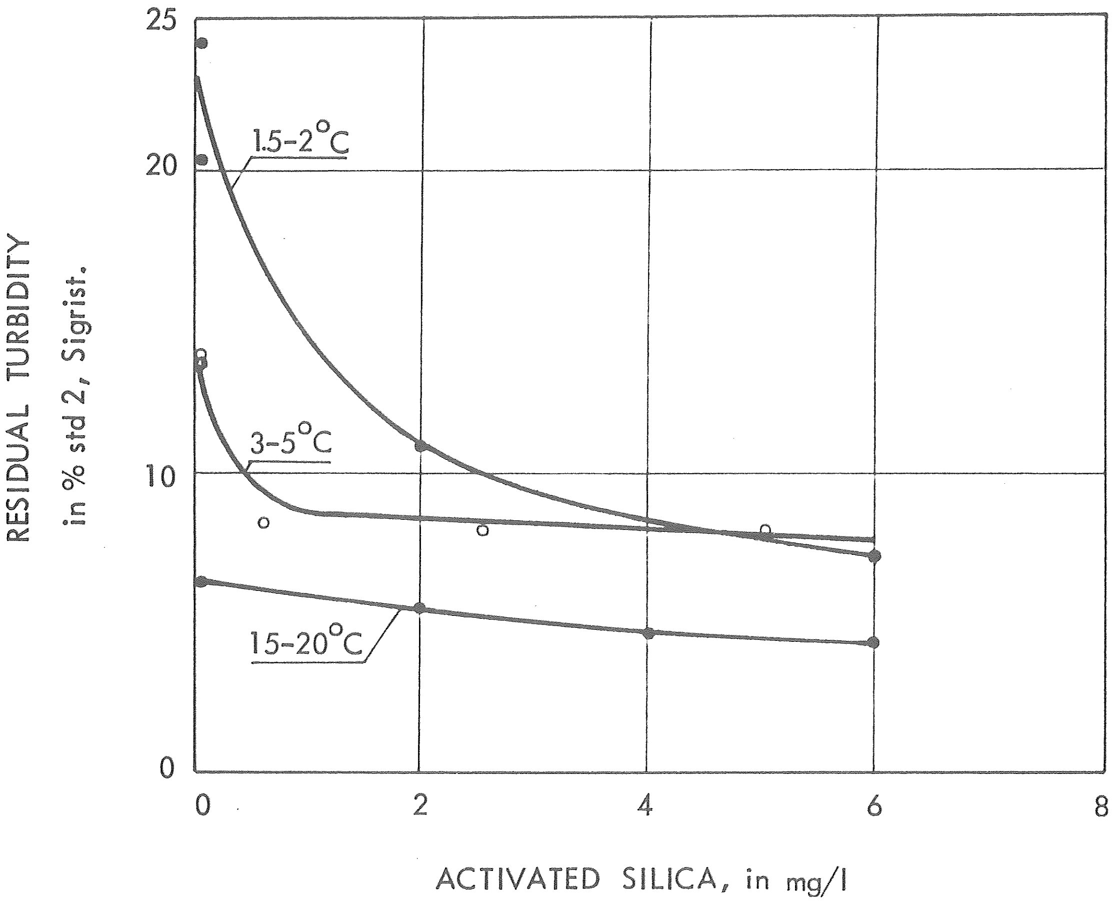


FIG. 3-5 Residual turbidity after sedimentation at a low overflow rate (0.2 m/h) as a function of activated silica addition (Pilot plant No. 3)

3.4 Effect of temperature on the optimum pH-value

An increased residence time in the flocculation reactor decreases the sensitivity to changes in pH-value. In general, the optimum pH-value is some tenths lower at a high temperature than at a low temperature. In the investigations, for example, the lowest residual turbidity was obtained at pH 6.2-6.3 at a water temperature of 15-16°C and at pH 6.4-6.6 at a water temperature of 4°C. The change in the optimum pH-value is affected by the raw water characteristics and by the type and amount of chemicals used.

3.5 Design of reactors

Several design parameters affect the result of the flocculation. By studying the detention time distribution curves from a reactor, one obtains information on the hydraulic conditions in a single reactor and in several reactors in series. The distribution function of a reactor is determined by means of a tracer, normally sodium chloride. The theoretical expression of the distribution function may be calculated (Eq. 2-19) and compared with the result obtained by tracer studies. A discrepancy between theory and practice sometimes occurs, however, and as several factors affect the hydraulic conditions in a reactor, it is important to know to what extent each contributes. In the following, examples are presented, showing the influence of geometry, velocity gradients, rate of water flow, and baffles. The examples represent results obtained in pilot and full scale plants.

3.5.1 Geometrical design of the reactor

Investigations carried out on flocculation units of different designs and on different scales have shown that the geometry does not significantly affect the hydraulic conditions in the relatively normal units which have been used. (Appendix 1). The geometry can be characterized by the width-depth ratio. In the present study, however, this relationship has been almost constant.

3.5.2 Velocity gradient

A flocculation reactor will naturally act more like a completely mixed flow reactor as the stirring intensity increases. This relationship is illustrated in FIG. 3-6.

The velocity gradient is of great importance to the hydraulic conditions. A high G-value (100 s^{-1}) implies that the time distribution is similar to that of a completely mixed flow reactor while a low G-value (12 s^{-1}) implies that the reactor corresponds to about 2.5 ideal CMF reactors. The G-values mentioned are of the same magnitude as those used in practice in the beginning and at the end of the flocculation process. Thus, it is obvious that the hydraulic conditions even in identically shaped reactors may be quite different, and it is necessary to consider this in the mathematical model of the flocculation process.

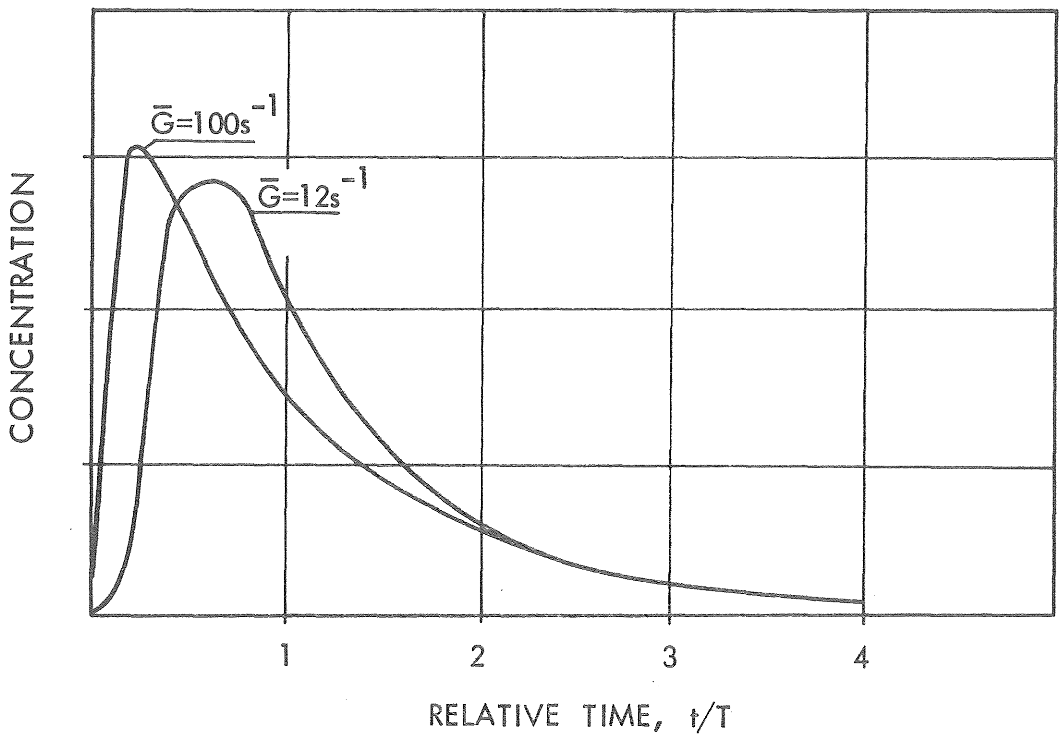


FIG. 3-6 The influence of the velocity gradient on the distribution curve

3.5.3 Rate of flow through the reactor

With decreasing residence time (increasing rate of flow) the tendency to remix the water decreases and the amount of dispersion becomes smaller, which is illustrated in FIG. 3-7 for different paddle designs; (see section 3.5.6). Thus, an increased rate of flow has a positive effect on the flocculation process and counteracts the negative effect of the simultaneously decreasing residence time. During some experiments a minimum residual turbidity has been observed at a certain "optimum" flocculation time. The difference in residual turbidity between flocculation at a long residence time and at an "optimum" residence time is, however, small.

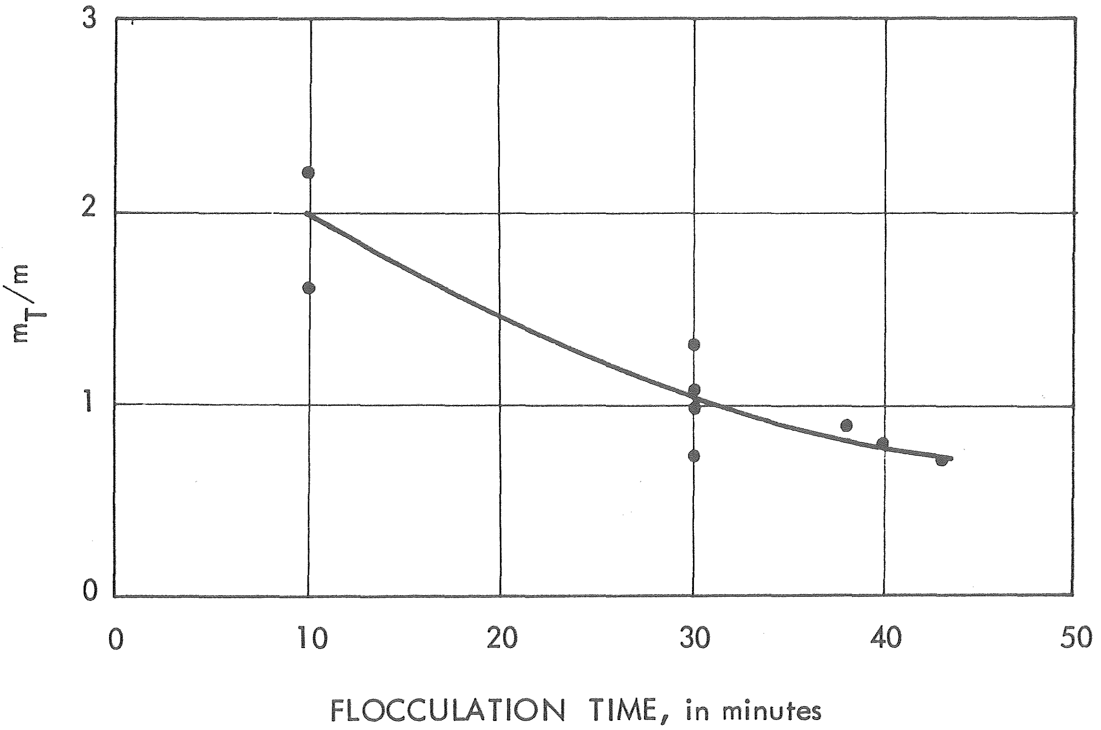


FIG. 3-7 Relationship between m_T/m (equivalent calculated number of CMF-reactors/actual number of reactors) and the mean residence time. Different tests in pilot plant No. 5

3.5.4 Baffles

The installation of baffles in a reactor has a marked effect on the detention time distribution curve. In FIG. 3-8 the discrepancy from the theoretical calculation is great due to the baffles installed in the flocculation unit.

The test was performed at the Lackarebäck Water Treatment Plant. (The plant is described in Appendix 1). The extra compartmentalization of the volume attained by the horizontal baffles gives the reactor a small amount of dispersion. The total flocculation unit is thus divided in 18 unit volumes, but as each volume has a smaller dispersion than a CMF reactor, the time distribution curve of the flocculation system has a dispersion smaller than that obtained in 18 CMF reactors in series. But as shown earlier (FIG. 2-2), the advantages with compartmentalization of a flocculation unit into more than 6 compartments are of little significance for the flocculation performance.

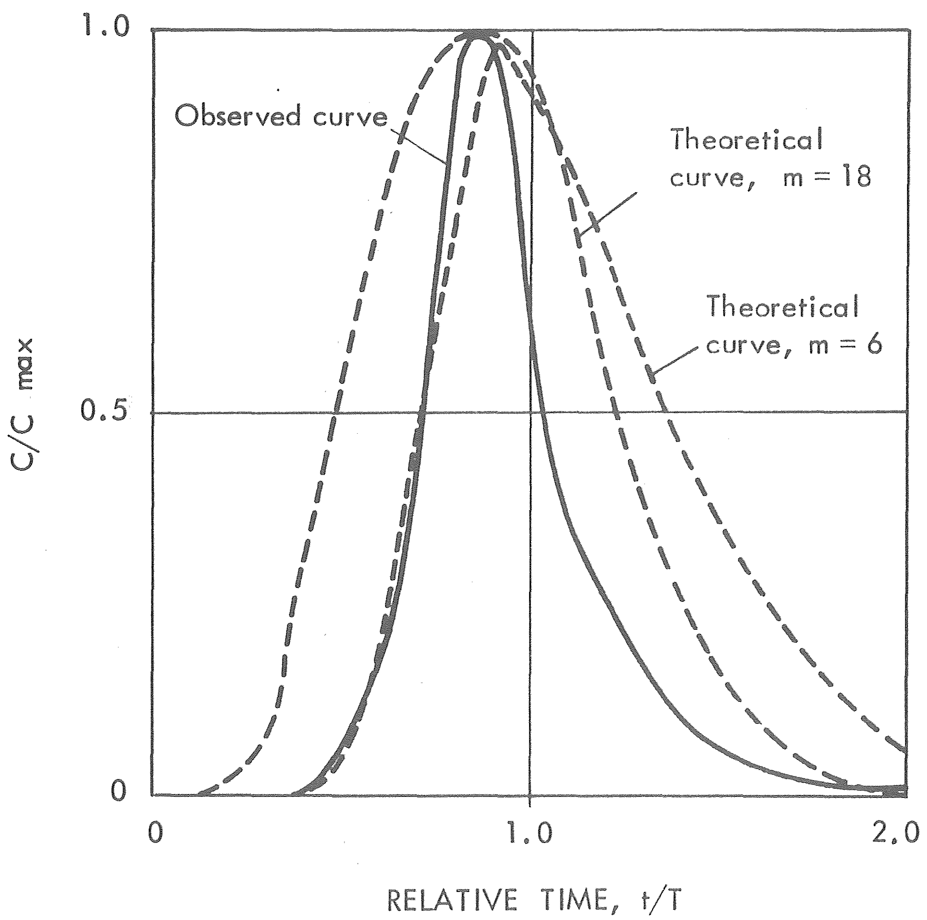


FIG. 3-8 Distribution curve in a flocculation unit with 6 tanks, each divided by 2 baffles, compared to a theoretical calculation for 6 and 18 CMF reactors. Full scale plant.

3.5.5 Direction of flow through reactors

Investigations – described in detail by Hedberg (1974) – have shown that the main direction of flow through the reactor (FIG. 3-9), does not significantly affect the result. Therefore, the simplest design, type 1, is preferred. The intention of this study was to investigate whether the detention time of the flocs in relation to the detention time of the water was of importance.

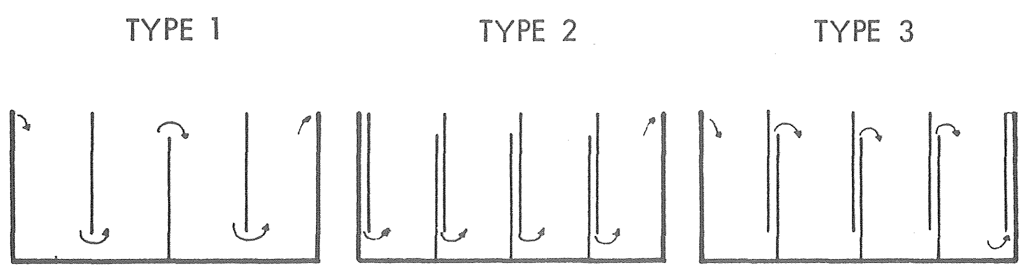


FIG. 3-9 Different principles of flow direction

3.5.6 Paddle design

The paddle design is important for the floc formation. Different paddle designs are used in practice and are also described in the literature. However, diverging opinions exist concerning the significance of paddle design and the energy distribution properties of the paddle. One predominant opinion is that the main collision of floc occurs within the small eddies formed at the edge of the paddle blades. The energy input must be adapted to the floc size and floc strength, and the energy distribution in the reactor must be as even as possible.

In the pilot plants and the full scale plant in which the studies have been performed, the stirring equipment consisted of vertical pipes. Other paddle designs were also briefly studied (FIG. 3-10).

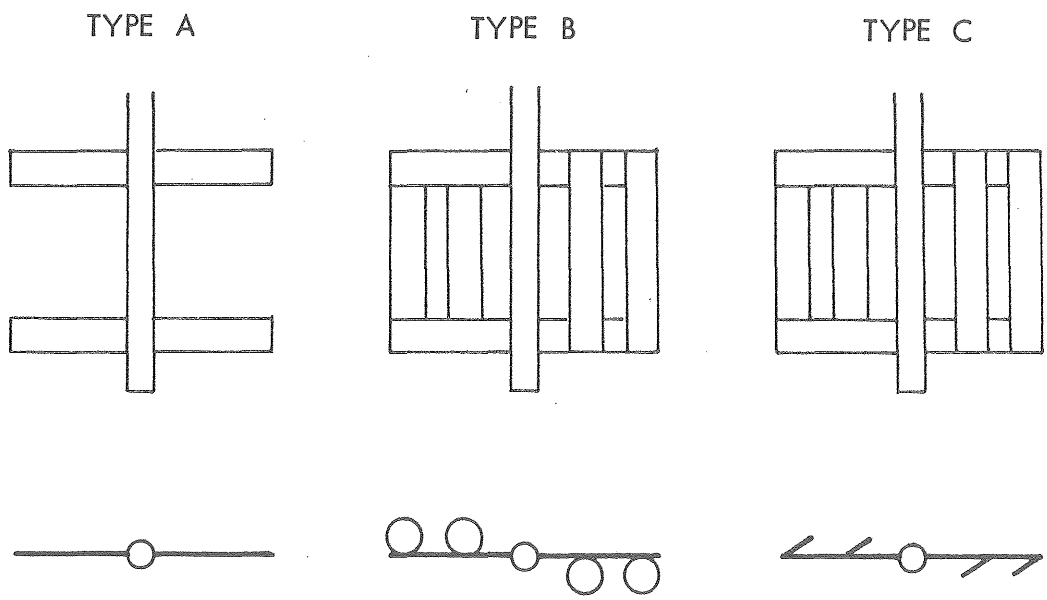


FIG. 3-10 Different paddle designs

The relationship between the paddle area and the cross-section of the flocculation tank varied for the different types from 16 % (type A) to 34 % (types B and C). The difference between types B and C is the shape of the paddle (see FIG. 3-10). The total length of the edges of the paddle L_p, which is regarded as very important in floc agglomeration, is in the three cases, type A: L_p = 3.3 m, type B: L_p = 5.7 m, and type C: L_p = 8.9 m.

For the paddles type A and B, the velocity gradients G have been calculated by Camp's formula – with an estimate of the relative velocity of the paddle v_p with respect to the fluid v_r of 0.5. In the case of paddle design type C, the calculation is more complicated. The projected paddle area is the same as for the cylindric paddle, but the paddle type C consists of plates obliquely-angled 30^o to the direction of motion. The drag coefficient C_p which depends upon paddle shape and flow conditions ought to be of the magnitude of 2.0. It is, however, possible that the rotation of water increases simultaneously, and this is of greater importance. If an increase of the drag coefficient from 1.2 to 2.0 causes a simultaneous increase in water rotation so that the value of v_r decreases from 0.5 to 0.41, the velocity gradient G will remain the same. Considering to this uncertainty the G-values of the obliquely-angled stirrer have been calculated in the same way as for the cylindric stirrer.

As a reference, the results obtained from two series in which paddles with "pipes" have been used, are presented. The tests have been performed on two occasions with some difference in water temperature; consequently, without further consideration the results are not comparable. The overflow rate at the breakup point has, therefore, been calculated for the lamella sedimentation unit used. The breakup point for the sedimentation operation has been chosen at the point where the turbidity exceeds the residual turbidity c_o by 10 %. The relative value of 1 has been chosen for the overflow rate for the tests with the paddle design of type B. The relative values of the overflow rates for the other types of paddle designs can thus be calculated. FIGURE 3-11 shows the results obtained as a function of the earlier mentioned factor K_j (Eq. 2-23). The number of tests is small, but the result indicates that the paddle design is of great importance. An increase of the factor K_j means that the performance of the sedimentation unit can be increased due to an advantageous floc size distribution. The test shows that K_j should be as large as possible.

The factor K_j gives information about the energy distribution in

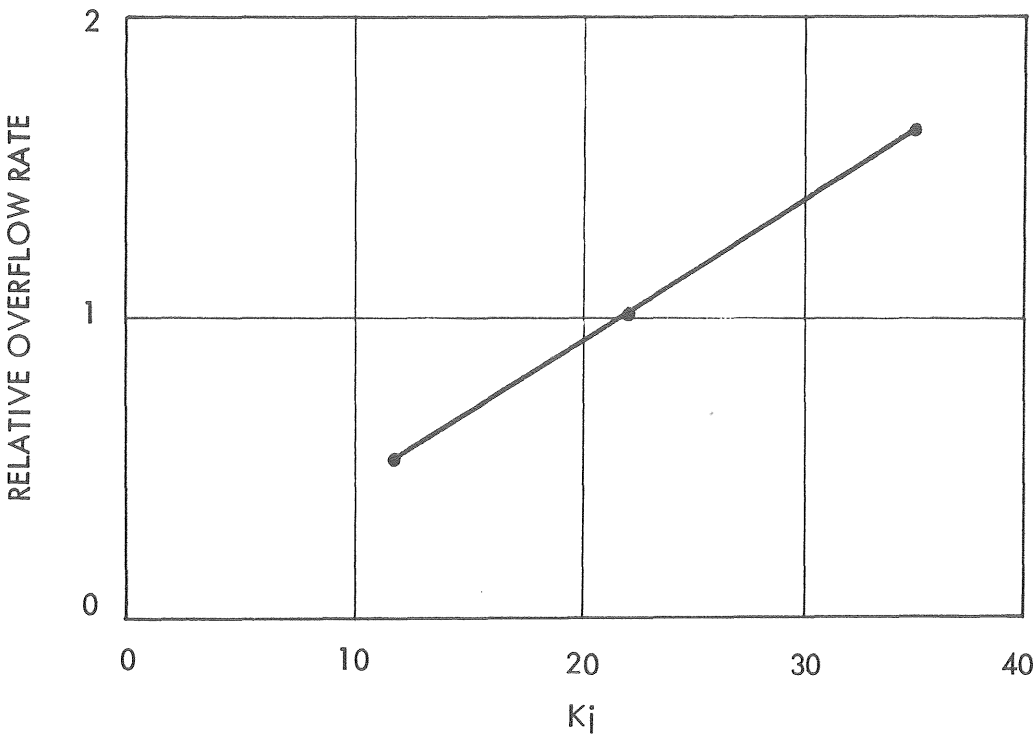


FIG. 3-11 Relative overflow rate in the sedimentation unit at the breakup point as a function of the paddle design factor K_j

the reactor. The rough picture of the flocculation obtained by calculation of the velocity gradient alone is thus complemented by the suggested paddle design factor. However, it is an empirical measure based on intuitive considerations: the higher the K_j-value, the more even the distribution of the energy input represented by the G-value. A more even energy distribution means less risk for breakup of flocs and also gives a uniform floc size distribution.

One way to increase the K_j factor is to increase the number of paddle blades. A larger paddle area, however, increases the tendency to purely rotary motion which is believed to be relatively ineffectual in generating eddy currents. A larger projected area of paddle assembly than 1/3 of the cross sectional water area is to be avoided according to Hyde and Ludwig (1944). Paddle design in the form of a coarse net might be preferable in order to simultaneously obtain a high value of the K_j factor and a small total paddle area. This should be investigated in future flocculation studies.

3.6 Principle of distribution of the total reactor volume

3.6.1 General test conditions

Normally, a certain number of reactors in series with an equal residence time are used, but a theoretical analysis has shown that a successively increased residence time in each reactor improves the performance of the flocculation unit. This hypothesis was tested in a number of experiments. A practical application of the principle with a successively increasing residence time means that the volume in the first flocculation tanks is smaller than in the last ones. However, to be able to use the existing pilot plants, and in order to easily vary the degree of the detention time distribution and obtain comparisons with conventional residence time distribution, we simulated this experiment by shunting a certain amount of water from each flocculation tank in the pilot plant as shown in FIG. 3-12. The investigations have been carried out in two pilot plants, No. 3 and No. 5.

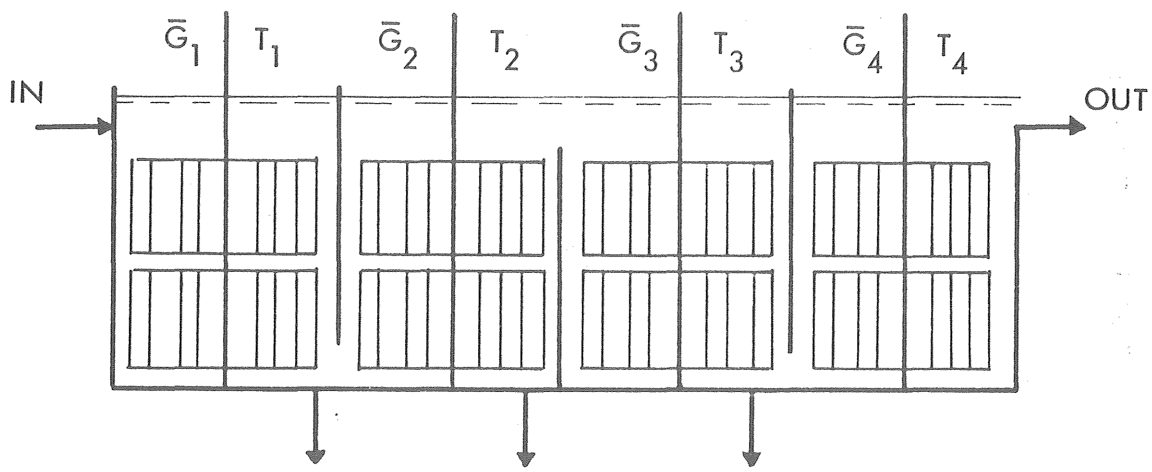


FIG. 3-12 Sketch of the principle of the test with a successively increased residence time

After each test a comparison test with a conventional time distribution was carried out.

3.6.2 Results

The results from the investigations with different time distributions in the flocculation units are summarized as follows. Altogether, about 30 tests, each lasting at least one day, have been carried out. The study was, however, of a qualitative character and more systematic investigations will be necessary in the future in order to confirm the observations made.

The research indicates that it is possible to obtain the same result measured in turbidity units after sedimentation at a shorter residence time in the flocculation unit if a successively increased flocculation time is used in the reactors compared to when an equal detention time is used. It is necessary to compensate for the short residence time in the first reactors by an increase in the velocity gradient, so that the product G_i t_i is of the same magnitude in the two systems. The results obtained within the temperature interval of 11.8-14.5^0 are exemplified in FIG. 3-13.

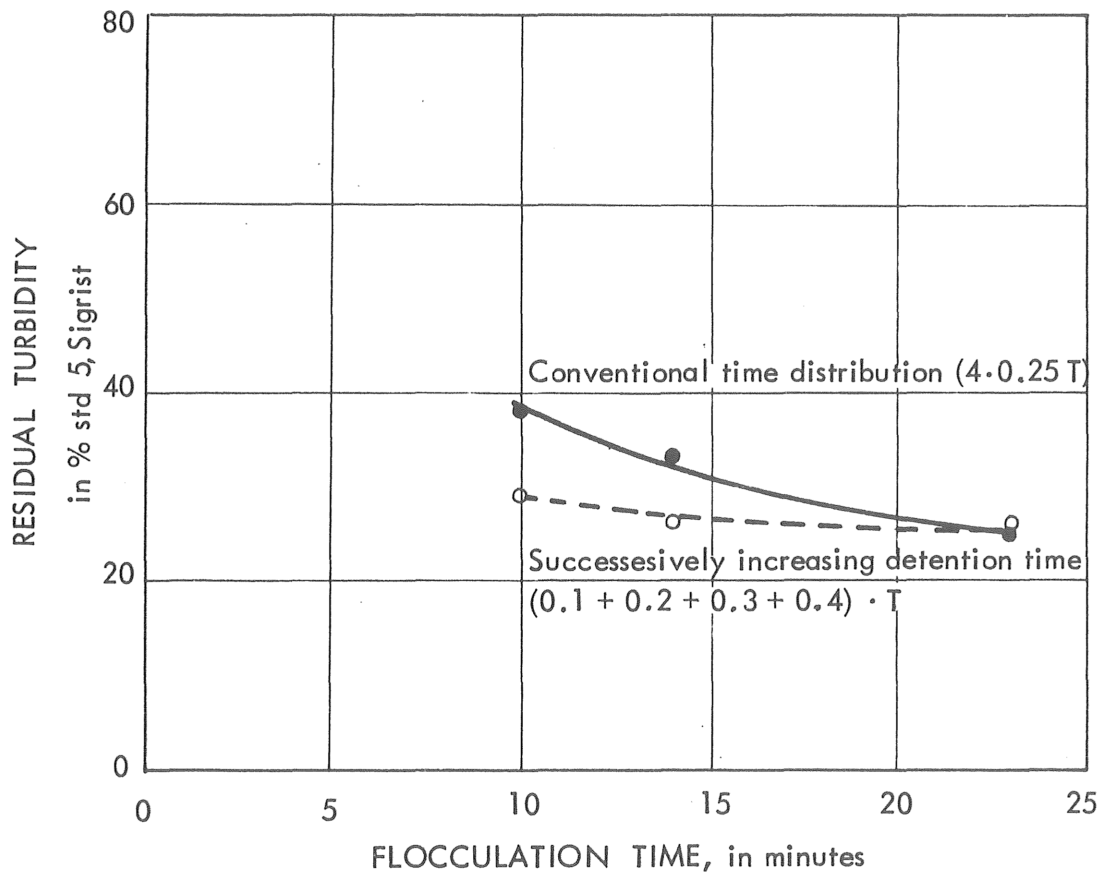


FIG. 3-13 Comparison between conventional and successively increased flocculation residence time. Temperature at 11.8-14.5^{0}C, pH-value 6.2-6.5. Overflow rate in the sedimentation unit: 0.5-0.6 m/h.

From FIG. 3-13 one can see that the relative advantages of the successively increasing flocculation time are increasing with decreasing total flocculation time. In the investigation carried out the mean residence time in the first reactor was about 10 % and in the second, third, and fourth reactors 20 %, 30 %, and 40 %, respectively, of the total flocculation residence time. Other combinations are, of course, also possible. The results obtained give ideas for many alternate designs of the flocculation unit in a water treatment plant.

3.7 Mathematical model

The experimental investigation has included many of the parameters found in the mathematical model. In this chapter the practical application of the flocculation model is presented, illustrating how the model agrees with the results obtained.

3.7.1 Particle measurements

In the experimental investigation, the flocculation performance was measured in terms of turbidity units. Turbidity measurement is, from a practical point of view, a suitable measure of the concentration of particles. From a theoretical point of view, it is more difficult to use. However, turbidity is judged to give a useful measure of the number of particles as long as the suspension characteristics studied do not vary to a great extent. The particles are assumed to have constant light absorption and reflection characteristics. It is also important that the particle size distribution is as similar as possible on every occasion of measurement. One way to ensure this is to break up the floc to a uniform size by intense stirring of the sample. Normally, the sample is taken after sedimentation and will thus contain particles with a slower settling volocity than that which corresponds to the overflow rate in the sedimentation unit. It has been shown, at least in laboratory tests, that there is a good correlation between the number of particles and the turbidity in the water (Hedberg, 1975).

3.7.2 Effect of temperature on flocculation and sedimentation

In the basic mathematical model the effect of temperature on the flocculation process has not been considered; however, a certain dependence on temperature has been observed. This means that the coefficients K_1 and K_2 in Eq. (2-35) ought to be dependent on the temperature. As the flocculation effect is measured indirectly by sedimentation in real sedimentation units, the temperature effect in the sedimentation is also included. The actual design of the sedimentation unit is important due to flocculation effects during the sedimentation. In order to consider the influence of temperature on the sedimentation operation, one can introduce a temperature dependent term K_R in the mathematical model, or include the temperature influence on the sedimentation in the coefficients K_1 and K_2 in Eq. (2-35). Both methods have been tried. The temperature dependent term K_R may be expressed as:

$$K_R = \frac{K_{R10} \cdot 60}{42 + 1.8 \cdot t} \qquad (3\text{-}1)$$

in which K_{R10} is a coefficient determined at a temperature of 10^oC

t is the temperature in ^oC

$78.5/(42+1.8 \cdot t) \cdot 10^{-6}$ is Hazēn's approximation of the variation of viscosity with temperature within the normal temperature range

3.7.3 Coagulant aids

The residual turbidity measured after a certain sedimentation time is normally decreased when the dosage of activated silica is increased. As mentioned, an increase in water temperature has a similar effect. At higher temperatures the influence of activated silica on the residual turbidity is relatively small. For optimization purposes it is necessary to be able to describe the effect of the activated silica at different temperatures, although the mechanisms affecting the floc size distributions have not been studied.

The effect of activated silica can be regulated by the coefficients K_1 and K_2 in the flocculation model but as such a method has not been sufficiently investigated at present, it seems more convenient to control the effect of activated silica by the term K_R introduced in the previous section.

The influence on the residual turbidity c_o of the activated silica addition A at different temperatures, can be expressed as: (FIG.3-5).

$$f_a = \frac{c_o\ (A=0)}{c_o\ (A>6)} = 1 + \frac{a}{1+(1+bt)\ (1+d \cdot A^2)} \qquad (3\text{-}2)$$

in which f_a is a factor expressing the influence of activated silica addition on the residual turbidity

A is the activated silica addition

$c_o(A=0)$ is the residual turbidity when A = 0

$c_o(A>6)$ is the residual turbidity when A > 6 mg/l

a, b, d, are coefficients

t is the temperature

3.7.4 Paddle design

The investigations have shown that the paddle design is of great importance for the residual turbidity as well as the floc size distribution. As suggested by Argaman and Kaufman (1970), a paddle design coefficient may be included in the coefficient K_1 in the floc growth equation (Eq. 2-24). However, the investigations carried out so far are insufficient to suggest a quantitative value for a paddle design coefficient. An approximate expression of the relationship between residual turbidity c_o and the paddle design coefficient K_j in Eq.(2-23) has been suggested, but it is considered incorrect at this stage to use such an expression.

3.7.5 Hydraulic conditions

In the previous section 3.5 it has been shown that a reactor may function as a completely mixed flow reactor or a reactor with less dispersion, depending on the velocity gradient and the rate of flow through the reactor. The residence time distribution depends on baffles or other arrangements within the reactor. Thus, in each case, the hydraulic conditions must be examined with tracer studies or at least be estimated in order to calculate the equivalent number of completely mixed reactors in series which give the same residence time distribution as the actual reactor. As reactors may function differently under different conditions, it has been considered correct to use the equivalent number of completely mixed reactors in series in the flocculation model.

3.7.6 The working mathematical model

As already shown, several factors affect the flocculation process, and since the different factors are probably not independent variables, the mathematical model can be very complicated. It has been suggested that the effect of water temperature and the dosage of activated silica can be regulated by the coefficients K_1 and K_2 or by a residual term K_R in the flocculation model. An adaption of the basic model equation (Eq. 2-34) gives the following general expression of the flocculation performance when the number of particles is expressed in turbidity units measured after sedimentation:

$$c_{o,i} = \frac{100-K_R}{100} c_{o,i-1} \left[\frac{1+K_2 \frac{\bar{G}_i^P}{c_{o,i-1}} \cdot \frac{R_i T}{F_i}}{1+K_1 \bar{G}_i \cdot \frac{R_i T}{F_i}} \right]^{F_i} + K_R \qquad (3\text{-}3)$$

in which $c_{o,i}$ is the turbidity after sedimentation. (Per cent of the initial turbidity c_{100})

- K_1 is a coefficient for floc growth
- K_2 is a coefficient for floc breakup
- $\bar{G}_i$ is the mean velocity gradient for the reactor i (mesaured by Camp´s formula)
- R_i is the relation between the residence time in reactor i and the total residence time
- T is the total residence time
- F_i is a coefficient describing the hydraulic conditions in the reactor i (F_i=1 for a CMF reactor, F=∞ for a plug flow reactor)
- K_R is a term expressing the temperature effects in the sedimentation and flocculation units
- P is the floc breakup exponent

$$K_R = K_{R10} \left(\frac{60}{42+1.8\ t}\right) \left(1 + \frac{a}{\{1+bt\}\ \{1+dA^2\}}\right)$$

in which K_{R10} is a coefficient

A is the activated silica addition

a, b, d are coefficients

3.7.7 Determination of coefficients

Due to the complexity of the flocculation model it is rather difficult to calculate the different coefficients, especially when the number of reactors is large. The calculations have been carried out on a Hewlett-Packard calculator, which has plotted the residual turbidity c_0 as a function of time. The curves have then been compared with the observed result.

As a first data set the result obtained in pilot plant No. 5 has been used. The following variables have been studied:

Temperature	2.5-14.5^{o}C
Residence time in the flocculation unit	6 - 41 minutes
Velocity gradient	5 - 450 s^{-1} (calculated with the Camp formula)
Residence time distribution	$\frac{R_1}{R_4}$: $\frac{0.08}{0.52}$ to $\frac{0.25}{0.25}$

The following parameters have been constant

Coagulant dose	40 mg/l alum
Activated silica	4-6 mg/l
Flocculation pH-value	6.2-6.4 (optimum value, see section 3.4)
Paddle design	
Number of reactors	1 to 6

As a measure of the initial particle concentration c_{100}, a value of 8 Formazin Turbidity Units (FTU) has been chosen, based on several measurements of the turbidity in the flocculation tank. 8 FTU corresponds to ≈ 160 % Std 10 Sigrist ≈ 1000 Zeiss Pulfrich units.

In the flocculation model there are several possible magnitudes of the floc breakup exponent p. Parker *et al.*, (1971) have on a theoretical basis suggested a value of 2 or 4 depending on the actual hydraulic conditions. In order to investigate the magnitude of the floc breakup exponent in this specific case, we tried several values between 1 to 4 (1, 4/3, 5/3, 2 and 4).

The results obtained in the pilot plant can be predicted when the floc breakup exponent is between 1 and 2. At higher values it was impossible to get a satisfying correlation. At values of the exponent between 1 and 2, it is possible to choose the coefficients K_1 and K_2 (and even K_R) so that a very good agreement between theory and practice is obtained (corr. coeff. 0.95). This is true in particular when all the data available are tested. But if the tests carried out with a successively increasing residence time in the flocculation unit are considered, the picture is somewhat different. The floc breakup coefficients 1 and 2 gave a relatively weak correlation between theory and practice. (corr. coeff. 0.55). A value of 4/3 gave the highest correlation (corr. coeff. 0.85) and therefore the floc breakup exponent has been fixed to 4/3. Thus, the result obtained in this investigation does not support the theoretical studies carried out by Parker *et al.*.

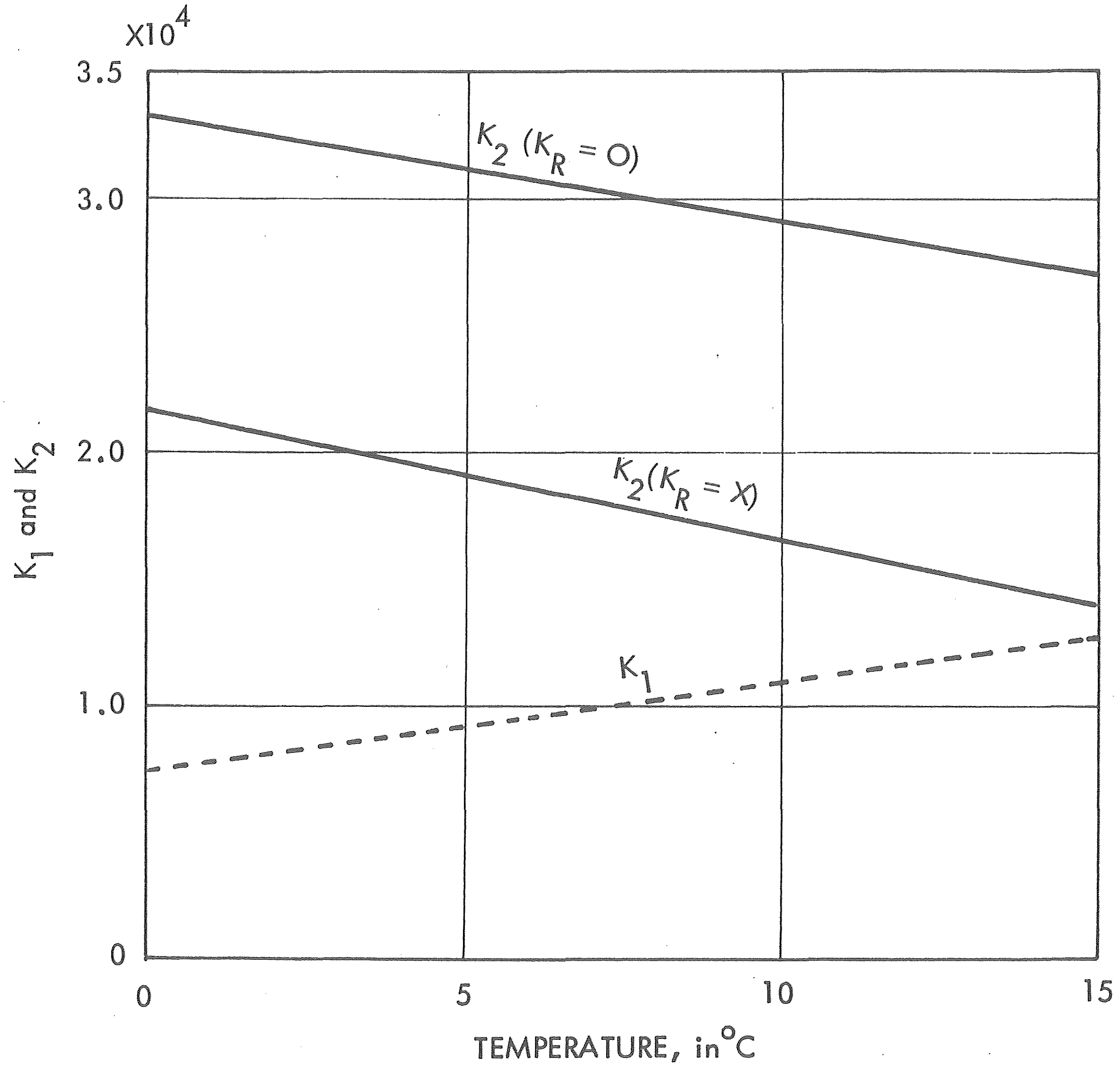


FIG. 3-14 The magnitude of the coefficients K_1 and K_2 and their dependence on temperature

3.7.7 Determination of coefficients

Due to the complexity of the flocculation model it is rather difficult to calculate the different coefficients, especially when the number of reactors is large. The calculations have been carried out on a Hewlett-Packard calculator, which has plotted the residual turbidity c_0 as a function of time. The curves have then been compared with the observed result.

As a first data set the result obtained in pilot plant No. 5 has been used. The following variables have been studied:

Temperature	2.5-14.5^{o}C
Residence time in the flocculation unit	6 - 41 minutes
Velocity gradient	5 - 450 s^{-1} (calculated with the Camp formula)
Residence time distribution	$\frac{R_1}{R_4}$: $\frac{0.08}{0.52}$ to $\frac{0.25}{0.25}$

The following parameters have been constant

Coagulant dose	40 mg/l alum
Activated silica	4-6 mg/l
Flocculation pH-value	6.2-6.4 (optimum value, see section 3.4)
Paddle design	
Number of reactors	1 to 6

As a measure of the initial particle concentration c_{100}, a value of 8 Formazin Turbidity Units (FTU) has been chosen, based on several measurements of the turbidity in the flocculation tank. 8 FTU corresponds to ≈ 160 % Std 10 Sigrist ≈ 1000 Zeiss Pulfrich units.

In the flocculation model there are several possible magnitudes of the floc breakup exponent p. Parker *et al.*, (1971) have on a theoretical basis suggested a value of 2 or 4 depending on the actual hydraulic conditions. In order to investigate the magnitude of the floc breakup exponent in this specific case, we tried several values between 1 to 4 (1, 4/3, 5/3, 2 and 4).

The results obtained in the pilot plant can be predicted when the floc breakup exponent is between 1 and 2. At higher values it was impossible to get a satisfying correlation. At values of the exponent between 1 and 2, it is possible to choose the coefficients K_1 and K_2 (and even K_R) so that a very good agreement between theory and practice is obtained (corr. coeff. 0.95). This is true in particular when all the data available are tested. But if the tests carried out with a successively increasing residence time in the flocculation unit are considered, the picture is somewhat different. The floc breakup coefficients 1 and 2 gave a relatively weak correlation between theory and practice. (corr. coeff. 0.55). A value of 4/3 gave the highest correlation (corr. coeff. 0.85) and therefore the floc breakup exponent has been fixed to 4/3. Thus, the result obtained in this investigation does not support the theoretical studies carried out by Parker *et al.*.

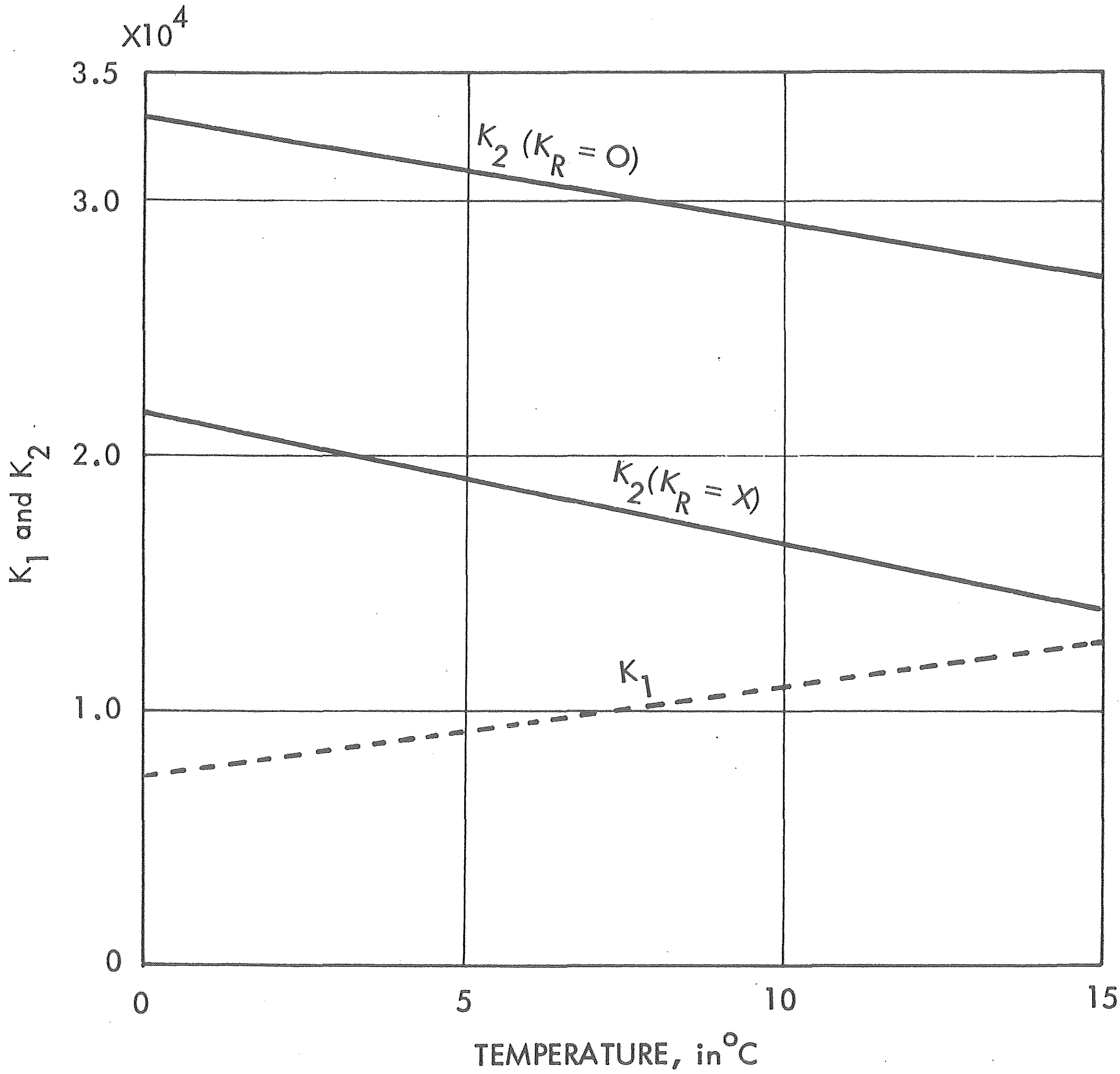


FIG. 3-14 The magnitude of the coefficients K_1 and K_2 and their dependence on temperature

It has also been found that the coefficients K_1 and K_2 in Eq.(3-3) are temperature dependent. This relationship and the magnitude of the coefficients are shown in FIG. 3-14.

As mentioned earlier the flocculation model can be used with or without a residual term K_R partly regulating the temperature dependence of the sedimentation. This residual term may be set at zero, in which case the temperature dependence of the flocculation and sedimentation must be considered when calculating the coefficients K_1 and K_2. In the case of K_R=0 the floc breakup coefficient K_2 must be somewhat higher than when the residual term is used. In both cases the floc growth coefficient K_1 is the same. As both equations are approximations, the most convenient one may be used until further investigations give more detailed information.

3.7.8 Correlation between theoretical calculations and test results

The agreement between theory and practical observations in a pilot plant is shown in FIG. 3-15.

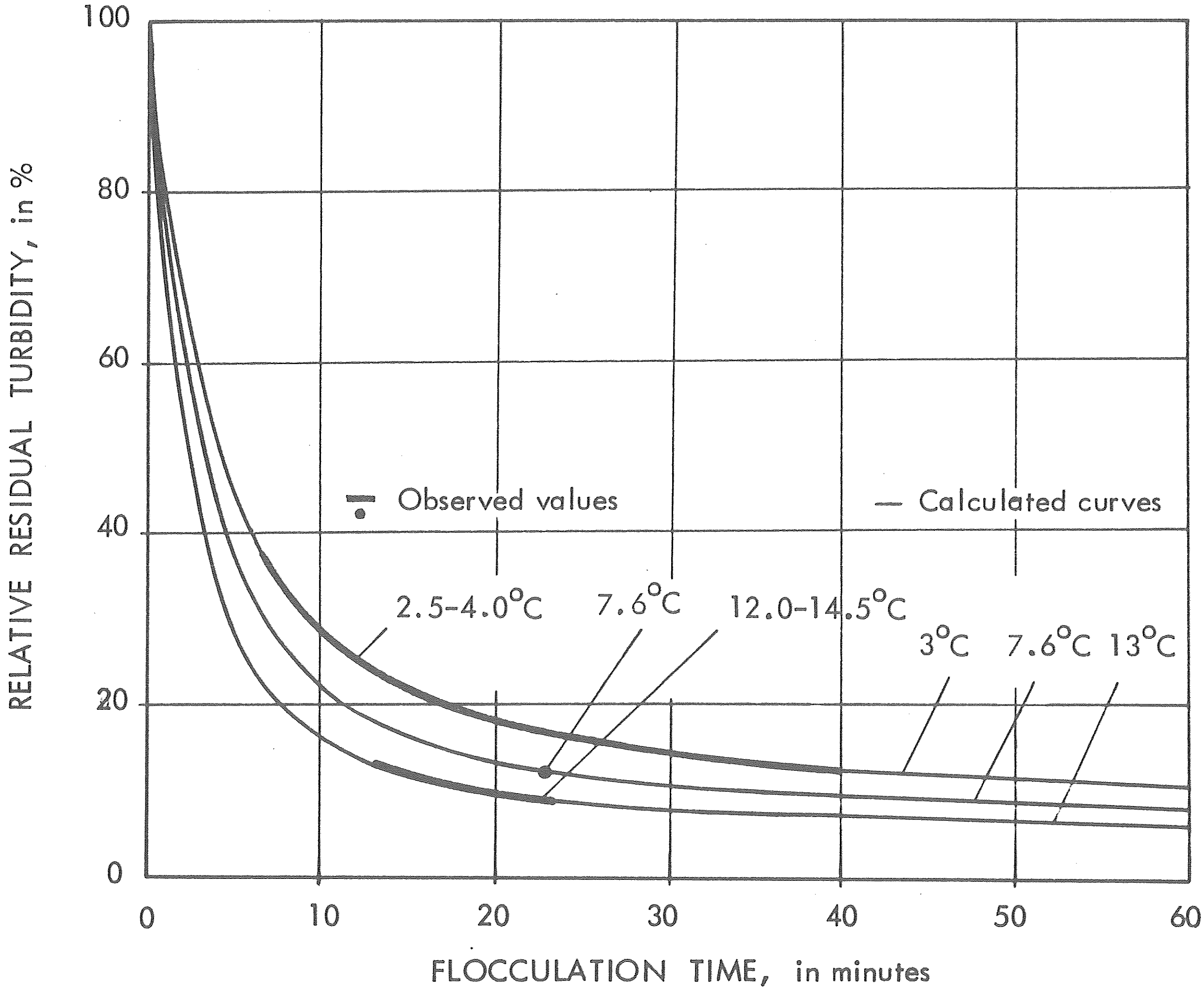


FIG. 3-15 Theoretical calculations of the residual turbidity at 3.0, 7.6 and 13.0°C (fine lines) and some observations in a pilot plant (heavy lines). (K_R=0)

All the results obtained have been compared with theoretical calculations, and the correlation is shown in FIG. 3-16.

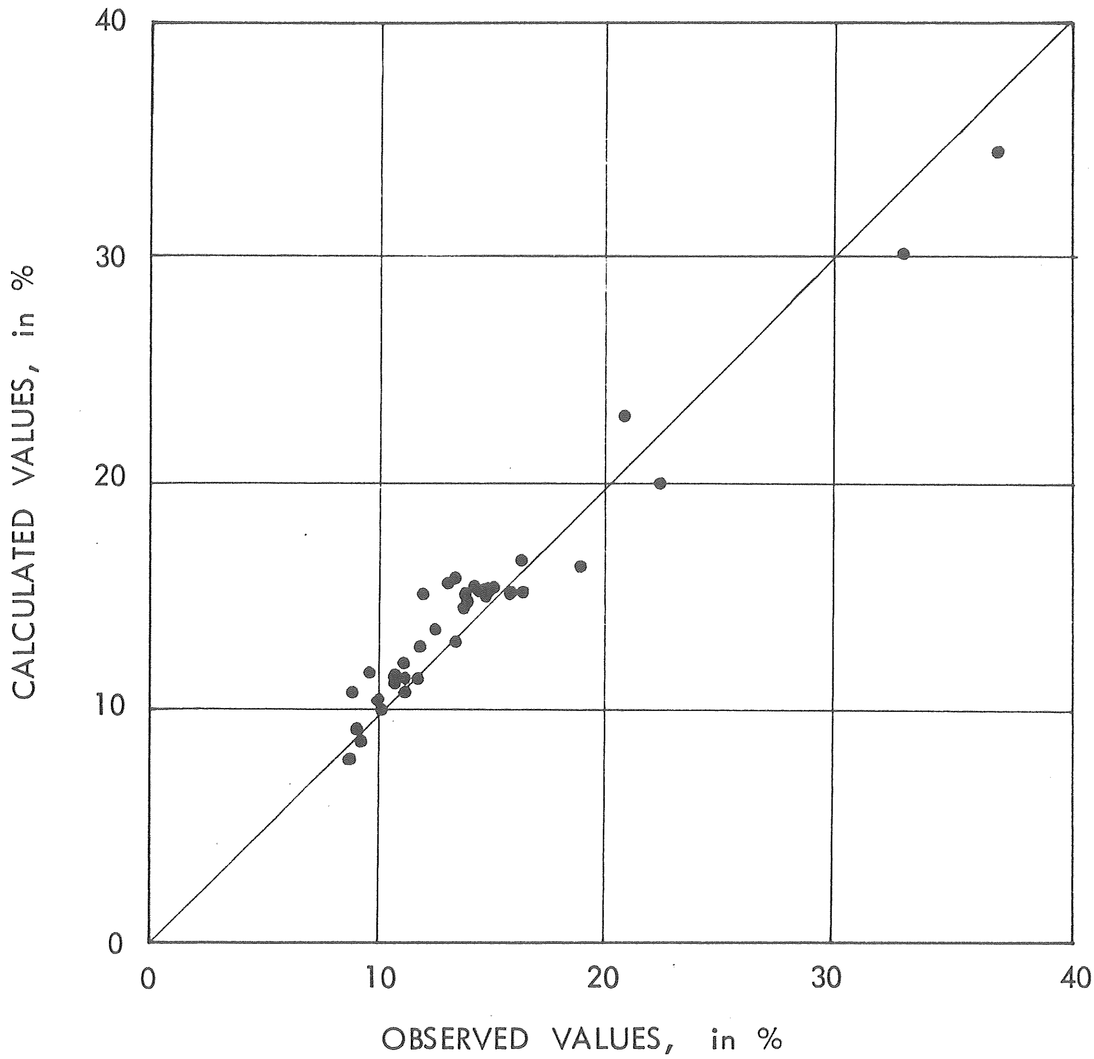


FIG. 3-16 Correlation between observed and calculated values. (Number of tests 41, r = 0.97)

As can be seen from FIG. 3-15 and FIG. 3-16, it is possible to use the flocculation model to predict the results, at least within the limits mentioned.

In addition to the investigations carried out in the relatively small pilot plant (total volume 1 m^3), a large number of tests have been performed in a bigger pilot plant (total volume 14.4 m^3, pilot plant No. 3) in which a varying number of reactors in series have been used. It has been impossible to carry out the investigations in full scale with the same variation of the system variables as in the pilot plants. However, two studies have been performed at the Water Treatment Plant at Lackarebäck. The flocculation systems examined were quite different; one system was the same as the conventional system at the plant with six reactors in series; the other system consisted of only one reactor. The water treatment plant is described in Appendix 1.

In the flocculation system with several reactors in series it was not possible to control the flocculation at different residence times in any other way than by sedimentation analysis in the different reactors. The depth of sedimentation in the special sedimentation column during this analysis was 0.3 m, and the residual turbidity was determined after 30 minutes of sedimentation time. The sedimentation depth is of great importance which will be discussed l-ter. The sedimentation depth had to be considered as the turbidity of the water in the settling column was compared with the turbidity in the effluent from the conventional sedimentation unit - the Lovö basin. The turbidity of the effluent from the Lovö basin was of the same magnitude to that from the lamella sedimentation unit used in the pilot plant experiments. The results obtained in the pilot plant and in the full scale operation have been compared with the calculated values according to the flocculation model. The result is presented in FIG. 3-17 and FIG. 3-18, and it shows a satisfactory correlation between theoretical calculated values and experimental results, even for full scale operation.

The result of the full scale test is, however, somewhat uncertain due to the testing procedure, with the exception of the result from the whole flocculation system, which was obtained after sedimentation in a real sedimentation basin.

As the flocculation and sedimentation operations proved to be affected by temperature, the combined effect can be calculated by Eq. (3-3) and compared with the observed values. One example is shown in FIG. 3-19.

It can be concluded that, although the flocculation model suggested can be further improved, it can very well be used to predict the flocculation performance within the limits mentioned in this thesis.

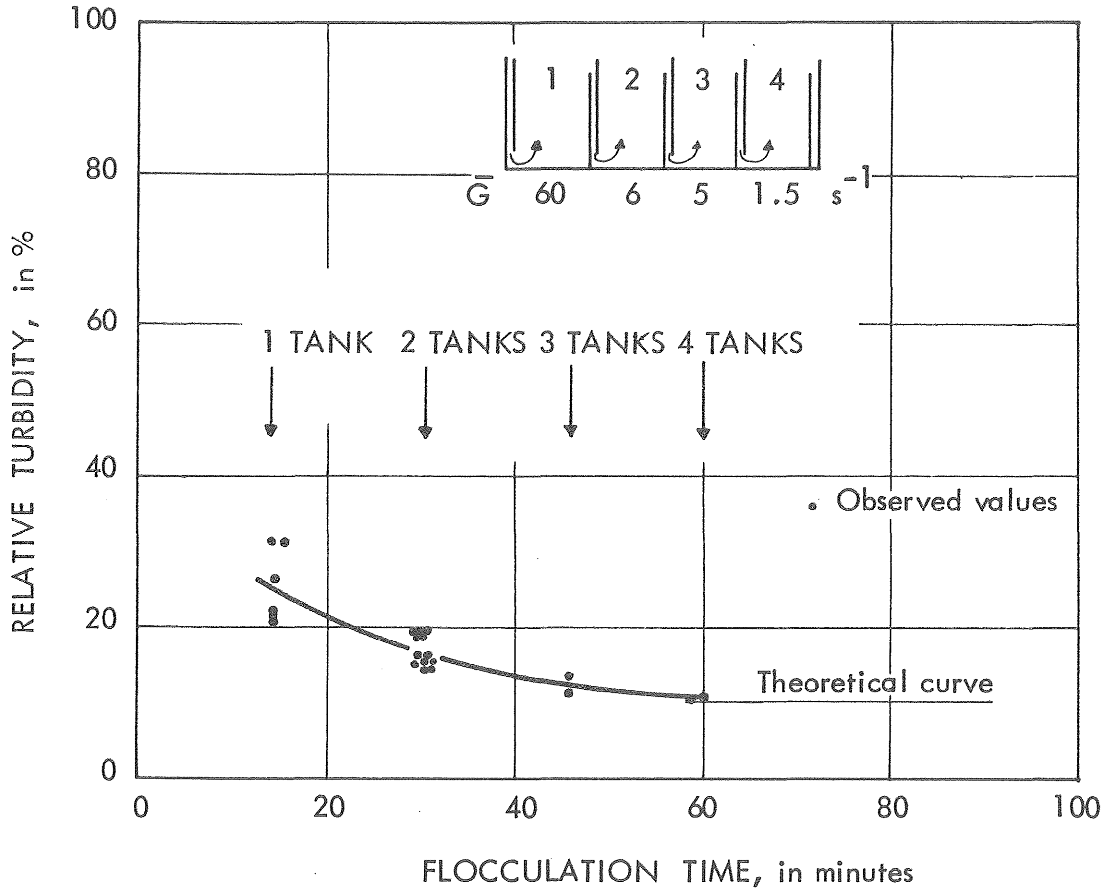


FIG. 3-17 Comparison between observed and calculated values. Pilot plant No. 1. Temperature 10°C. Activated silica 8 mg/l.

(K_R= 0. K_1= $1.1 \cdot 10^{-4}$, K_2=$2.9 \cdot 10^{-4}$)

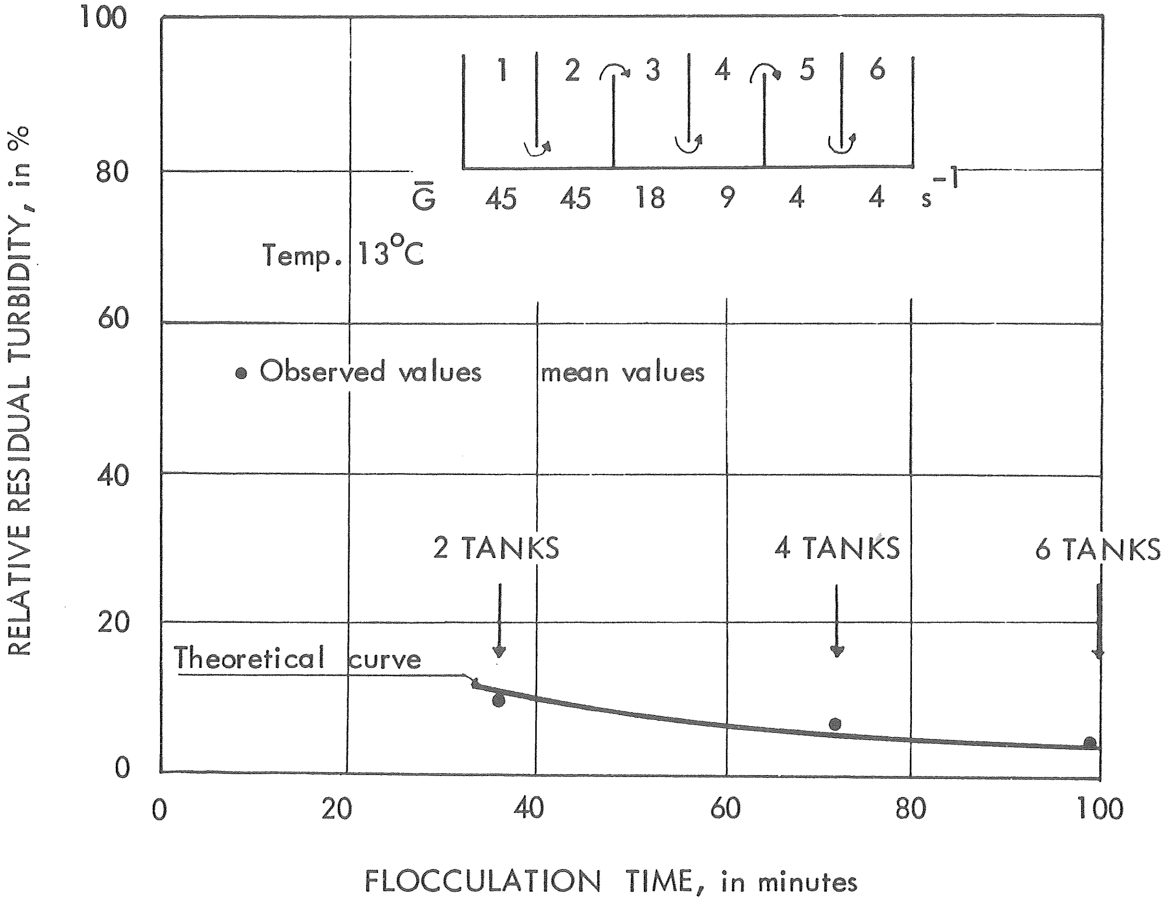


FIG. 3-18 Comparison between observed and calculated values. Mean values (total number of analyses 60). Activated silica 8 mg/l.
Full scale operation. Lackarebäck Water Treatment Plant. Temperature 13°C.
(K_R = 0. $K_1 = 1.2 \cdot 10^{-4}$ $K_2 = 2.8 \cdot 10^{-4}$)

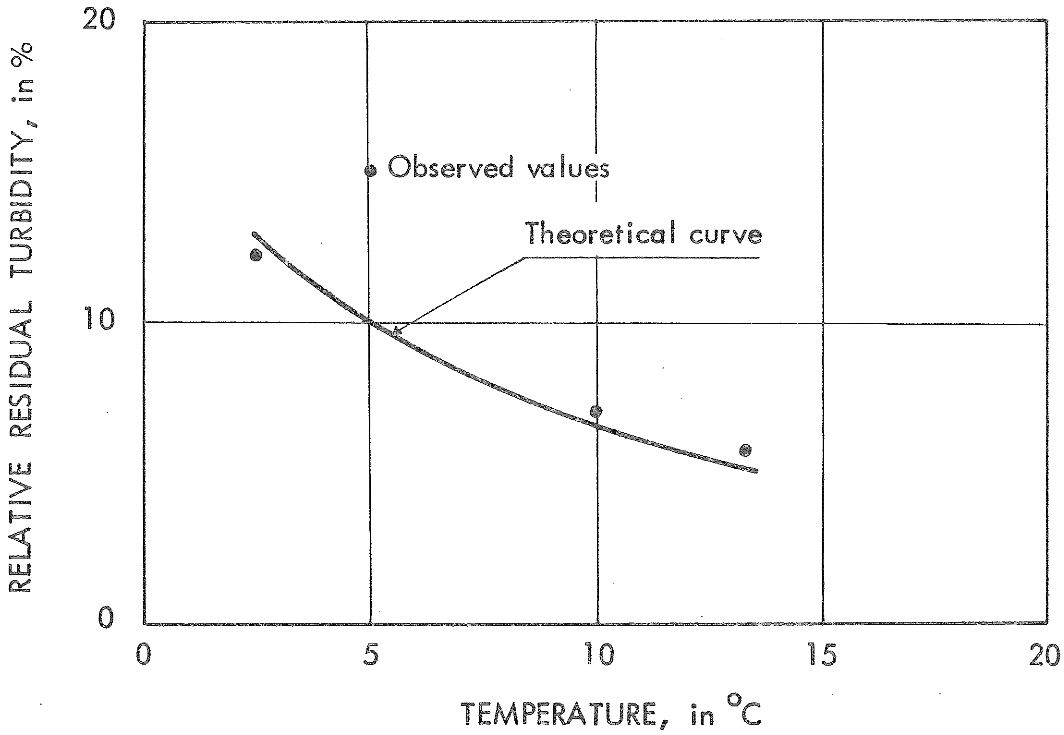


FIG. 3-19 Comparison between observed and calculated values of the turbidity in effluent from a sedimentation basin, as a function of temperature. Full scale operation

3.8 Discussion of the flocculation model

In developing and using a theoretical model, one is forced to make some approximations not only with respect to simplicity and usefulness but also because of lack of knowledge of the different factors affecting the operation. In the following pages an estimation of the effect of the different approximations is summarized.

3.8.1 Temperature

In the flocculation model the temperature influence has been considered for both the flocculation and the sedimentation operations. Two methods have been suggested, and the result, at least within the temperature interval of 4 to 16^oC, is very close to measured values.

At lower temperatures than 4^oC a rapid increase in the turbidity of the effluent from the sedimentation unit has been observed, especially when activated silica is not added. But as activated silica normally is used at such low temperatures, it is possible that the model might be used at all temperatures normal for surface waters. The influence of temperature on the flocculation is expressed in the coefficients K_1 (floc growth coefficient) and K_2 (floc breakup coefficient). When the temperature decreases, K_1 decreases while K_2 increases. Thus, an equilibrium of the energy input at which the floc growth is equal to the floc breakup may be derived as follows:

$$c_{o,i} = c_{o,i-1} \quad \text{which gives}$$

$$G_i \text{ equilibrium} = \left(\frac{K_1}{K_2}\right)^3 \cdot c_i^3$$

This means, that when $c_{o,i}$ is assumed to be constant, the relationship between those G-values at which the floc growth is zero at for example temperatures of 3^oC and 13^oC according to FIG. 3-15, can be expressed as:

$$\frac{G^{\text{equilibrium}}_{(13^oC)}}{G^{\text{equilibrium}}_{(3^oC)}} = \left(\frac{0.42}{0.26}\right)^3 = 4.2$$

Obviously, it is theoretically possible to increase the energy input at the higher temperature, which would result in a more marked difference in the flocculation performance at high and low temperatures. Thus, it is necessary to focus the research on flocculation methods which reduce the influence of low temperatures. An increase of the particle concentration in the flocculation unit has been observed to have this effect. This fact has also been reported by others (Tesarik, 1967).

3.8.2 Coagulants and coagulant aids

The flocculation model is valid for optimum flocculation pH-values and particle concentrations measured in terms of turbidity. The equation does not take into account any change in the flocculation pH-value. The coefficients for floc growth and floc breakup are determined for a specific dosage of a given coagulant (aluminium sulphate 40 mg/l), and for the use of a coagulant aid (activated silica 4-6 mg/l). Other dosages of activated silica are considered in the model by the residual term K_R. This is an approximation awaiting further research, but as activated silica normally is used, especially at low temperatures, the approximation is considered acceptable. It must be pointed out that the preparation and the addition of activated silica are of fundamental importance for the floc characteristics.

3.8.3 Hydraulic conditions

The hydraulic conditions in a reactor change with the rate of flow, energy input, and compartmentalization of a tank volume. Normally, a reactor is not completely mixed, and the flocculation model thus gives results with some safety. The deviation from a completely mixed reactor is taken into account in the flocculation model and this should be considered at least when a reactor is divided into two or several volumes by baffles. However, the influence of compartmentalization is very small if the number of tanks or compartments is higher than 4.

3.8.4 Paddle design

The paddle design and the energy distribution affect the flocculation performance. Some studies have been performed and some qualitative remarks have been made, but it is still difficult to consider these in the present model.

3.8.5 Velocity gradient

The velocity gradient G is calculated by the Camp formula but, as this calculation is based upon the relative velocities between the paddle and the rotation of water which can only be estimated, there is some degree of inaccuracy.

3.9 Safety factor

In practice it is often necessary to have a certain degree of safety. The safety factor should be chosen with consideration given to the variations in the unit operation in question and to the effect of this unit on the consecutive unit operations. As the flocculation operation is of fundamental importance for the separation operations, it is especially important to have some safety margin in this process. It has been observed that some disturbances caused by variations of pH-value and chemical dosages can be reduced at longer residence times. Thus, the theoretical residence time can be corrected by a relationship giving a larger safety factor at a shorter residence time. A safety factor of unity is suggested at a residence time of 1 h and a safety factor of 1.5 at a residence time of 10 minutes. This relationship may be written:

$$T_{practice} = T_{theory} \cdot 10^{\frac{1.7}{T_{theory}}} \qquad (3\text{-}4)$$

in which $T_{practice}$ is the residence time in flocculation (minutes)

T_{theory} is the theoretical residence time (minutes)

3.10 Settling characteristics of particles

There is no standardized method for testing the settling properties of floc particles. A floc suspension is affected to a certain extent by the method used; therefore, it is extremely important to adapt the method chosen to the separation unit in question, or, to be able to describe how the settling properties are changed in the separation unit in relation to the result obtained in an analysis carried out in a sedimentation column. The sedimentation depth is of great importance for the testing of the settling properties of floc particles in a settling column. In the investigations the depth of the sedimentation has been 0.3 m in most of the tests but 0.12, 0.15, 0.95 and 1.0 have also been used. The results obtained at the different depths are shown in FIG. 3-20. Definitions according to FIG. 2-6 have been used.

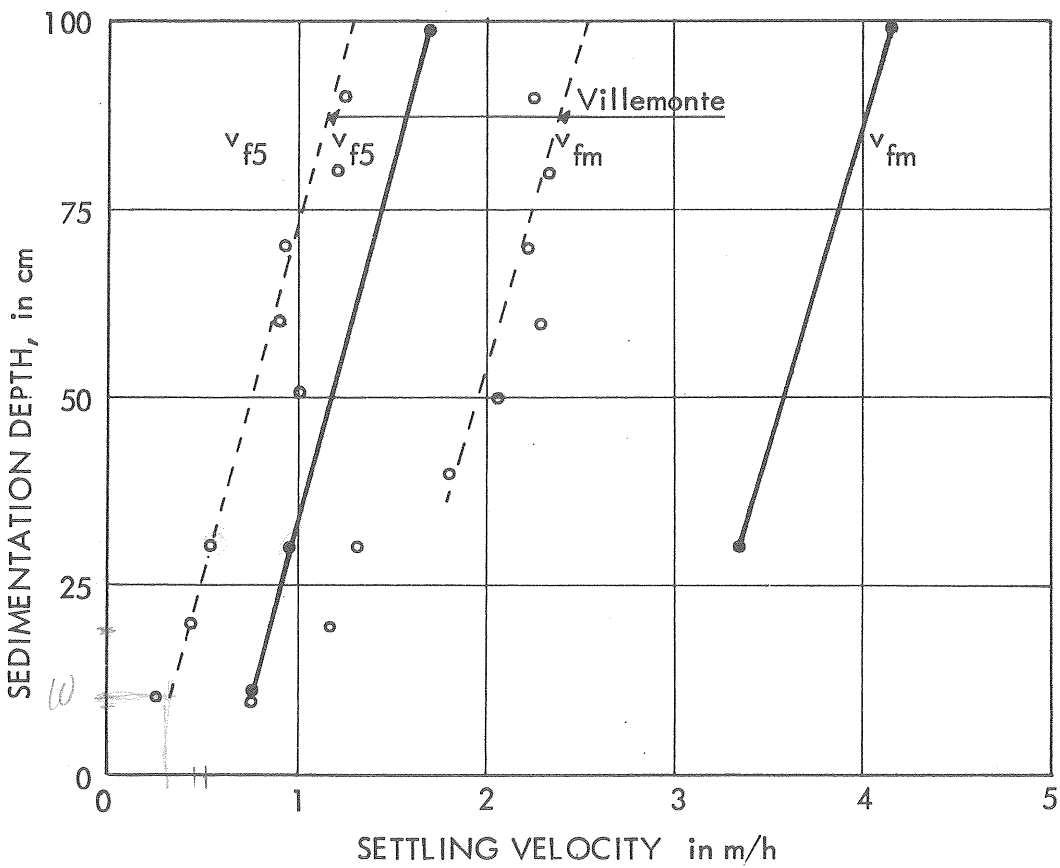


FIG. 3-20 Effect of the sedimentation depth on the settling velocity. Mean values (turbidity: $0.1 \cdot c_{100}$). (Comparison with Villemonte *et al.*, 1966).

In FIG. 3-20 mean values of the settling velocity of the flocs are shown, but as only a few sedimentation depths have been examined, a comparison with results obtained by Villemonte *et al.*, (1966) is included. The data represent the settling characteristics of aluminium hydroxide flocs. The dependence of the settling analysis depth on the result is quite similar in the two cases.

In section 2.9 the hypothesis was suggested that the settling velocity v_f and the standard deviation in the settling-velocity curve, which has an approximately Gaussian distribution, might be related to the residual turbidity. Such a relationship would be of great importance for the development of a model of the total performance in a flocculation/sedimentation system, which is necessary for optimization purposes.

Therefore the residual turbidity has been plotted as a function of the settling velocity of the floc. FIGURE 3-21 represents the results obtained at a sedimentation depth of 0.3 m.

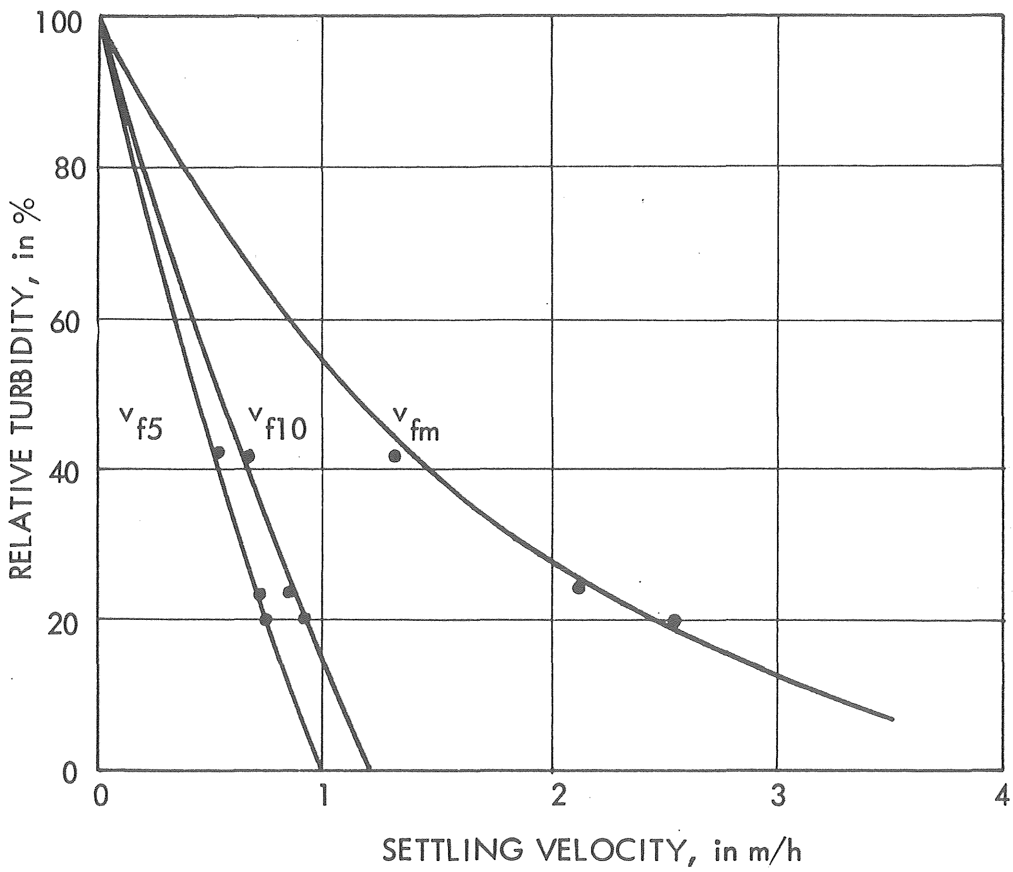


FIG. 3-21 Relationship between residual turbidity and settling velocity. Mean values (Number of analyses 40). (Pilot plant No. 7)

Sedimentation analyses have been carried out under several conditions, and the results are shown in Appendix 4. For each system it is possible to find a mathematical expression of the relationship between the residual turbidity and the settling properties of the flocs. In this investigation the following expression was found to well describe the result

$$c_o = \frac{100 + a}{1+bv_f^{1.25}} - a \qquad (3\text{-}5)$$

in which c_o is the relative residual turbidity in per cent of c_{100}

v_f is the settling velocity of the floc

a,b are coefficients depending on which settling velocity (v_{fo}, v_{f5} or v_{fm}) is considered and the sedimentation depth used.

The relationship between theoretical calculations and test results for the mean velocity v_{fm} according to Eq. (3-5) is presented in FIG. 3-22. Original data from Rosên (1967) have been used.

Due to practical difficulties in determining settling characteristics, the deviation in the results is always large. Because of particle size variation, the same is true of the turbidity measurements. The relationship illustrated is valid for residence times in the flocculation unit of 15 to 45 minutes. The correlation coefficient was r = 0.75. By considering a narrower interval of the residence times, for example results obtained at a residence time of about 40 minutes, a better correlation is achieved (r=0.95). The total number of analyses for this particular sedimentation depth was 64.

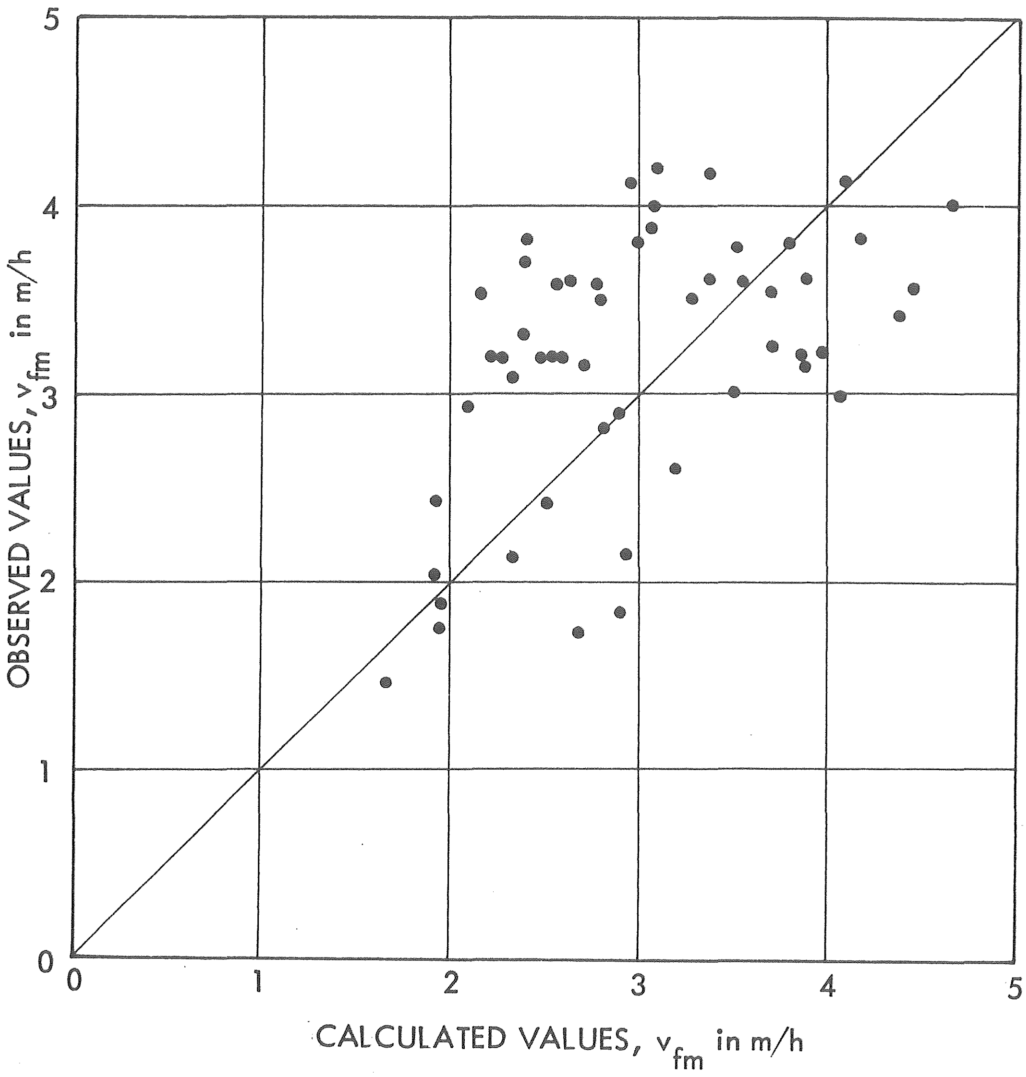


FIG. 3-22 Correlation between observed and calculated values of the mean settling velocity of the flocs. (Pilot plant No. 1. Sedimentation depth 1.0 m) a = 20, b = 0.5.

The relationship between theoretical and measured values for the settling velocity v_{fo} according to Eq. (3-5) is presented in TABLE 3-1. Data from tests in both pilot plant and full scale operation are included. As the sedimentation operation has not yet been discussed, it is necessary to calculate a relative value of the settling velocity v'_{fo} with the guidance of the value of the settling velocity of the flocs which has been calculated on the basis of Hazen's theory of the overflow rate. This value is called the observed value. Thus, it is the observed and the calculated relative values which are going to be compared with the predicted value according to Eq. (3-5). The table shows the change in the residual turbidity due to change in the flocculation residence time and change in the water temperature.

TABLE 3-1 Correlation between observed and calculated settling velocity (a=80, b=0.5 in Eq. (3-5))

Turbidity, c_0 Absolute ZP-units	Relative %	Settling velocity Observed v_{fo} m/h	Relative v'_{fo} m/h	Predicted Eq. (3-5) v''_{fo} m/h	Ratio v''_{fo}/v'_{fo}
Change in the residual turbidity due to change in residence time during flocculation. (Pilot plant No. 3)					
40	5.6	0.40	1.88	(1.88)	(1)
58	8.3	0.37	1.74	1.79	1.02
75	10.7	0.34	1.60	1.71	1.06
Change in the residual turbidity due to change in water temperature					
Lamella sed. (Pilot plant 7)					
125	16.5	1.07	1.55	(1.55)	(1)
210	28.0	0.80	1.16	1.25	1.07
Lovö-basin (Pilot plant 6)					
75	6.3	1.75	1.85	(1.85)	(1)
150	12.5	1.50	1.58	1.66	1.05

From the table it can be seen that the correlation between the relative and the predicted value is good, and it may be concluded that the relationship according to Eq. (3-5) can be used for calculation of the settling velocity of the flocs under several different conditions.

In order to completely describe the Gaussian distribution of the settling velocity, one also needs information on the standard deviation σ. Earlier, the hypothesis that also σ may be related to the residual turbidity has been mentioned. Before this can be shown, the calculation of the deviation must be discussed. Rosén (1967) has developed a simplified method for calculation of the standard deviation:

$$\sigma = 2\ (v_{f40} - v_{f60}) \tag{3-6}$$

in which v_{f40}, v_{f60} are the settling velocities corresponding to the relative turbidity of 40 % and 60 %, respectively (see FIG. 2-6)

If this equation is used for suspensions with different mean settling velocities, a positive correlation ($r = 0.94$, number of analyses = 64, sedimentation depth = 1.0 m) between the residual turbidity and the settling velocity v_{fm} can be obtained, which is not in agreement with the results obtained from tests in sedimentation units. This false result is due to the calculation method. If instead the time-turbidity curve is used as a basis for calculation of the standard deviation, a negative correlation between the residual turbidity and the settling velocity is obtained. This argumentation is illustrated in FIG. 3-23.

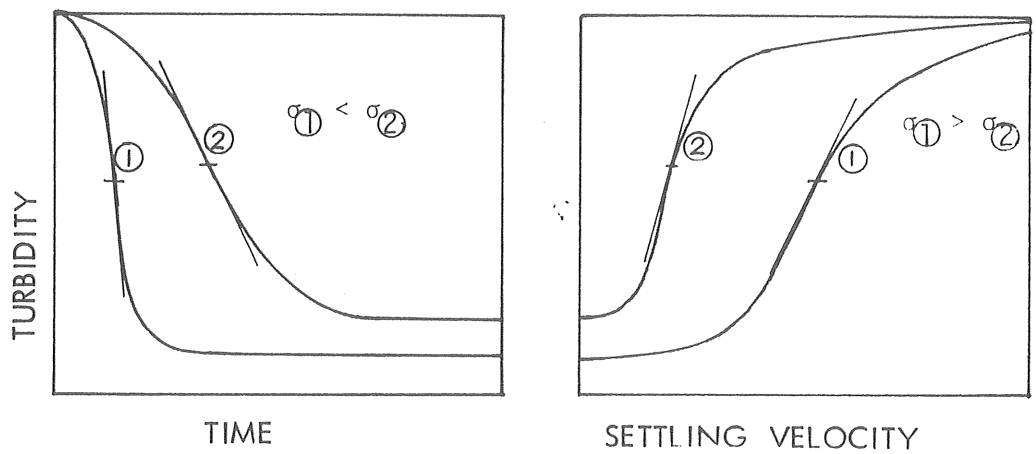


FIG. 3-23 Picture of the principle for calculation of the standard deviation

In the determination of the standard deviation of the settling velocity distribution curve, different values were obtained depending on the sedimentation depth used. For a floc suspension with a mean settling velocity of 3.0 m/h, a standard deviation of 0.8 m/h was obtained at a sedimentation depth of 0.3 m and a deviation of 0.4 m/h at a 1.0 m sedimentation depth. This shows that the settling characteristics of the flocs change during settling. It also indicates that different sedimentation units may affect the suspension. It is important to be able to determine an absolute value of the standard deviation σ from sedimentation analysis for each suspension. In order to determine the magnitude of the standard deviation, we have used all the analyses carried out, and the standard deviation has been calculated at the mean value of the time t_{50} (FIG. 2-6). The relationship between the standard deviation and the mean settling velocity of the flocs is illustrated in FIG. 3-24.

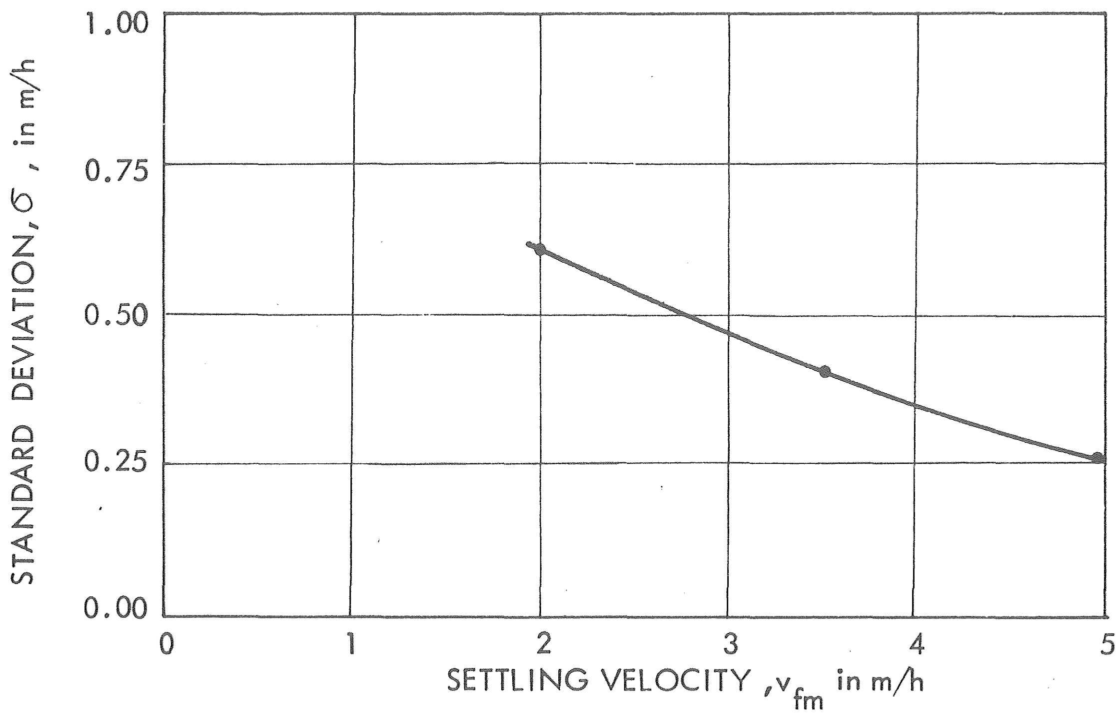


FIG. 3-24 Relationship between standard deviation σ and mean settling velocity v_{fm}. Mean values (correlation coefficient r=0.95, sedimentation depth 1.0. Pilot plant No. 1)

From the settling analysis carried out it has been observed that the deviation from the Gaussian distribution increases when the mean settling velocity increases.

Even though an almost linear relationship has been found between the residual turbidity and the mean settling velocity of the flocs, it has been regarded more correct to give a mathematical expression as follows:

$$\sigma = \frac{1}{1+k_\sigma \cdot v_{fm}} \qquad (3\text{-}7)$$

in which k_σ is a coefficient dependent on the characteristics of the suspension and the sedimentation depth

Equation (3-7) and the equation describing the relationship between turbidity and mean settling velocity, Eq. (3-5), give the relationship between the residual turbidity c_0 and the standard deviation σ as follows:

$$c_0 = \frac{100 + a}{1+b\left(\frac{1-\sigma}{k_\sigma \cdot \sigma}\right)^{1.25}} - a \qquad (3\text{-}8)$$

in which a, b, and k_σ are coefficients.

This expression can be simplified to a linear expression as follows:

$$c_0 = k_g \cdot \sigma + l_g \qquad (3\text{-}9)$$

in which k_g and l_g are coefficients dependent on the characteristics of the suspension and the sedimentation depth.

The relationship between residual turbidity and standard deviation is illustrated in FIG. 3-25.

If the original time-turbidity curve is used as a basis for calculation of the standard deviation, it can be shown that an almost linear relationship between the residual turbidity and the standard deviation exists. (r=0.82 number of analyses = 40).

Several factors affect the sedimentation analysis and the suggested relationship may not be generally used. As a conclusion it can be stated that there is a relationship between the residual turbidity and the settling properties measured by the mean settling velocity and the standard deviation in the assumed Gaussian distribution of the settling velocity.

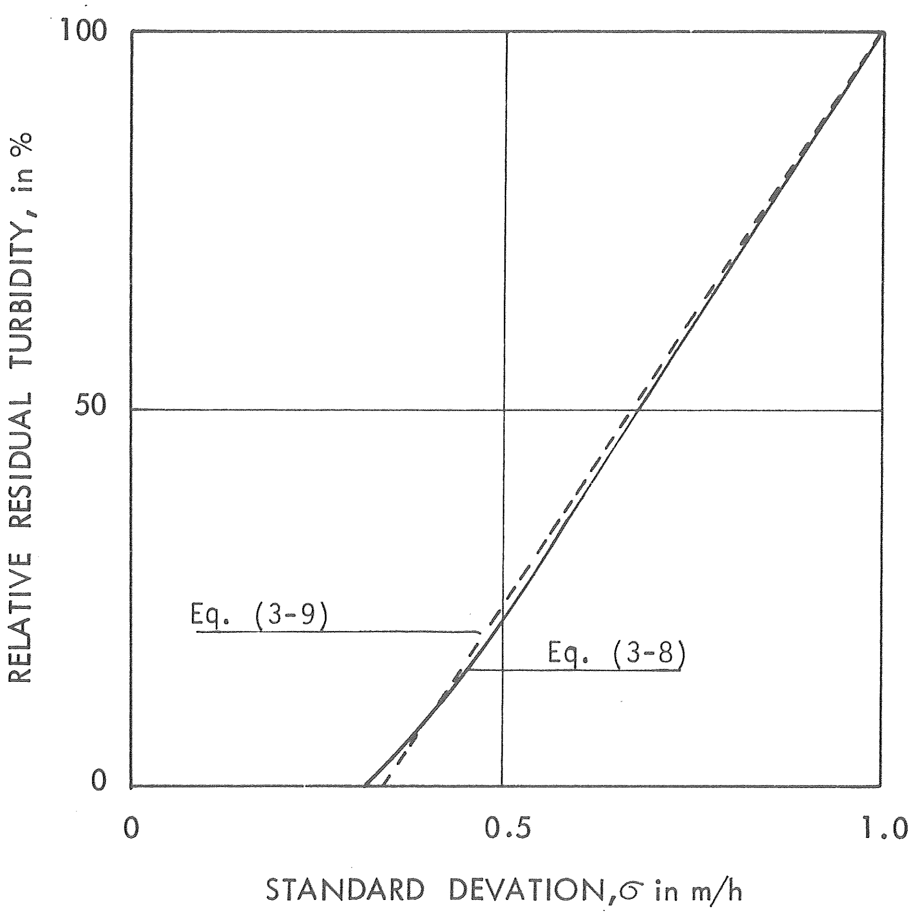


FIG. 3-25 Relationship between residual turbidity and standard deviation. (Sedimentation depth 1.0 m, k_g = 154, l_g = 54, Pilot plant No. 1).

Thus, it is possible to combine Eqs.(2-38), (3-5), and (3-9) in order to describe the settling velocity frequency curve:

$$f(v_f) = \frac{1}{\sqrt{2\pi}} \exp \left[- \frac{\left[v_f - \frac{1}{b} \left(\frac{100 - c_o}{a + c_o} \right)^{4/5} \right]^2}{2 \left[\frac{c_o - l_g}{k_g} \right]^2} \right] \tag{3-10}$$

in which c_o is the relative residual turbidity calculated by the flocculation equation, Eq.(3-3)

- a and b are coefficients of the mean settling velocity
- k_g and l_g are coefficients of the standard deviation
- v_f is the settling velocity of the floc

The coefficients are dependent on the characteristics of the suspension and the sedimentation analysis method used.

3.11 Discussion of the model describing the settling characteristics of the flocs

The developed equation is based on earlier assumptions by Rosén (1967) who found that the settling velocity of the flocs has approximately a Gaussian distribution. In general, turbidity measurements are used in order to determine the settling characteristics, but as the value obtained is a measure of the product of number and size of the particles, it is necessary to achieve in some way, e.g. by vigorous mixing, a homogenous particle size. However, it is believed difficult to obtain a similar breakage of the flocs originating from different flocculation conditions. In certain sedimentation analyses it has been observed that the settling velocity distribution function deviates from the Gaussian distribution, which is probably due to the experimental method used. An analysis of particle size is necessary in order to draw correct conclusions about the nature of the settling velocity distribution of the flocs.

The sedimentation depth plays an important part in the result, and if the suspension characteristics are unknown, several depths may be used.

The suggested linkage between the residual turbidity and the mean settling velocity of the flocs as well as between the residual turbidity and the standard deviation of the settling velocity of the flocs seems to be quite useful for optimization purposes as several factors such as temperature, energy input, and residence time automatically affect the settling velocity distribution of the floc.

4 SEDIMENTATION, BACKGROUND

4.1 General

The development of the sedimentation technique has, during the past decade, increased rapidly, and sedimentation units with very high efficiency per unit volume have been designed. The new constructions being used are one example of a practical application in agreement with the theories already developed in the beginning of the 1900's.

4.2 Sedimentation theory

The design of sedimentation units is based on Hazen's ideal sedimentation theory developed in 1904. The ideal sedimentation concept is based upon discrete particles settling in an ideal sedimentation basin. An ideal continuous flow basin has the following characteristics:

1. The direction of flow is horizontal, and the velocity is the same in all parts of the settling zone.
2. The concentration of suspended particles of each size is the same at all points in the vertical cross section at the inlet end of the settling zone.
3. A particle is removed from the suspension when it reaches the bottom of the settling zone.

The settling in an ideal sedimentation basin takes place in exactly the same manner as in a quiescent settling column of the same depth. Hazen (1904) and Camp (1946) have suggested that the terminal velocity of a particle which settles a distance equal to the effective depth of the basin in a detention period can be thought of as an overflow rate. (FIG. 4-1):

$$v_0 = \frac{H_0}{T} = \frac{H_0}{V/Q} = \frac{H_0}{A \cdot H_0/Q} \qquad (4\text{-}1)$$

in which H_0 is the depth of the basin
T is the detention time
V is the volume of the basin
A is the surface area
Q is the flow of water

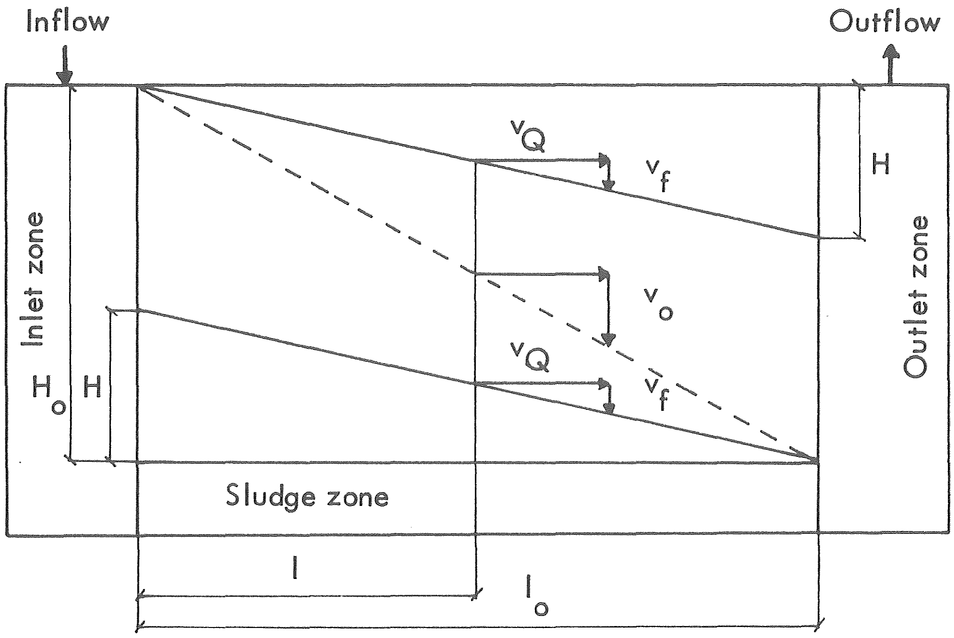


FIG. 4-1 Discrete particle settling paths in an ideal basin

Ideal discrete particles which have settling velocities v_f greater than v_o will be completely removed, and particles with settling velocities v_f less than v_o will be removed in proportion to the ratio v_f/v_o.

In the case of an ideal horizontal basin and ideal particles, the fraction C_u of the particles leaving the basin with a subsiding velocity of v_o or less can be written (Ingersoll *et al.*, 1955):

$$C_u = C_i \cdot \int (1 - \frac{v_f}{v_o}) \cdot f\ (v_f)\ dv_f \qquad (4\text{-}2)$$

in which C_i is the total concentration of particles
v_f is the settling velocity of the particles
v_o is the overflow rate

If the concentration of particles is measured in turbidity units, the expression can be written:

$$C_o = (c_{100} - c_o) \cdot f + c_o \qquad (4\text{-}3)$$

in which $f = \int (1 - \frac{v_f}{v_o})\ f(v_f)\ dv_f$

C_o is the effluent turbidity
c_{100} is the total initial turbidity
c_o is the residual turbidity

The corresponding expression for an ideal vertical sedimentation basin can be derived:

$$C_o = (C_{100} - c_o) \cdot f^1 \cdot c_o \qquad (4\text{-}4)$$

in which $f^1 = \int f\,(v_f)\,dv_f$

In practice, ideal conditions do not prevail, and therefore discrepancies from the above equations occur. Currents induced by inertia of incoming fluid, turbulent flow, and density and temperature gradients result in a reduction of settling basin efficiency. On the other hand the settling efficiency can be improved by aggregation of particles with different settling velocities during the sedimentation.

In order to describe the relationship in efficiency under ideal, y_o, and real, y, conditions, Fair, Geyer, and Okun (1968) have suggested an expression for the amount of sediment of settling velocity v_o reaching the tank bottom in time T:

$$\frac{y}{y_o} = 1 - (1 + n\,k\,T)^{-1/n} \qquad (4\text{-}5)$$

in which n is a coefficient for the actual sedimentation unit
k is a coefficient characterizing the settling properties of the particle suspension

$$(k = 1/T_o = v_o/H_o)$$

Equation (4-5) can also be written:

$$\frac{y}{y_o} = 1 - (1 + n\,v_o/\{Q/A\})^{-1/n} \qquad (4\text{-}6)$$

in which $v_o/(Q/A) = T/T_o$

The maximum removal for a value of $v_o/Q/A$ equal to unity and n equal to zero, which means good hydraulic conditions, can be calculated to 63 % instead of the 100 % which is obtained in ideal conditions. The n-value can be approximately estimated by studying the hydraulic conditions in the actual basin by means of a tracer (Thomas and Archibald, 1952).

In order to preyent light flocs from being lifted up from the sludge zone, Ingersoll, McKee and Brooks (1956) have suggested that the rate of flow v_Q should be kept below a certain value:

$$v_Q = (8/f)^{1/2} \cdot v_f \qquad (4\text{-}7)$$

in which f is the Weisbach-Darcy friction factor.

The hydraulic conditions in the actual basin are of great importance for the removal efficiency,and they can be described by the Reynold's number:

$$R = \frac{v_Q \cdot d}{\nu} \qquad (4\text{-}8)$$

in which v_Q is the rate of flow
d is a characteristic length
ν is the viscosity (kinematic)

When Reynold's number is less than 500, the flow is considered to be laminar. When R is greater than 500, the flow is turbulent and particles are submitted to random pulses in every direction. In conventional sedimentation the flow is generally turbulent. Gomella (1974) has stated that, on an average, pulses over the horizontal plane are more or less opposed to the desired settlement, and he has expressed the vertical component w of the turbulence as follows:

$$w = \frac{v_Q}{k} \qquad (4\text{-}9)$$

in which k is a constant

Various investigators have found quite similar values of k between 20 and 30. (Gomella, 1974).

According to Gomella (1974), it has been found that the settling velocity of particles arrested in a settling basin is given by:

$$v_f = \frac{Q}{A} \left(1 + \frac{L}{k\,H_0}\right) \qquad (4\text{-}10)$$

in which H_0 is the depth of the basin
L is the length of the basin

Apart from these negative effects during sedimentation, there is also a flocculation process during the sedimentation operation. The floc aggregates formed in water treatment are relatively fragile. As they grow in size, velocity gradients across them grow larger. This will break them up at some limiting size. As a rule, flocculant suspensions entering settling tanks have not yet reached this limit and sedimentation is improved materially by further floc growth. According to Camp (1946) flocculation in a sedimentation basin is due to two causes:

1. Differences in settling velocities of the flocs
2. Differences in velocity gradients in the water

Camp and Stein (1943) have formulated expressions for the number of contacts per unit volume and time N_s due to differences in settling velocities:

$$N_s = n_1 \, n_2 \, \pi \, g \, \frac{(\rho - 1)}{72\nu} \, (d_1 + d_2)^3 (d_1 - d_2) \qquad (4\text{-}11)$$

and due to differencies in velocity gradients in the water N_v (See also Eq. (2-3)):

$$N_v = n_1 \, n_2 \, \frac{1}{6} \, G \, (d_1 + d_2)^3 \qquad (4\text{-}12)$$

in which n_1, n_2 is the number of particles with diameter d_1 and d_2
ρ is the specific gravity of the particle
ν is the viscosity (kinematic)
G is the mean temporal gradient in the water

The velocity gradient caused by the drag on walls and floor is:

$$G = \left(\frac{f}{8\nu \, H} \, v_Q^3\right)^{1/2} \qquad (4\text{-}13)$$

in which f is a friction factor
H is the hydraulic radius
ν is the viscosity (kinematic)
v_Q is the rate of flow

Inserting Eq. (4-13) into Eq. (4-12) gives

$$N_v = \frac{n_1 n_2}{6} \left(\frac{f}{8\nu H} v_Q^3\right)^{1/2} (d_1 + d_2)^3 \qquad (4\text{-}14)$$

From Eq. (4-11) it can be seen that the rate of flocculation is greatest for a high concentration of particles of large size, large relative weight, and large size difference in water of high temperature. Thus, in deep sedimentation basins the flocculation will be more evident near the bottom. Flocculation due to differences in velocity gradients, Eq. (4-14), is proportional to $v_Q^{3/2}$ and inversely proportional to $H^{1/2}$, which means that this mode of flocculation is important in shallow units with relatively high velocities, for example lamella sedimentation units.

4.3 Design of sedimentation basins

Several sedimentation units have been developed in order to create favorable hydraulic conditions for the removal operation. However, the settling basins have two functions: primarily to remove the suspended solids and secondly to collect and store the sludge solids in as small a volume of water as possible to facilitate subsequent sludge handling and processing. In conventional horizontal basins the sludge is stored for some time in the basin, and therefore the hydraulic conditions are continuously changing.

In order to decrease the effect of the turbulence of the flow and stabilize the flow in basins with a relatively great depth, Lindquist (1949) has suggested two methods:

1. Flow acceleration
2. Stratification

The technique using an accelerated rate of flow has been applied in the Lovö basin, which to a great extent is used in Sweden, (FIG. 4-2).

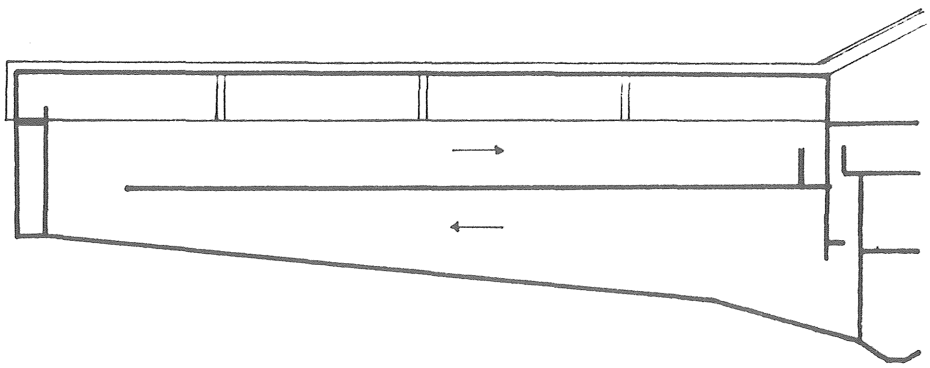


FIG. 4-2 Sedimentation basin, Type Lovö

In another type of basin, the so called Fischerström basin, the tank has been divided into sixteen horizontal channels in order to increase the settling area and to decrease the Reynold's number. The development is theoretically based upon the concept advanced by Camp (1946). The two types of sedimentation basins mentioned can be regarded as series-connected and parallel-connected conventional basins.

Gomella (1974) has shown that maximum gain can be obtained by parallel-connected compartments delimited by extra floors. But in these conventional basins, the Reynold's number is relatively high and there are practical difficulties in obtaining an even distribution of water to the different channels, especially in the parallel-connected Fischerström basin. The sludge handling is also a problem.

Due to the fact that the depth of sedimentation is independent of the overflow rate, research has lead to the development of sedimentation units with a very small depth and consequently with improved hydraulic conditions in the basin. However, in compliance with the demands on sludge handling, these units must be inclined so that the sludge is continuously removed. Such basins are called high rate sedimentation units or lamella settling basins, (FIG. 4-3).

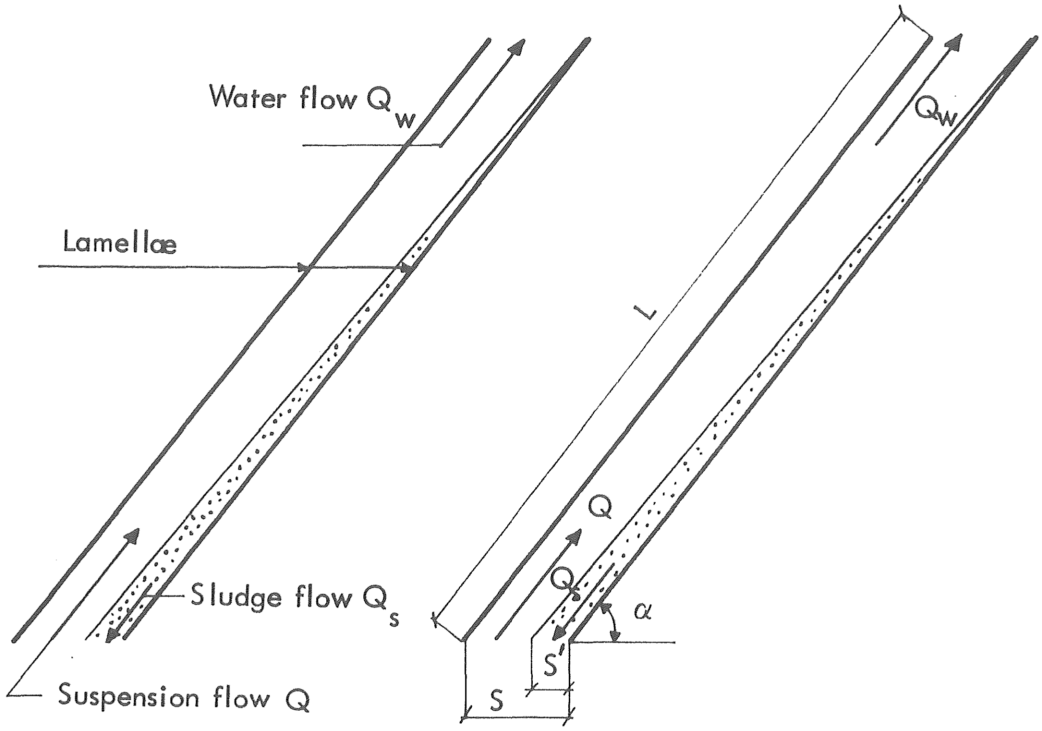


FIG. 4-3 Principle of lamella sedimentation

There are three different types of lamella separators:

1. Countercurrent or upflow unit in which the suspension and water flow are in the opposite direction to the sludge flow
2. Cocurrent or downflow unit in which the suspension and water flow directions are the same as the sludge flow direction
3. Crosscurrent unit in which the suspension and the water flow are in a direction at right angles to the sludge flow direction

The great advantage of the lamella sedimentation principle is that greatly improved hydraulic conditions are achieved and that constant hydraulic conditions are obtained due to the fact that the sludge is continuously removed.

4.4 Lovö sedimentation basin

The Lovö sedimentation basin is a double bottom basin developed from a single basin in order to stabilize the flow and decrease the effect of countercurrents. An illustration of the currents in this basin are shown in FIG. 4-4.

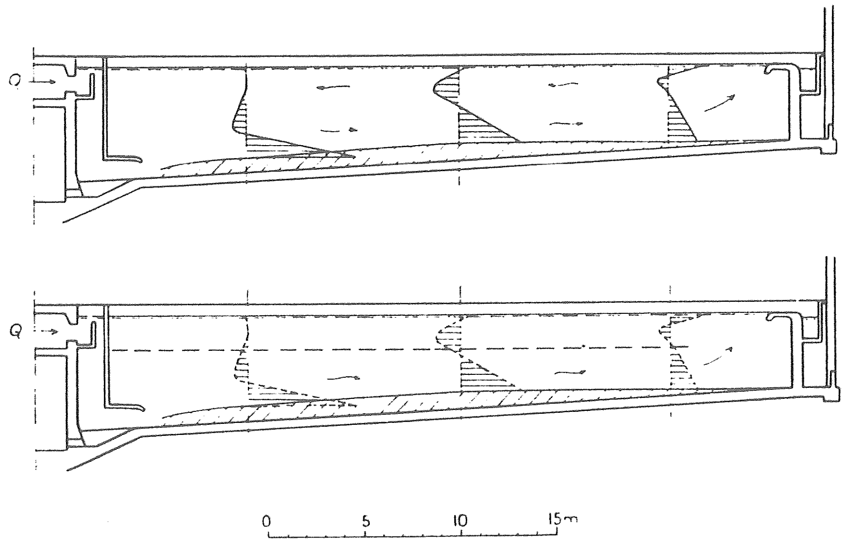


FIG. 4-4 The Lovö basin (After Lindquist, 1945).

The basin is normally designed with a relatively great depth due to the sludge storage. On the average, the sludge is withdrawn every three or four weeks. This means that the basin functions under variable hydraulic conditions. Measurements carried out have shown that the main part of the sludge is removed in the lower compartment. In the passage from the lower to the upper part, there is some mixing of water, and particles which have been close to the bottom in the lower compartment might be brought up to the top of the upper compartment by currents. The hydraulic conditions expressed by the Reynold's number are very unfavorable (R=1000-25000) and also variable. For this reason and due to the fact that the entire basin is not available (inlet zone and mixing zone) for the removal of particles, the overflow rate must be reduced to some extent. In accordance with the theories of flocculation during sedimentation (Eq. 4-11 and Eq. 4-14), some aggregation of flocs occurs primarily due to differences in settling velocities of the flocs. This effect ought to be most evident in the lower compartment.

In order to summarize the discrepancies between theory and practice, one can suggest a simplified expression:

$$\frac{Q\ \text{practice}}{Q\ \text{theory}} = \frac{Q\ \text{practice}}{v_f \cdot 2\ A} = K_{\text{Lovö}} \qquad (4\text{-}15)$$

in which Q practice is the maximum flow during practical operation

Q theory is the maximum flow during ideal conditions

v_f is the settling velocity of the floc removed

A is the surface area of the basin

$K_{Lovö}$ is a coefficient expressing the design conditions, disturbances due to turbulence, and flocculation effects. In this coefficient the significance of the method used in the determination of the settling characteristics of the flocs can also be included.

The value of $K_{Lovö}$ can be estimated with the guidance of the theories mentioned and the design of the Lovö basin to have a magnitude of 0.5-0.7.

4.5 Lamella sedimentation unit

The lamella sedimentation unit combines constant favorable hydraulic conditions with simple sludge handling. A sketch of the principle of the upflow lamella unit is illustrated in FIG.4-5.

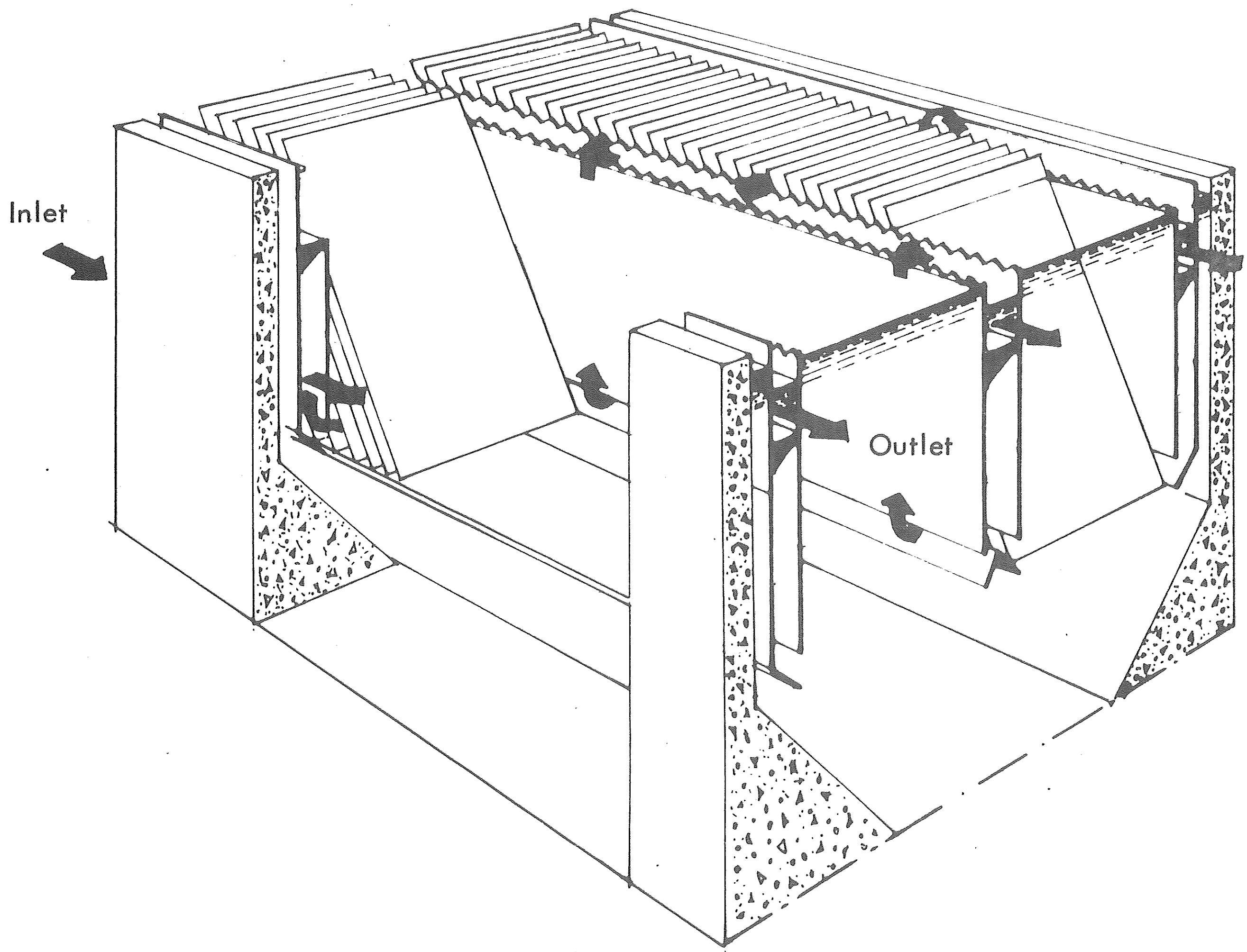


FIG. 4-5 Upflow lamella sedimentation unit

This type of sedimentation unit, which is relatively new, has so far been designed with the length of the lamellæ of the magnitude of 2-3 meters and with a horizontal clearance between the lamellæ of about 0.1 meter. The water containing flocs can be brought into the unit in different ways. In the illustrated unit (FIG. 4-5) the suspension inlet is on both sides at the bottom of the unit and the outlet is at the top. A basic condition for obtaining high performance is that the particle suspension which is led into the unit is separated or protected as effectively as possible from the sludge deposits which slide off the lamellae. In the unit shown, this is achieved to a certain extent. In this construction the inlet and outlet system is of a relatively simple design and as a consequence of this, the distribution of water over the available area is not complete. However, this might be compensated for by the fact that the separation operation is influenced in a positive direction by this counterflow principle.

Weijman-Hane (1963) applied the theory of overflow rate developed by Hazen on the lamella sedimentation unit (FIG. 4-3) and derived the following expression:

$$\frac{Q}{A} = Y_a = v_f \left(\frac{L}{S} \cos \alpha + 1\right) \qquad (4\text{-}16)$$

in which
- Q is the flow of water per lamella
- A is the horizontal area $S \cdot B$
- Y_a is the surface loading
- L is the length of the lamella
- S is the horizontal clearance between the lamellæ
- B is the width of the lamella
- α is the inclination angle

The relative significance of the different variables in Eq.(4-16) can be illustrated as shown in FIG. 4-6 and FIG. 4-7.

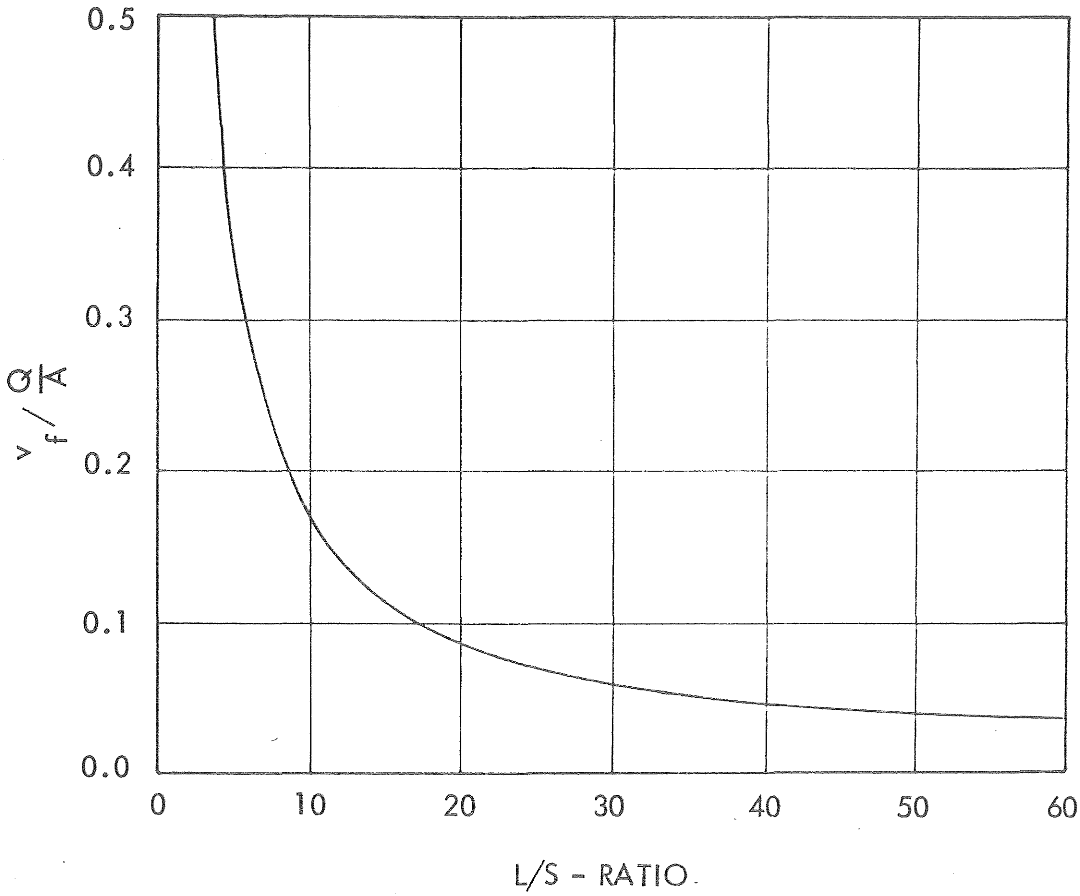


FIG. 4-6 The settling velocity v_f at a constant surface loading as a function of the L/S-ratio.

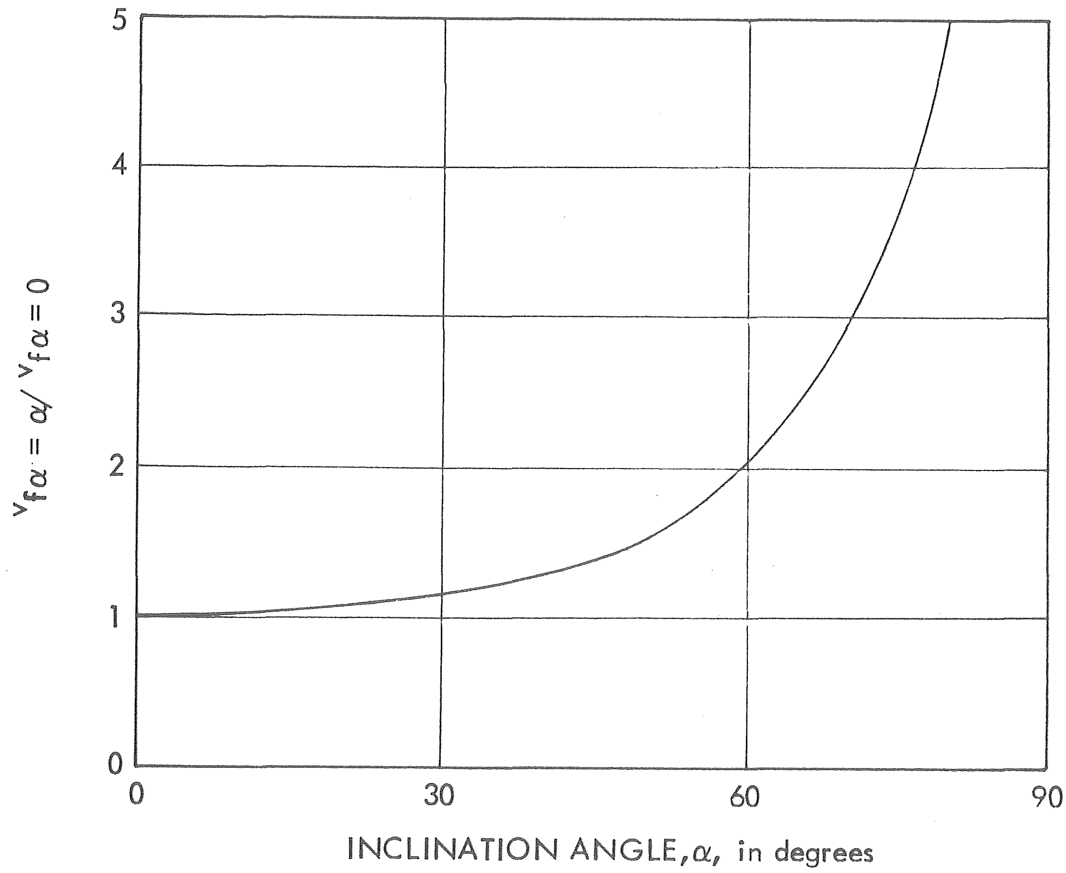


FIG. 4-7 The settling velocity at different inclinations in relation to the settling velocity with $\alpha = 0$ as a function of the inclination angle.

From FIG. 4-6 and FIG. 4-7 it can be seen that a lamella settling unit ought to be designed with a L/S-ratio of about 40 and with an inclination as small as possible. However, the sliding properties of the sludge must be considered.

As the value of L/S cos α >> 1, Eq. (4-16) may be simplified to

$$\frac{Q}{A} = v_f \frac{L}{S} \cos \alpha \qquad (4\text{-}17)$$

or $$\frac{Q}{B \cdot L} = v_f \cos \alpha \qquad (4\text{-}18)$$

From Eq. (4-18) it can be seen that the overflow rate is independent of the "depth" which corresponds to the clearance S. At maintained flow Q and decreasing S, the velocity of the flow v_Q increases and a critical velocity v_c is obtained when the deposits are re-suspended due to shearing forces between the water and the sludge. This critical velocity sets the lower limit of the clearance S. The value of the clearance is dependent on the characteristics of the flocs, especially the volume concentration. According to FIG. 4-3 the following expression can be derived:

$$\frac{Q}{B(S-S^1) \sin \alpha} \leq v_c \qquad (4\text{-}19)$$

in which S^1 is the thickness of the sludge
v_c is the critical velocity

Due to the small "depths" used in lamella sedimentation units, the Reynold's number is extremely low and the flow is laminar. The Reynold's number can for the lamella sedimentation unit be expressed as follows:

$$R = \frac{v_Q \cdot H}{\nu} = \frac{Q}{2\nu\ S(B+S \sin \alpha)} \qquad (4\text{-}20)$$

in which H is the hydraulic radius
ν is the viscosity (kinematic)

By combining Eqs. (4-18) and (4-20), the following relation is obtained:

$$R = \frac{v_f\ L \cos \alpha}{2\ \nu\ S} \cdot \frac{B}{(B+S \sin \alpha)} \qquad (4\text{-}21)$$

When S sin $\alpha << B$, Eq. (4-21) can be simplified as follows:

$$R = \frac{v_f \, L \cos \alpha}{2 \, \nu \, S} \qquad (4\text{-}22)$$

The Reynold's number increases with increasing L/S. For example, Reynold's number can be calculated to be of the magnitude of 100 for lamella separators with normal designs. This means that the flow is laminar.

It is now significant to note that the model Hazen and Camp used to develop the concept on overflow rate was an ideal, open, rectangular tank with a uniform flow across the cross section of the tank. In that model the suspended particles would follow straight lines. In the narrow conduits used in lamella sedimentation units, laminar flow is developed, and as a result, the particle paths are not straight lines. Yao (1970) has modified the Hazen model under some ideal conditions by considering the velocity distribution in the conduits. He has assumed that the flow is laminar and one--dimensional and that the suspended particles are discrete particles, which do not aggregate. No consideration has been given to the sludge. The result of these theoretical studies is presented in FIG. 4-8.

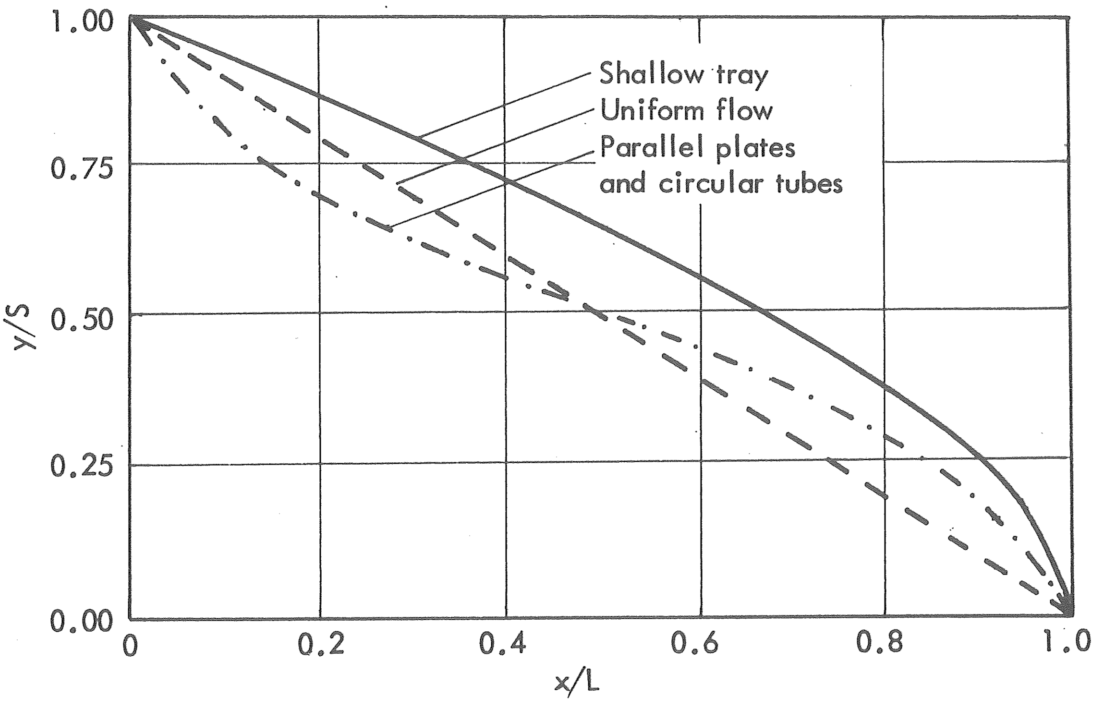


FIG. 4-8 Limiting trajectories in different types of settlers (after Yao, 1970)

Due to the velocity distribution in laminar flow, the rate of settling is higher near the plates. If the sludge current is considered, the velocity distribution is more complex and varies along the length of the lamella.

In practical applications the units are connected to the flocculation unit and there is some sort of inlet arrangement with a relatively large opening. At the entrance to the plates there exists a transition zone in which the flow may be a mixture of turbulent and laminar flow. The performance of a lamella settling unit with uniform flow is either comparable to or better than a similar system with laminar flow. The existence of the transition zone should not significantly affect the removal efficiency of the system. An expression of the transition zone L_T has been suggested by Yao (1970), Gomella (1974), and Daily *et al.*, (1956). Their expressions can be modified for lamella sedimentation purposes to:

$$L_T = \text{const} \cdot S \cdot \sin\alpha \cdot R \qquad (4\text{-}23)$$

in which S is the horizontal distance between the lamellæ
α is the inclination angle
R is the Reynold's number
The value of the constant
is 0.232 (Yao)
0.10 (Gomella)
0.065 (Daily)

Equations given earlier, which expressed the surface loading, provide uniform distribution of the suspension over the entire lamella plate. However, because of the inlet system, differences in the distribution of suspension exist. Therefore, the equations must be modified. In FIG. 4-9 some useful definitions have been illustrated.

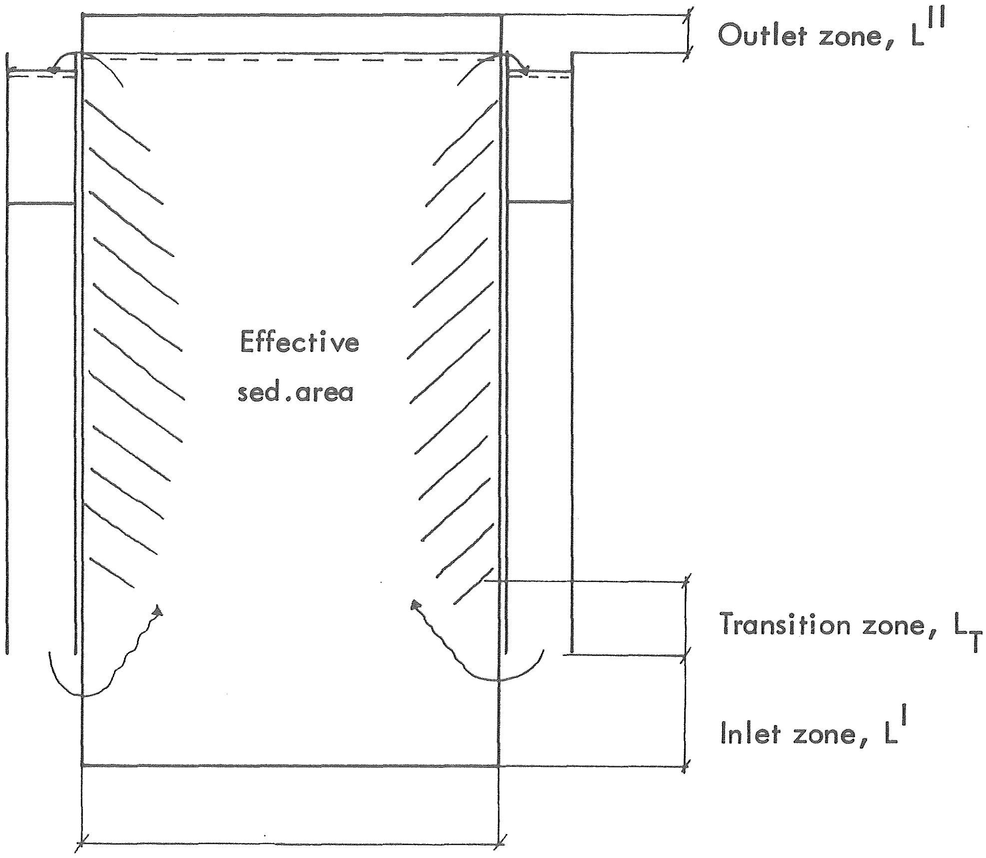


FIG. 4-9 The distribution of the suspension over the lamella plate

The inlet and outlet system and the relationship between the length and the width of the lamella affect the effective sedimentation area. The effective part can be expressed by a factor $K_{lamella}$ which is design dependent and to a certain extent current dependent. The relationship between practice and theory can be expressed:

$$\frac{Q\ \text{practice}}{Q\ \text{theory}} = \frac{Q\ \text{practice}}{v_f \cdot BL \cos\alpha} = K_{lamella} \qquad (4\text{-}24)$$

in which Q practice is the maximum flow during practical operation

Q theory	is the maximum flow during ideal conditions
v_f	is the settling velocity of the flocs removed
B	is the width of the lamella
L	is the length of the lamella
α	is the inclination angle
$K_{lamella}$	is a coefficient expressing the design conditions, hydraulic conditions affected by disturbances due to turbulence and flocculation effects. In this coefficient the significance of the method used to determine the settling characteristics of the flocs can also be included.

According to FIG. 4-9 a further division of the coefficient $K_{lamella}$ can be suggested. From the design point of view there is some part of the lamella length which is not functioning. The effective lamella length is $L_{eff} = L - L_i$. Depending on the design of the inlet and outlet system, the distribution of water over the lamella plate is more or less complete. This can be regulated by a coefficient K_A. But even if the water is distributed over the area equal to $B \cdot (L - L_i) \cdot K_A$, it is not certain that this area is available for sedimentation.

The effective sedimentation area may thus be written:

$$A_{eff} = B \, (L - L_i) \, K_A \cdot K_T \qquad (4\text{-}25)$$

in which L_i is the ineffective length of the lamella
K_A is a coefficient dependent on the design of the inlet and outlet system of water
K_T is a coefficient dependent on the hydraulic conditions.

The equation for the lamella sedimentation unit may thus be written:

$$\frac{Q}{v_f \, B \, (L - L_i) \cos \alpha} = K_A \cdot K_T \qquad (4\text{-}26)$$

From an optimization point of view and in order to obtain comparisons between conventional sedimentation (Lovö basin) and lamella sedimentation, it is desirable to be able to evaluate the hydraulic conditions in the different basins and the function in general. On the basis of a standardized settling velocity analysis, it is also of interest to describe the particle concentration leaving the sedimentation unit under various conditions.

5 SEDIMENTATION, EXPERIMENTAL INVESTIGATION

5.1 Objectives of the sedimentation investigation

The objectives of the investigation have been to combine earlier tests of lamella sedimentation units and with supplementary tests carried out in this investigation under real full scale operation, create a model of the sedimentation operation, useful for optimization purposes. The intention has also been to compare conventional sedimentation with the new lamella sedimentation technique.

5.2 Lamella sedimentation

5.2.1 Scope of the investigation

During development of the lamella sedimentation units, different units with more or less ideal design were studied. Some studies have already been reported (Weijman-Hane, 1963) but an analysis of all the units used has not been made. In this report the results obtained during full scale operation at the Water Treatment Plant at Lackarebäck are also discussed.

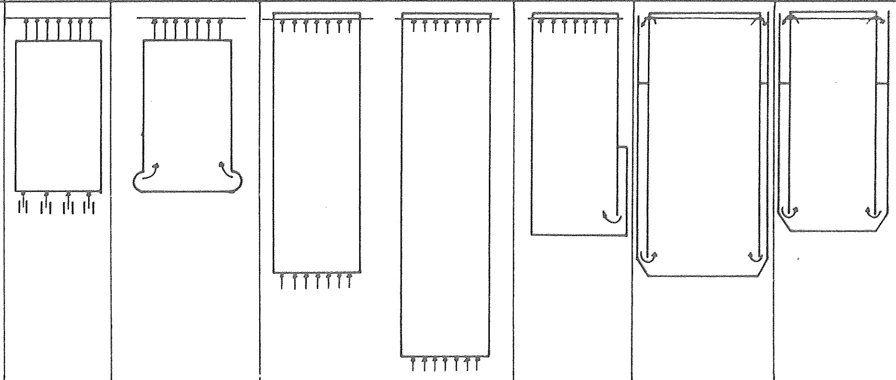
The following variables concerning the lamella sedimentation have been studied:

Inclination angle α
Length of lamella L
Horizontal distance between the lamellæ S
Width of lamella B
Inlet and outlet system

In FIG. 5-1 the units examined are illustrated by some data.

During the investigations the conventional basins have been used as reference units. The number of tests carried out was very large. A full scale unit has been studied during a period of almost four years under varying conditions. A detailed description of the different units can be found in Appendix 1.

LAMELLA TYPE	1	2		3 A	3 B	4	5 A	5 B



DATA:								
No. of lamellae	5	5	9	1	1	2x29	4x55	8
Lamella distance m	0.1	0.1	0.06	0.15-0.35	0.15-0.35	0.1	0.08	0.02-0.2
Lamella length m	1.8	1.8	1.8	3.0	4.0	2.5	3.0	2.5
Lamella width m	1.0	1.0	1.0	1.0	1.0	1.0	1.25	1.0
Inclination angle °	40-60	55	55	55	55	55	55	55
Flocculation time min	60	45	30-60	60	60	60	60-100	30-60
Alum mg/l	50	50	40	45	45	50	40	40
Act. silica mg/l	0-8	8	0-6	8	8	9	8	0-10
Raw water	Göta älv	Göta älv	Delsjön	Göta älv	Göta älv	Göta älv	Delsjön	Bolmen
Plant	1	1	3	-	-	2	8	4
Report	B 63:3	B 69:1	Here	B66:4	B 66:4	Here	Here	Sydvatten

FIG. 5-1 Sketch of the principles of the different units investigated

5.2.2 Inclination angle

Weijman-Hane (1963) showed that an inclination angle α of at least 55° is necessary in the upflow lamella sedimentation unit in order to get the sludge to slide off. Ever since, this inclination angle has been used in all the units tested.

5.2.3 Relationship between settling velocity of flocs and surface loading

Determination of the settling characteristics of the flocs creates some difficulties as the floc properties are always affected more or less during the analysis; but, if the analysis is carried out in the same way each time, a useful relative value of the settling velocity can be calculated, and comparisons with results obtained in a particular sedimentation unit can be made.

In two pilot plants (Pilot plant No. 1 and pilot plant No. 4) different floc suspensions have been tested and some information on the relationship between the settling velocity and the surface loading has been obtained. The result is summarized in TABLE 5-1.

TABLE 5-1 Settling velocity of flocs and surface loading. Mean values

Chemicals mg/l	Settling velocity of floc v_{fo} m/h	Surface loading Observed m/h	Surface loading Calculated Eq. 4-17 m/h	Pilot Plant No.	Sed. depth m	Number of settling analyses
Alum 40 Lime Superfloc. 0,0.1	0.7	18	20.5	4	0.12	6
Alum 40 Lime Act. silica 3	0.8	25	23.5	4	0.12	-
Alum 40 Chalk 80 Carbon-dioxide 15 Superfloc. 0.1	1.0	30	29.5	4	0.12	5
Ferric-chloride 70 Lime Act. silica 5	1.6	50	47.5	4	0.12	3
Alum 50 Lime Act. silica 9	0.9	10	10.0	1	0.30	12
Alum 50 Chalk 80 Carbon dioxide 15 Act. silica 9	1.1	13	13.0	1	0.30	22

The table shows that an increase in settling velocity leads to a corresponding increase in the surface loading which agrees with the theory expressed in Eq. (4-17). It must be pointed out that the settling velocity analysis has been carried out at a relatively small sedimentation depth and has thus been adapted to the lamella sedimentation unit.

Villemonte *et al.*, (1966) have also suggested that a "critical settling depth" can be used as a performance criterion in several ways, for example (a) as a measure of the effectiveness or efficiency of a given basin to handle a given floc; or (b), as a criterion for comparing the performance of various inlet systems.

5.2.4 Relationship between the L/S-ratio and the surface loading

During investigations of the lamellæ of design types 3 and 4 (FIG. 5-1), Weijman-Hane and Rosén (1966) found that low values of the L/S-ratio (values of about 10) gave an overflow rate (Q/BL) lower than that obtained at higher values of the L/S-ratio, although the overflow rate ought to have been the same according to the theory, Eq. (4-18). The surface loadings obtained corresponded, according to Eq. (4-17), to a settling velocity of between 1.5-2.5 m/h. The explanation of the higher overflow rate obtained at a high L/S-ratio was assumed to be due to the hydraulic conditions in the unit. At a high L/S-ratio the flow was regarded as stable, while the flow was unstable at a low L/S-ratio. The flow characteristics were calculated by the Froude's number F ($F=v_Q^2/g \cdot H$) which is supposed to be less than 10^{-5} for a stable flow. It is, however, doubtful if the Froude's number can be used in this sedimentation technique as a measure of the stability of the flow. The floc suspension is affected by the design of the particular sedimentation unit, and the settling velocity of the floc may be increased as the possibilities of aggregation are improved at higher L/S--ratios according to the Eq. (4-14). This might be the explanation for the above observation. Results obtained in another pilot plant, type 7, (FIG. 5-1), with a constant lamella length (L=2.5 m) and with varying distances between the lamellæ, S=0.05 to 0.20 m, did not show any difference in overflow rates at different L/S-ratios (TABLE 5-2).

TABLE 5-2 Results from pilot plant No. 4

L/S-ratio m/m	Surface load, $\frac{Q}{A}$ observed m/h	Settling velocity v_{fo} calculated m/h
2.5/0.20	7	0.85
2.5/0.10	13	0.84
2.5/0.05	25	0.84

The difference in the design between the lamella sedimentation units type 4 and type 7 are the lamella length and the inlet system. In the latter type the suspension inlet is on both sides at the bottom of the lamellæ, and the water is withdrawn at the top on both sides. The reason why the overflow rate does not increase with increasing L/S-ratio in a sedimentation unit with a simple inlet system may be due to the fact that the flocs are affected by too high shearing stresses at the inlet and thus a further aggregation of flocs is prevented. However, some results in a simplified lamella sedimentation unit show that a longer lamella gives an improved result – lower residual turbidity – which may be due to favorable hydraulic conditions being established.

Thus, in some lamella sedimentation units a result in agreement with the theory has been obtained. In some units in which the L/S-ratio has been considerably changed, other conditions were also simultaneously changed and this in turn affected the result. Equation (4-18) shows that the distance S between the lamellæ does not have any significance for the waterflow Q through the lamella unit. The distance consequently ought to be as small as possible in order to increase the performance. However, it is necessary to consider the floc volume in the suspension in order to avoid critical velocities v_c in the lower part of the unit (FIG. 4-3). The critical velocity determines the lower limit of the horizontal distance S between the lamellæ in accordance with Eq. (4-19)

$$\frac{Q}{B(S-S^1)\sin\alpha} \leq v_c$$

in which S^1 is the thickness of the sludge

Investigations carried out with very small horizontal distances show that the critical velocity typical for this particular suspension is reached at a distance of about 0.04 m (FIG. 5-2). At distances greater than 0.05 m, the flow of water is relatively constant. The value of the critical velocity can be calculated to be 0.005 m/s.

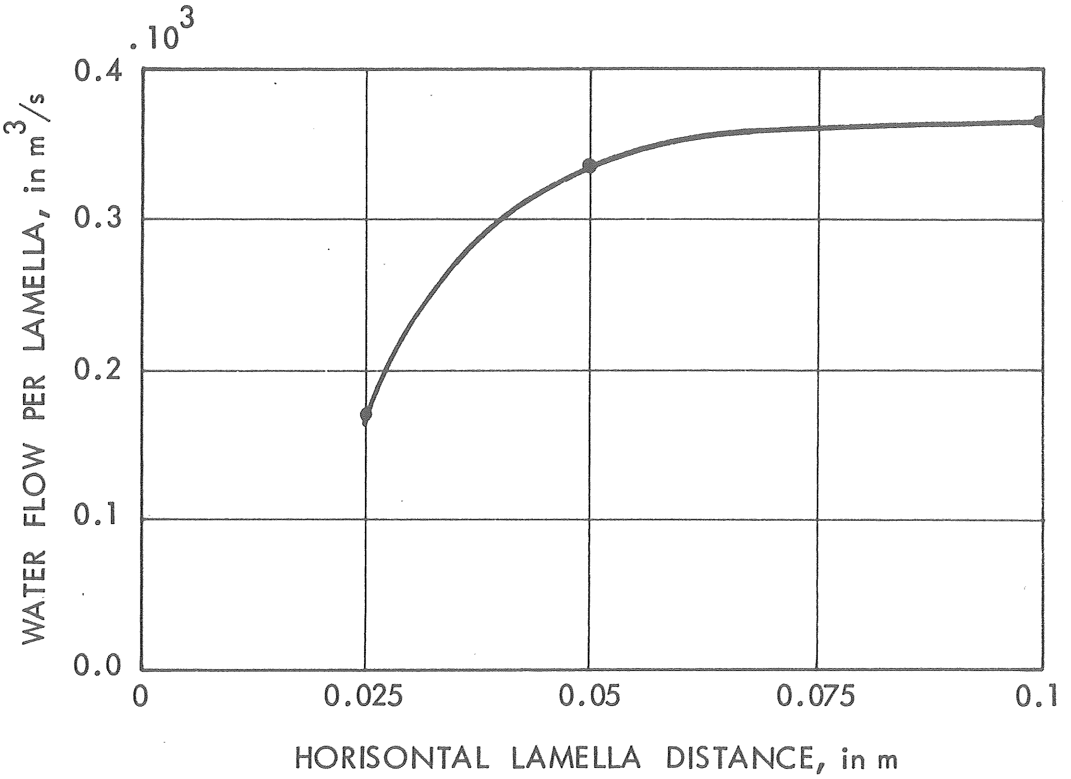


FIG. 5-2 Flow of water per lamella as a function of the horizontal distance between the lamellæ (Pilot plant No.4)

5.2.5 Settling velocity – sedimentation analysis

So far different variables and their effect on the sedimentation operation have been accounted for. With the settling analysis as a basis, it still remains to try to obtain an absolute value of the settling velocity v_{fo} of the flocs which are completely removed and also to estimate the change in the floc characteristics during the sedimentation in relation to the result obtained in the settling analysis.

Earlier investigations (Weijman-Hane, 1963) have clearly shown that laminar flow is established in the lamella sedimentation unit. Reynold's number has been calculated at less than 100, which is considerably less than the 500 mentioned as a limit of laminar conditions. Thus, there is no reason to adjust the sedimentation efficiency because of turbulence, except at the inlet zone. The results obtained in the different units have been coordinated in TABLE 5-3. In the table the effect of some corrections have been shown. With consideration given to the particular de-

TABLE 5-3 Result of lamella sedimentation. Mean values. (Definitions, see FIG. 4-9)

1 DESIGN													
Type of lamella unit		1	2	2	3	3	3	3	3	4	5	5	5
Lamella length L,	{m}	1.8	1.8	1.8	3.0	3.0	0.4	4.0	4.0	2.5	3.0	2.5	2.5
Lamella distances S,	{m}	0.1	0.1	0.06	0.3	0.1	0.35	0.20	0.15	0.1	0.08	0.05	0.1
Lamella width B,	{m}	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.25	1.0	1.0
L/S-ratio		18	18	30	10	30	11.5	20	26.6	25	37.5	50	25
2 RESULT													
Water temperature t,	{°C}	14	10	15	16	16	16	10	5	2	10	10	10
Turbidity red	{%}	94	90	94	94	97	93	94	91	90	90	90	90
Surface rate Q/A,	{m/h}	14	9	11	8	47	14	18	30	16	18	26	13
3 CALCULATION OF SETTLING VELOCITY v_{fo},													
3a Without correction (Eq.4-18)													
Settling velocity	{m/h}	1.4	0.8	0.7	1.4	2.7	2.1	1.6	1.9	1.1	0.8	0.9	0.9
3b Correction: L_{eff} and A_{eff}													
L' + L"	{m}	0.2	0.3	0.3	0.1	0.1	0.1	0.1	0.1	0.25	0.35	0.35	0.35
A_{eff}		1	0.8	0.8	1	1	1	1	1	0.8	0.7	0.7	0.7
Settling velocity	{m/h}	1.5	1.3	1.0	1.5	2.8	2.2	1.6	2.0	1.6	1.4	1.5	1.5
3c Correction: L_{eff}, A_{eff} and L_T													
L' + L"	{m}	0.2	0.3	0.3	0.1	0.1	0.1	0.1	0.1	0.25	0.35	0.35	0.35
L_T	{m}	1.3	0.7	0.4	6.4	4.2	15.3	5.6	4.6	1.0	0.9	0.5	1.0
A_{eff}		1	0.8	0.8	1	1	1	1	1	0.8	0.7	0.7	0.7
Settling velocity	{m/h}	2.2	1.6	1.2	2.5	6.9	5.6	3.2	4.2	2.0	1.9	2.0	2.0

sign and the transition zone, the effective lengths of the lamellæ have been calculated. In calculation of the transition zone L_T Eq. (4-23) has been used. If the vertical component of the turbulence (w=v/K) is considered, the settling velocity of the flocs can be calculated by the following simplified expression.

$$v_{fo} = \frac{Q}{A}\frac{S}{L}\left(\frac{1}{\cos\alpha} + \frac{L_T}{20 \cdot S \sin\alpha}\right) \qquad (5\text{-}1)$$

Equation (5-1), together with the calculation of the transition zone L_T, Eq. (4-23), seems to give unreasonable results, at least for distances between the lamellæ larger than 0.1 m. It is therefore not recommendable to use this equation (Eq. 5-1). For a conventional sedimentation basin a decrease in sedimentation performance of 20 to 40 per cent due to turbulence is a consequence of Eq. (4-10), suggested by Gomella. In sedimentation units with small horizontal distances, the overflow rate is reduced by about 20 per cent in accordance with Eq. (5-1). Bearing in mind that the flow conditions in a lamella sedimentation unit are not fully investigated, one could estimate the reduction in overflow rate due to disturbances in the inlet zone to maximum 20 per cent.

In the different lamella sedimentation units the effective sedimentation area A_{eff} is mainly dependent on the design of the inlet and outlet water system. The effective area in a unit with a simple inlet system (Type 5, FIG. 5-1) has been estimated to be 0.6 to 0.7 of the entire area. The test was carried out with tracer studies, and the value is valid for relatively normal surface loadings

The value of K_T (Eq. 4-26) can be estimated approximately at 0.8. The settling velocity (v_{fo} to v_{f10}), which has been calculated in TABLE 5-3, corresponds to a result which has been reached at different water temperatures. The settling velocity, useful for design purposes, may be estimated at 1.6 ± 0.2 m/h at a removal efficiency of 90 %. The settling velocities mentioned agree with those obtained during the settling velocity analysis when the sedimentation depth was between 0.5 and 1.0 m.

5.2.6 Lamella sedimentation model

For optimization purposes, information on the particle concentration leaving the sedimentation basin is necessary even at higher surface loadings than those corresponding to the settling velocity v_{fo}. It is also necessary to consider the increasing amount of sludge which has to be removed at higher surface loadings. Thus, it is important to formulate a more general equation than Eq. (4-26). This can in principle be expressed as:

$$\frac{Q}{v_f\, B(L_{eff} - K_S)\cos\alpha} = K_A \cdot K_T \qquad (5\text{-}2)$$

in which K_T is a coefficient of disturbances and flocculation effects due to turbulence

K_A is a coefficient of the available effective sedimentation area. The coefficient is dependent on inlet system, outlet system, width of lamella and flow rate

K_S is a term for the thickness of sludge which affects the relative lamella distance and indirectly adjusts the effective part of the lamella length

L_{eff} is the effective length of lamella ($L_{eff} = L - L_i$)

The influence of turbulence has been discussed earlier, and it was assumed that the overflow rate was reduced with a maximum factor of 0.2 of the theoretical value. This value must always be related to the method used to determine the settling characteristics of the flocs. From Eq. (4-14) it can be seen that the collision frequency of the flocs is increased with an increased rate of flow.

In order to compare the hydraulic conditions in lamella units of different designs, especially the inlet and outlet system, one can plot the relative settling velocity of the floc v_f/v_{fo} against the turbidity in the water leaving the unit. Such a plot is illustrated in FIG. 5-3 in which the result from a unit with an ideal suspension distribution over the whole width of the lamella plate is compared with that obtained in a lamella unit with the inlet on both sides at the bottom. In the latter unit two different floc suspensions, one with low settling velocity and one with a higher settling velocity, are shown.

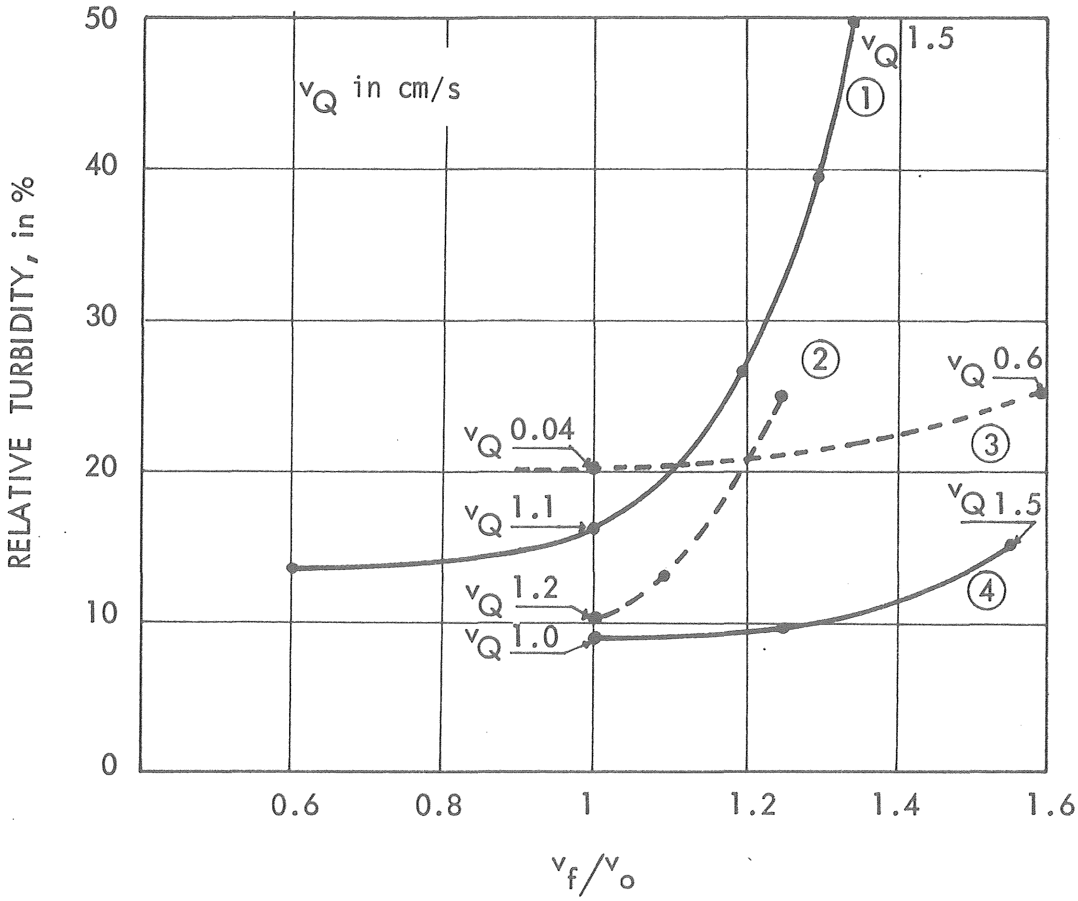


FIG. 5-3 Relationship between v_f/v_o and turbidity in the effluent

1. inlet from the sides L/S = 3/0.08 A_{eff} 0.7
2. inlet from the sides L/S = 2.5/0.05 A_{eff} 0.7
3. inlet from the sides L/S = 2.5/0.1 A_{eff} 0.7
4. ideal distribution L/S = 4/0.15 A_{eff} 1.0

A_{eff} is the area available for sedimentation
v_Q is the rate of flow

From the result obtained it is evident that the two units function quite differently. The conclusions which may be drawn from this comparison are that the hydraulic conditions are of great importance and that due to the inlet system they differ. It seems to be difficult to suggest a definite value of the critical velocity v_c at which the deposits are taken away, but the rate of flow v_Q ought to be kept below 1.0 to 1.5 cm/s. This value can be compared to the value determined by Eq. (4-7). The value of the friction factor for this particular type of sediment has been suggested to be $f=2.5 \cdot 10^{-2}$ by Fair *et al.* (1968). From the Eq. (4-7) with a floc settling velocity of 1.6 m/h, it can be calculated that flocs are eroded at a rate of flow of 0.8 cm/s.

In a lamella unit, with inlets only at the sides, the rate of flow is rapidly changed at the inlet zone and the critical velocity can be exceeded. Also, at the outlet zone critical velocity might be reached if the water is withdrawn only at the sides. In order to get an agreement between the theory based on the settling velocity distribution function obtained at the settling analysis and the result obtained in practice, one must adjust the available sedimentation area in the particular lamella unit as a function of the rate of flow if this rate exceeds critical values for sedimentation. Such an adjustment demands knowledge of the flow distribution in different designs and at different rates. However, such investigations have not yet been carried out; therefore, only an assumption can be made about the change in effective area.

The surface loading increases with decreasing distance between the lamellæ in accordance with Eq. (4-16). At small distances the thickness of the sludge S^1 must be considered in accordance with Eq. (4-19). It is difficult to measure the exact value of the thickness of the sludge as the sludge deposited on the plates does not slide off continuously and smoothly. It is more likely that a certain amount of sludge is brought together and the "mound" of sludge slides or rolls down the plate. Such observations have been made in units equipped with sides of plexiglass. However, the thickness of the sludge can be estimated at about 1-2 cm for the suspension tested at normal surface loadings with length of lamella of up to 2.5 or 3.0 m. The sludge thickness is dependent on the volume concentration of the actual suspension and on the surface loadings. At high volume concentrations, which means a reduced value of $(S - S^1)$, it has been observed that the length of lamella functioning decreases at the same time as the removal efficiency somewhat improves due to a higher collision frequency of flocs. The term changing the length of lamella functioning at a relatively decreased value of $(S - S^1)$ might be expressed as:

$$K_S = \frac{\gamma'}{S - \gamma'' v_Q} \qquad (5\text{-}3)$$

in which γ' γ'' are coefficients

v_Q is the velocity of flow between the lamellæ

S is the horizontal distance between the lamellæ

Thus, Eq. (5-2) can in terms of surface loading be written:

$$\frac{Q}{A} = Y_a = v_f \cdot K_T \cdot K_A \frac{1}{S} \left(L_{eff} - \frac{\gamma_0}{S - \gamma Y_a}\right) \cos \alpha \qquad (5\text{-}4)$$

in which v_f is the settling velocity of the flocs

K_T is a coefficient dependent on the hydraulic conditions

K_A is a coefficient dependent on the design of the inlet and outlet systems of water

S is the horizontal distance between the lamellæ

L_{eff} is the effective length of the lamella from the design point of view

γ_0, γ are coefficients for the sludge volume

α is the inclination angle

5.3 Conventional sedimentation – Lovö basin

5.3.1 Scope of the investigation

In connection with the studies of the lamella sedimentation, comparisons have been carried out with the Lovö basin at the Alelyckan Water Treatment Plant as well as the Lackarebäck Water Treatment Plant. The investigations have included studies of the removal efficiency at varying surface loadings and water temperatures.

5.3.2 Lovö basin, results

The results obtained under normal operation at the water treatment plants at Alelyckan and Lackarebäck at different water temperatures are briefly summarized in FIG. 5-4 and FIG. 5-5.

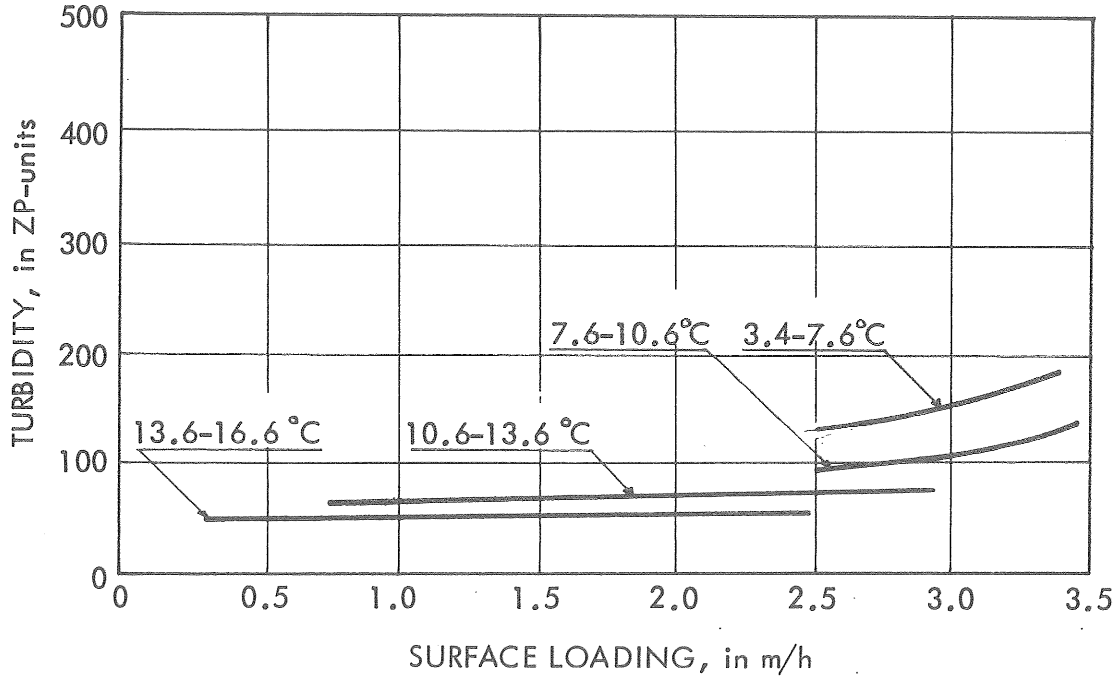


FIG. 5-4 Turbidity in effluent water as a function of surface loading. Lovö basin at the Alelyckan Water Treatment Plant.

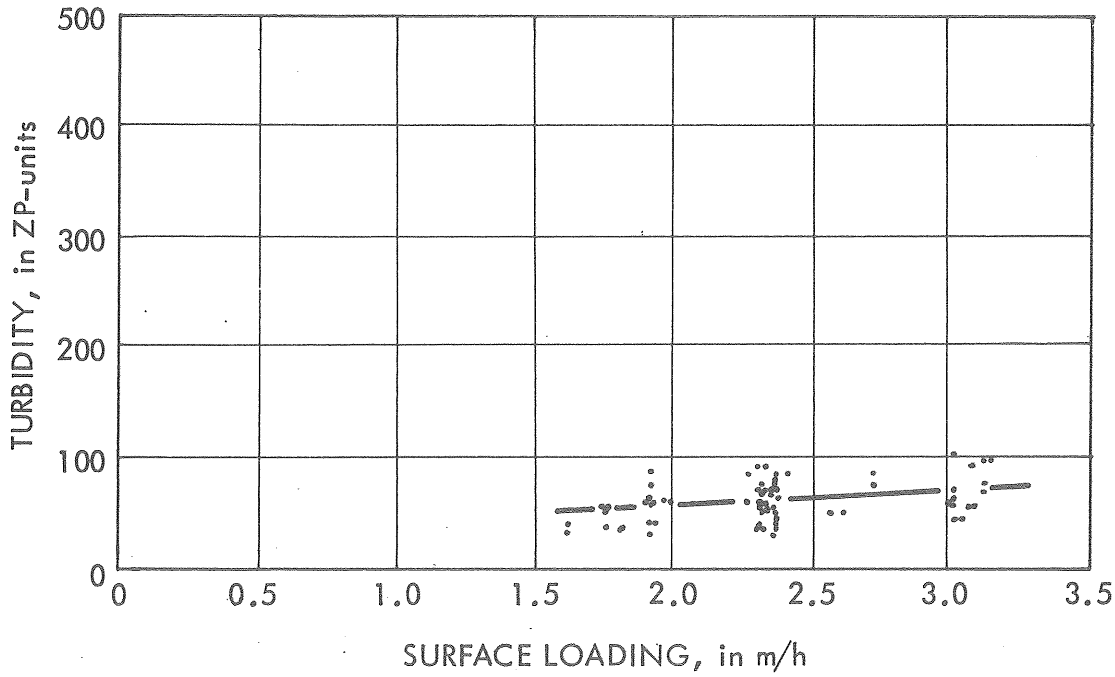


FIG. 5-5 Turbidity in effluent water as a function of surface loading. Lovö basin at the Lackarebäck Water Treatment Plant.

From the results it can be seen that no definite breakup has been reached. There are, however, some practical difficulties in carrying out tests in full scale. The results shown represent a normal operation. Apart from this test, a special capacity study was carried out at Lackarebäck in October 1975 with the same result.

5.3.3 Conventional sedimentation model

Equation (4-1) may be used as a basis for a model of the conventional Lovö basin. Only the extra bottom in the Lovö basin has to be considered. The equation under ideal conditions can be expressed in accordance with Eq. (4-15)

$$\frac{Q}{A} = v_f \cdot 2 \qquad (5\text{-}5)$$

Under real conditions the removal efficiency is, as mentioned earlier, affected by different factors. The relationship expressed in Eq. (4-15) between practice and ideal conditions is complicated as the decrease in removal efficiency due to turbulence (Eqs. 4-6 and 4-10) and the increase in removed efficiency due to floc aggregation, Eq. (4-11) and Eq. (4-14), are dependent on the rate of flow. An assumption that these two factors completely neutralize each other gives a constant value of the coefficient $K_{Lovö}$. A magnitude of about 0.7 can be suggested and Eq. (4-15) can be written:

$$\frac{Q_{practice}}{A} = 0.7 \cdot v_f \cdot 2 \qquad (5\text{-}6)$$

in which Q is the flow of water
A is the surface area of the basin
v_f is the settling velocity of flocs

5.4 Model describing the concentration of particles in the effluent from sedimentation basins

Earlier, two equations (Eq. 4-3 and 4-4) describing the particle concentration (turbidity) in the effluent from an ideal horizontal and a vertical basin have been given. The discrepancies between the lamella sedimentation unit and the Lovö basin, and the ideal horizontal basin. are great. It is assumed that the model of the ideal vertical basin can be applied. In describing the settling characteristics of the flocs Eqs. (2-38), (3-5), and (3-9) will be used. With respect to Eqs. (3-5) and (3-9), it has been demonstrated earlier that the sedimentation depth affects the shape of the settling velocity distribution curve. It is desirable to describe the turbidity in the effluent from a conventional basin as well as from a lamella sedimentation unit by a standardized settling analysis. However, the investigations have shown that small horizontal distances between lamellæ or small values of the differences $(S - S^1)$ give a high removal efficiency and also affect, to a measurable extent, the standard deviation of the settling velocity distribution. Thus, if the two types of basins should be compared, the difference in characteristics must be considered. In accordance with Eq. (4-14) the flocculation effect is proportional to the hydraulic radius, raised to 1/2, and thus it seems reasonable to suppose that the standard deviation varies in some proportion to the hydraulic radius. But, as there is almost the same relationship between the depths of the conventional and lamella sedimentation basins as there is between the hydraulic radius of the two units, it is suggested that the sedimentation depth can be used as a parameter adjusting the floc characteristics.

The aforementioned Eq. (3-9) may thus be written as follows:

$$c_o = \frac{k_g \sigma}{f(S)} + l_g \qquad (5\text{-}7)$$

in which k_g, l_g are coefficients

$f(S)$ is a function of the sedimentation depth. In conventional sedimentation basins, the vertical depth may be used, and in lamella sedimentation the horizontal distance S between the lamellæ may be used.

In describing the turbidity as a function of the surface load, one can combine Eqs. (5-4) and (5-6) for the lamella unit and the Lovö basin, respectively, with Eqs. (2-38), (3-3), (3-5), (4-4), and (5-7) mentioned earlier. The equations for the model describing the turbidity in the effluent can thus be summarized:

Equations for the lamella and the Lovö basin

$$C_o = (c_{100} - c_{o,\ i}) \int f(v_f)\ d\ v_f \qquad (4\text{-}4)$$

$$f(v_f) = \frac{1}{\sqrt{2\ \pi}} \exp \left[- \frac{(v_f - v_m)^2}{2\ \sigma^2} \right] \qquad (2\text{-}38)$$

$$c_o = \frac{100 + a}{1 + b\ v_f^{1.25}} - a \qquad (3\text{-}5)$$

$$c_o = \frac{k_g\ \sigma}{f(S)} + l_g \qquad (5\text{-}7)$$

Basic values of $c_{o,\ i}$ are obtained from Eq. (3-3)

Equation for the lamella sedimentation,

$$\frac{Q}{A} = v_f \cdot K_T\ K_A \cdot \frac{1}{S} \left(L_{eff} - \frac{\gamma_o}{S - \gamma\ Q/A}\right) \cos \alpha \qquad (5\text{-}4)$$

Equation for the Lovö basin,

$$\frac{Q}{A} = 0.7 \cdot 2 \cdot v_f \qquad (5\text{-}6)$$

in which C_o is the turbidity in the effluent

c_{100} is the turbidity in the influent

c_o is the residual turbidity

v_f is the settling velocity of the flocs

v_m is the mean settling velocity of the flocs

σ is the standard deviation of the settling velocity of the flocs

a, b are coefficients depending on the depth used in the sedimentation analysis

$f(S)$ is a function of the sedimentation "depth" in the sedimentation unit

k_g, l_g are coefficients dependent on the characteristics of the suspension

K_T is a coefficient dependent on the hydraulic conditions

K_A is a coefficient dependent on the design of the inlet and outlet system of water

γ_o, γ are coefficients for the sludge volume

The values of the coefficients are shown in Appendix 9.

The above summarized system of equations has been tested using some interesting data and the result is illustrated in FIG. 5-6 and in FIG. 5-7.

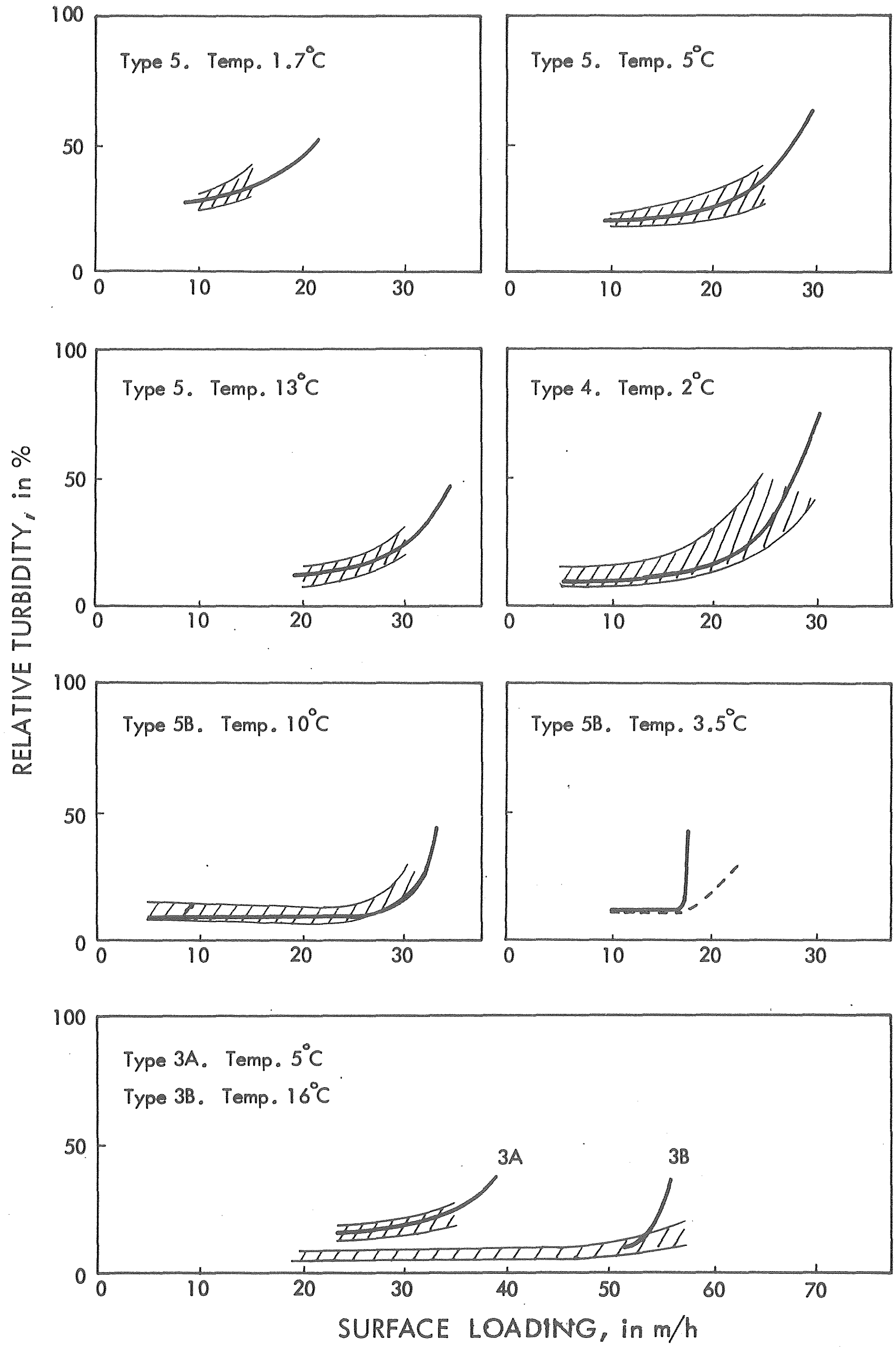


FIG. 5-6 Turbidity in the effluent from lamella sedimentation units of different designs.
Theory: heavy lines. Observed values: fields.

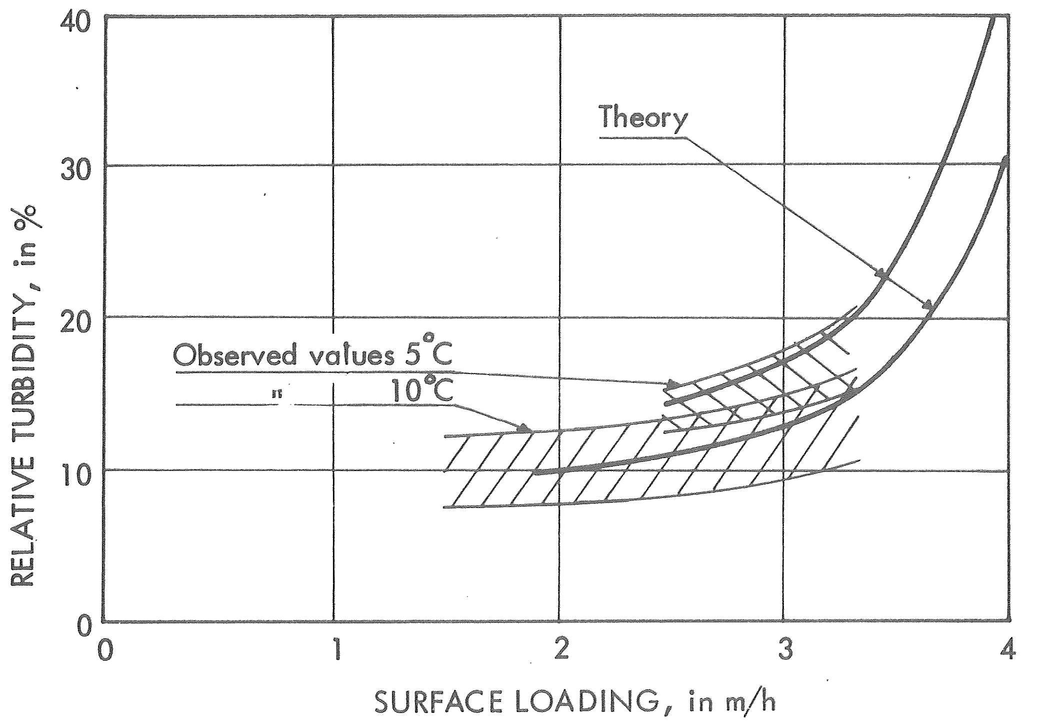


FIG. 5-7 Turbidity in the effluent from a Lovö basin

The graphs show that there is a very good agreement between the theoretical calculations and the measured result, and thus, the developed model can be used for optimization purposes.

5.5 Discussion of the model describing the turbidity in effluents from sedimentation units

In the proposed model describing the turbidity in the effluent from sedimentation units, several relationships, each containing some approximations are included. In describing the turbidity in effluents from different basins, based on settling analysis, corrections for the particular sedimentation unit must be done. The hydraulic conditions in the sedimentation basins change with the surface loads. This has not been completely considered; it has only been assumed that the increase in the settling velocity of the flocs due to a higher collision frequency with increasing rate of flow neutralizes the deteriorated hydraulic conditions.

More specialized studies are needed to verify the assumptions made but, from an optimization point of view, it is less important to have the exact relationship, provided that the problems to be studied do not to a great extent differ from the conditions reported.

5.6 Safety factor

The design of a sedimentation unit ought to be carried out in conjunction with the design of the filter operation and with the guidance of earlier studies of water treatment plants. The purpose of any experimental sedimentation measurement is that the characteristics of a critical subsiding particle can be defined well enough to permit proportioning of the tankage designed to process the suspension. Obviously, certain problems arise in translating the behavior of a suspension in a settling tube to its behavior in a full scale sedimentation plant. The choice of an acceptable safety factor requires judgement and experience. Safety might vary in different sedimentation units. In this investigation the influence of the hydraulic conditions on the performance has been discussed, but there is always a deviation in the results due to disturbances of different kinds. However, if the maximum capacity of the sedimentation unit has been estimated, the safety factor may be slightly above unity or at least less than 1.5.

5.7 Comparison between conventional and lamella sedimentation

The lamella sedimentation technique offers great advantage in reduction of the operation volume, stable and favorable hydraulic conditions, and simple sludge handling.

The removal efficiency in the two different types of sedimentation units — lamella and Lovö basin — is of course dependent on the actual load. In the investigations, the spread in the results obtained was relatively large, but it is possible to state that the two types of sedimentation units tested in full scale at the Water Treatment Plant at Lackarebäck have approximately the same removal efficiency at a surface loading of 2 - 3 m/h for the Lovö basin, and 18 - 19 m/h for the lamella sedimentation unit.

There is a great difference in sludge handling in the two units. The Lovö basin is emptied every 3rd to 4th week, whereas the sludge is continuously and automatically removed from the lamella unit. Penetrating studies of the sludge dewatering and sludge handling have not been carried out, but some tests of the sludge concentration may be mentioned. At the Lackarebäck plant the dry solid content in the sludge from the lamella sedimentation unit was about 3.5 g per liter. The sludge withdrawal corresponds to a maximum water loss of 0.5 %, a value which can be regarded as the maximum sludge discharge necessary. In the lamella sedimentation units there are possibilities of a sludge dewatering in the sludge container, which might further decrease the water loss and simplify additional treatment of the sludge. The water loss at the sludge discharge from a Lovö basin can be estimated at minimum 0.5 %.

The water quality in the lamella sedimentation basin ought to be superior to that of the conventional basin in which the sludge is stored for some time, causing taste and odor problems. Further details and results are shown in Appendix 5 and 6.

6 FILTRATION, BACKGROUND

6.1 General

Filtration through granular media is an important unit operation for solid-liquid separation in water treatment, which has been in practical use for a very long time. It was, however, not until the last decade that the practical and theoretical development of the operation intensified, and new filter designs – different from the conventional filters with a filter material of sand of about one meter in depth, with a grain size of about 0.5 to 1.0 mm, and a filtration rate of about 5 m/h – have been more extensively used.

The development within the filtration technique has moved towards the use of higher and higher filtration rates, and it is possible to maintain the filtrate quality on a satisfactory level even at these higher filtration rates. At high filtration rates, however, the filtration period is short, unless the filters have been designed to counteract this fact. In order to maintain satisfactory filtration periods, one may use the following filter designs (a) coarser filter media, (b) two-media filters, filters with two layers of material of different grain size and density, (c) multimedia filters, the same design as in (b) but several layers of material, (d) up-flow filters in which the suspension is filtered in an upward direction through a sand bed. The design of this unit operation is regarded, in spite of considerable development work, as an "art" rather than a rational engineering procedure. The difficulties in formulating mathematical models for this unit operation are due to the complexity of the mechanisms affecting the filtration operation. For a complete understanding of the removal ability of a filter, knowledge of the different factors affecting the filtration process is necessary. The suspended particles which are to be removed by a filter vary both in properties and origin, variations which may be due to seasonal changes. The suspended solids consist of alumhydroxide flocs, clay particles, bacteria, and algæ .

Even though the particles vary considerably, the variations in filter design are relatively small.

6.2 Optimum design of a filter

Usually there are three conditions to be considered in the design of a filter.

1. Water quality
2. Head loss
3. Time of filtration

The water quality must meet the standards set up by the authorities and thus the head loss must not exceed a certain value depending on characteristics of the particles to be filtered. Practical and economic conditions must be considered when determining the time of filtration. At smaller water treatment plants the practical considerations are more important than the economic ones, but at larger treatment plants, where the filters normally are equipped with automatic control panels, the economic significance of the filtration time is more important. In FIG. 6-1 the principle of the filtration operation is illustrated.

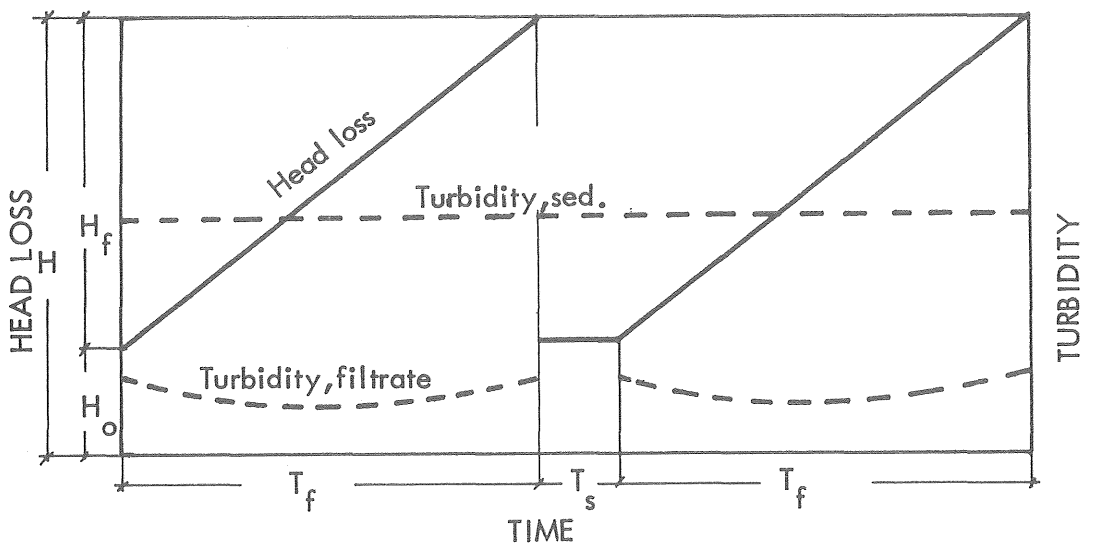


FIG. 6-1 Sketch of the principle of the filter operation

The optimum function of a filter is obtained when the head loss H_f reaches a limiting value $H_{f\ max}$ at the same time as the water quality C reaches the lowest acceptable value (highest permissible turbidity) in accordance with the quality standard. If a break-through occurs before a maximum head loss has been reached, the filter is said to be under-designed for the operational conditions in question. The opposite implies that the filter is over-designed.

By adapting the filter design or by changing the characteristics of the suspended particles and the size distribution of the particles, it is possible to obtain the optimum function of a filter. In accordance with the definitions shown in FIG. 6-1, a general equation concerning the performance of a filter can be written:

$$Q_p = v \cdot T_f - Q_s \tag{6-1}$$

or

$$Q_d = \frac{24}{T_p} \cdot v \cdot T_f - \frac{24}{T_p} \cdot Q_s \tag{6-2}$$

in which Q_p is the production of water in cubic meters per square meter of a filter in one filter period

Q_d is the production of water in cubic meters per square meter of a filter a day

v is the filtration rate (m/h)

T_f is the filtration time (h)

T_p is the filter period (h)

Q_s is the loss of water during washing in cubic meters per square meter of a filter

From the Eq. (6-2) it can be seen that daily production increases with a higher filtration rate as long as the filtration time $T_f >> T_s$. The filtration time T_f is approximately inversely proportional to the filtration rate v. Thus, there is an optimum production in cubic meters per square meter of a filter per day at a filtration rate which is depending on the actual filter design and the backwashing system of the filter. It is therefore important to be able to describe the relationship between the time of filtration and the filtration rate under different conditions for the filters concerned in order to optimize the operation. The operational characteristics of a filter consist, however, of a large number of factors. Each of the following parameters, some of which are interdependent, merits consideration at the design stage of a treatment process and in any assessment of the efficiency of a filtration technique.

Average grain size

Uniform grain coefficient

Filtration rate

Depth of media

Rate of head loss development

Specific capacity of filter bed

Influent and effluent suspended solids
Degree of penetration of arrested material
Run length
Overall throughput capacity
Backwash water consumption as a percentage of throughput

6.3 Principal mechanisms of filtration

In the literature several factors which may play an important role in filtration have been discussed. The predominant mechanisms depend on the physical and chemical characteristics of the suspension and the medium, the rate of filtration and the chemical characteristics of the water. Removal of solids, particularly smaller particles, depends on two types of mechanisms (Ives, 1969).

First, a transport mechanism which brings the small particles from the bulk of the fluid within the interstices close to the surface of the media is required. Transport mechanisms may include gravitational settling, diffusion, interception, and hydrodynamics; these are affected by such physical characteristics as size and density of the particles and the size of the filter medium, filtration rate and temperature.

Secondly, as the particle approaches the surface of the medium, or previously deposited particles on the medium, an attachment mechanism is required to retain the particle. The attachment mechanism may involve electrostatic interactions, chemical bridging, or specific adsorption, all of which are affected by the coagulants and coagulant aids, the chemical characteristics of the water, and the filter medium.

With the guidance of the investigations carried out and described in the literature, examining the relative significance of the different mechanisms, the following summary can be made.

The mechanisms of filtration have been used as a basis for some mathematical models which will be presented later.

TABLE 6-1 Principal mechanisms of filtration

Mechanism	Evaluation
Transport mechanism	
Straining	Significant if the water is difficult to coagulate
Gravitational settling	Significant
Diffusion	Insignificant
Interception	Insignificant
Hydrodynamic (Reynolds No.)	Probably significant
Attachment mechanism	
Electrostatic interaction	Probably significant at small distances < 100 Å
Attractive forces (Van der Waal)	" -

6.4 Filtrate quality pattern

In the search for answers to the fundamental mechanisms affecting the filtration performance and filtration efficiency, a wide variety of patterns of filtrate quality and head loss are observed in filtration investigations and in practice. This variety of patterns is not unexpected, considering the complexity of filtration and the possible changes in transport and attachment mechanisms. Within the filtration poeration two distinct modes of filtration can be observed – surface filtration and deep bed filtration or volume filtration.

6.4.1 Surface filtration

Surface filtration may be characterized by the formation of a cake of suspended particles on the surface of the filter medium, due to blocking of the pores in the uppermost layer of the filter. This is mainly a physical type of removal mechanism – straining – although some small particles may be removed by adhesion to the surface cake. The removal efficiency is mainly constant during the time of filtration. Surface filtration will occur when certain conditions are prevailing with respect to particle characteristics and media size, flow rate, and influent concentrations. Surface filtration may be of significance in cases of waters that are difficult to coagulate such as those with low colloid concentration and low alkalinity.

6.4.2 Volume filtration – deep bed filtration

Volume filtration may be characterized by removal in the entire depth of the filter medium and by different removal mechanisms acting simultaneously. Two different removal efficiency patterns have been observed for the volume filtration.

1. Steadily decreasing removal efficiency
2. Initially increasing efficiency, then constant or decreasing efficiency

These two types of filtration action can be illustrated as in FIG. 6-2 in which the effluent turbidity as a function of time is shown.

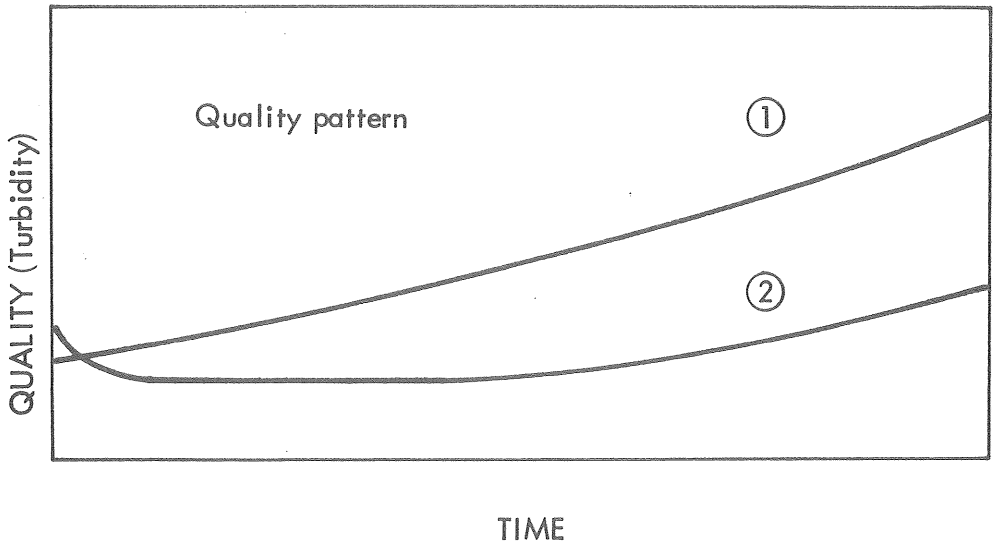


FIG. 6-2 Different quality patterns in volume filtration or deep bed filtration.

A steadily decreasing removal efficiency has been observed with the use of different suspensions – bentonite clay, latex spheres, algae bacteria (Agrawal, 1966); clay particles (Cleasby, 1969); (Ornatski, 1955); synthetic particles (Maroudas, 1965); silica and alumina particles (Herzig *et al.*, 1970); magnesium silicate particles (Diaper and Ives, 1965).

Eliasson (1935); Ling (1955); Heertjes and Lerk (1967); Cleasby (1969) who were using flocculant particles formed by ferric chloride ($FeCl_3$) or ferrous sulfite ($Fe\ SO_4$) observed a steadily decreasing removal efficiency pattern. In explanation of this action, (Wright *et al.*, 1970) have suggested two different hypotheses:

1. The specific surface of the filter media decreases as the particles are deposited in the pores

2. Attractive forces between the suspended particles and the deposited particles are not sufficient to overcome the increasing surface shearing forces

Concerning the first hypothesis, it might be noted that other investigators also consider the change in specific surface to be important (Stein, 1940); (Ling, 1955); (Ives, 1963).

When suspended solids are removed, the surface characteristics of the medium change , and the pore volume will decrease thus increasing the velocity. This latter effect tends to decrease the possibility that a particle will be deposited. Shearing forces also increase, and when the attractive forces are insufficiant, the possibility of removal decreases. The result of several investigations (Agrawal, 1966); (Heertjes and Lerk, 1962);(Alexander, 1963); (Maroudas and Eisenklam, 1965); (Cleasby, 1969), suggests that the monotonic decrease in removal efficiency is due to the magnitude of surface shearing forces in relation to the attractive forces.

<u>Initially increasing efficiency – constant or decreasing efficiency</u> has been observed in several experiments. Agrawal (1966) showed that the removal efficiency would increase with time at higher ionic strengths. Experiments carried out at a high temperature have also given the same result.(Deb, 1969).

Cleasby and Bauman (1962); Ives and Sholji (1965); Ives and Pienvichetr (1965); Agrawal (1966); Davis and Borchardt (1966) and others have all observed an increased removal efficiency at the beginning of the filter period. Different suspensions ranging from discrete particles to flocculant solids were investigated. As the period of the increased efficiency is normally very short in relation to the total filter period, it is of less importance to consider this from an engineering point of view when modeling the filtration operation. However, Ives has developed a filtration theory based on this pattern.

The explanation of this mode of filtration action may be (Wright *et al.*, 1970):

1. Smaller pores may be clogged, causing an initial increase in removal efficiency. When these pores have been filled, the removal efficiency begins to deteriorate.

2. Attractive forces between the suspended solids and the deposited solids are stronger than between the suspended solids and the filter media, and thus the removal efficiency increases when the media is covered. After that the surface shearing forces become predominant, and the efficiency decreases.

3. As suggested by Ives, the specific surface is increased initially, then decreased.

From the investigations carried out and described in literature, the second hypothesis can be regarded as the most relevant.

6.5 Mathematical models for deep bed filters

Several mathematical models for the filtration have been suggested. The most common approach has been to develop a model relating filtrate quality and head loss to time. Some of the relationships have been developed empirically, and the models do not attempt to consider the predominant mechanisms responsible for the observed removals. Other investigators have developed models by considering the basic mechanisms. All physical models of filtration, as originally noted by Iwasaki (1937), are based on the fact that the rate of removal per unit depth of filter is proportional to the local concentration of suspended particles,

$$- \frac{\partial C}{\partial L} = \lambda C \qquad (6\text{-}3)$$

in which C is the particle concentration

L is the length (depth) of the filter

λ is the filter coefficient which varies with time and depth in the filter

At the initial stage of the filtration the following limits are valid: $T_f=0$ and $\lambda=\lambda_o$, and the equation can be written:

$$\frac{C}{C_o} = e^{-\lambda_o L} \tag{6-4}$$

in which C_o is the influent concentration at L=0.

As the filtration operation proceeds, the geometry within the filter is changed due to the specific deposit σ_{ss}. The filter coefficient varies with time and depth in the filter.

The material balance in the filter between deposited particles and influent particles can be expressed approximately:

$$-\frac{\partial C}{\partial L} = \frac{\partial \sigma_{ss}}{v \partial t} \tag{6-5}$$

in which σ_{ss} is the volume of specific deposit per unit filter volume

v is the filtration rate $(\frac{Q}{A})$

From Eqs. (6-4) and (6-5) it can be seen that the filter coefficient is dependent on the initial concentration C_o and on the specific deposit. In order to obtain a complete mathematical model of the filtration operation when the probable removal mechanisms are known, one must find a relationship as follows:

$$\lambda = \lambda_o f\ (\sigma_{ss}) \tag{6-6}$$

The variation of the filter coefficient with specific deposit has been the subject of considerable research with different workers proposing different relationships. Ives (1969) has developed a general relationship between λ and σ_{ss} based on the hypothesis that the filter coefficient is a function of the specific surface and the interstitial velocity.

Ives assumes that the filter bed may be represented by individual cylindrical capillaries, and as the deposits accumulate, the interstitial velocity increases. The general equation describing the filtration action has been expressed as:

$$\frac{\lambda}{\lambda_o} = (1+\frac{\sigma_{ss}}{1-f})^y \ (1-\frac{\sigma_{ss}}{f})^z \ (1-\frac{\sigma_{ss}}{\sigma_u})^x \qquad (6\text{-}7)$$

in which f is the porosity in a clean bed

σ_{ss} is the specific deposit per unit volume of filter

σ_u is the ultimate value of σ_{ss} when the filter is ineffective at the depth

y, z, x are exponents expressing the influence of the different terms

Several other workers have developed mathematical models which are special cases of the general equation developed by Ives (1969).

Ives (1960) $\lambda=\lambda_o+a\ \sigma_{ss} - b\sigma_{ss}^2/(f-\sigma_{ss})$

Ives' equation is then seen to be a special case of the general Eq.(6-7).

Mackrle (1965) $\lambda/\lambda_o=(\{1+\sigma_{ss}/\{1-f\})^y\ (1-\sigma_{ss}/f)^z$

(x=0 in Eq. 6-7)

Heertjes and Lerk (1967)
Shekhtman (1961)
$\lambda/\lambda_o=(1-\sigma_{ss}/f)$ (6-8)
(y=x=0, z=1 in Eq. 6-7)

Maroudas (1967) $\lambda/\lambda_o=(1-\sigma_{ss}/\sigma_u)$
(y=z=0, x=1 in Eq. 6-7)

Mints (1951) $-\frac{\partial C}{\partial L} = \lambda_o\ C- \frac{\alpha\ \sigma_{ss}}{V}$

which can be written

$\lambda/\lambda_o=(1-\sigma_{ss}/\sigma_u)$
(y=z=0, x=1 in Eq. 6-7)

If the above expressions for λ are inserted in Eq. (6-3), the change in filtrate quality with depth and time can be calculated. Equation (6-7) and the simplifications, Eq. (6-8), can be solved, but there are a number of difficulties in the use of the equations. The variables in the filtration operation are numerous, and a considerable amount of research has also been carried out in order to quantify the filtration models.

6.6 Effect of filtration variables on the filter efficiency

Many different suspensions have been used to describe the filtration operation — clay particles, oxides, Al- and Fe-hydroxides, bacteria, virus, algae, PVC-microspheres, etc. A great number of qualitative remarks on the significance of the different factors such as particle size, media size, filtration rates, temperature, ionic strength, pH-value, coagulants, and coagulant aids have been made by the investigators. However, many statements are conflicting.

Considering the basic mechanisms of removal, Ives (1969) has analyzed and discussed the effect of the filtration variables on filter efficiency at the initial stage (t=0). These mechanisms were interception, diffusion, sedimentation, and hydrodynamics. The removal efficiency Λ in a unit layer with a thickness of one grain has been expressed as:

$$\Lambda = \lambda_o \cdot d = -\frac{\partial C}{\partial L} \cdot \frac{1}{C} \cdot d \qquad (6-9)$$

and the removal efficiency is a function of different mechanisms, which can be expressed as follows:

$$\Lambda = f(\text{interception, diffusion, sedimentation, and hydrodynamics}) \qquad (6-10)$$

or

$$\Lambda = \text{const}\left(\frac{d}{D}\right)^{\alpha} \cdot \left(\frac{kT}{3\pi\mu dvD}\right)^{\beta} \left(\frac{g\{\rho_s-\rho\}d^2}{18\ \mu v}\right)^{\gamma} \cdot \left(\frac{\mu}{\rho\ D\ v}\right)^{\delta} \qquad (6-11)$$

or after rearrangement

$$\Lambda = \text{const}\ \frac{d^{(\alpha-\beta+2\gamma)}}{\mu^{(\beta+\gamma-\delta)} \cdot d^{(\alpha+\beta+\delta)} \cdot v^{(\beta+\gamma+\delta)}} \cdot (kT)^{\beta} \cdot$$

$$\cdot (\rho_\sigma - \rho)^{\gamma} \cdot \rho^{-\delta} \qquad (6-12)$$

in which d is the particle diameter

D is the grain diameter

$\frac{3\pi\mu dvD}{kT}$ is the Peclet number

μ is the viscosity (dynamic)

v is the filtration rate

k is the Boltzmann's constant

T is the absolute temperature

ρ_s is the particle density

ρ is the water density

$\alpha\beta\gamma\delta$ are positive exponents

The equation shows in general an increase in removal efficiency with increased particle size, smaller grain size, lower filtration rates, and increased water temperature.

A quantitative relationship between the important system variables and the rate of deposition is needed in optimization studies. Some investigations have been carried out, but the quantitative remarks are often conflicting.

Concerning the particle size d there are difficulties in measuring and characterizing the suspension. Some synthetic unit suspensions have been used in order to study the importance of the particle size (Agrawal, 1966) and (Diaper and Ives, 1965). Most of the investigators agree that large particles (the size of the particles is though small in relation to the grain size) are easily removed (Eliasson, 1935); (Ling, 1955); and (Stanley, 1955). The predominance of any one mechanism varies with particle size (Herzig *et al.*, 1970). The removal efficiency due to changes in particle size may vary with $d^{-2/3}$ to d^2 (O'Melia, 1972).

When considering the significance of the media size D one can state that most of the workers seem to agree that the removal efficiency is improved in finer media (Baylis, 1937); (Stein, 1940); (Ling, 1955); (Stanley, 1955); (Ives and Sholji, 1965). The removal efficiency may vary with D^{-1} to D^{-3}.

An increased filtration rate implies in general a decreasing removal efficiency. However, there is little agreement on the quantitative form of this dependence. The removal efficiency may vary with v^0 to v^{-4}. A lower viscosity – high temperature – will cause a reduction in surface shear, and in general the efficiencies are higher at high temperatures (Baylis, 1937). Due to temperature changes, the removal efficiency may vary with μ^0 to μ^{-2}. Investigations have been carried out by Ives and Sholji (1965); Deb (1969); and Hunter and Alexander (1963).

The removal efficiency is also due to the chemical characteristics of the suspended particles and the media as well as the water. Several investigations have been carried out in which the importance of the pH-value, ionic strength, ionic species has been studied, (O'Melia and Stumm, 1967); (O'Melia and Crapps, 1964); (Stanley, 1955); (Hunter and Alexander, 1963); and (Gregory, 1967). However, no empirical or theoretical relationships have been suggested to account quantatively for these effects.

The different models suggested in Eq. (6-8) together with the conflicting correlations between the filter coefficient λ_0 and the system variables give an idea of the difficulties arising when agreement between theory and practice is to be established.

6.7 Consequences of the filtration theory on the filter design

According to Eq. (6-3) the rate at which the suspension concentration diminishes with respect to distance is proportional to the local concentration in the filter. In a uniform filter the reduction in concentration will be logarithmic with distance. A consequence of this fact is that in a uniform filter, layers of media farther from the surface remove less suspended particles. In practice, where the washing stratifies the size-graded media normally used, each succeeding layer is less efficient, thus removing a smaller proportion of the suspension in accordance with the theory. It follows that for a filter to give the highest possible overall throughput rate, an even deposition of material must be achieved throughout the full depth of the media, so that local development of head loss is avoided. As suggested by Ives (1969) this can be achieved by making each succeeding layer more efficient so that each takes out a greater proportion of the suspended matter flowing into it. Increased surface area of grains should lead to increased filter efficiency for removal of particles. The surface area of grains per unit filter bed volume S_0 is given by (Ives, 1969)

$$S_o = 6\ (\frac{1-f}{\psi D}) \qquad (6\text{-}13$$

in which f is the filter porosity
D is the grain size
ψ is the sphericity ($\leq$ 1.0)

The equation shows that the specific area is increased with the use of a more angular material but as most angular materials pack to a more open porosity there is little advantage in varying the angularity of the grains.

The best way to achieve increased surface area per layer of filter is to grade the filter media coarse to fine in the flow direction. Upward flow filtration is a logical development since theoretically an ideally graded bed results from the natural distribution of uniform density grains after the upflow backwashing. Dual media and multi-media beds only partially meet the requirements and are subject to mixing at the interfaces during back washing.

6.8 Head loss

The head loss h per unit of depth of a filter can be approximated very closely by a straight line in the following form (Deb, 1962)

$$h = h_o + k\ \sigma_{ss} \qquad (6\text{-}14)$$

in which h_o is the initial (clean filter) hydraulic gradient
k is a dimensionless constant
σ_{ss} is the specific deposit

The total head loss H to any depth L can be obtained by integration

$$H = \int_o^L h\ dL = \int_o^L h_o\ dL + \int_o^L k\ \sigma_{ss}\ dL \qquad (6\text{-}15)$$

If the filter material is uniform

$\int_o^L h_o\ dL = h_o \cdot L = H_o$ and k will be constant for all values of L

The initial head loss H_o can be calculated by the Kozeny-Carman equation at laminar flow through porous media

$$H_o = K \frac{\mu \, v \, S_o^2}{\rho \, g \, f^3} \qquad (6\text{-}16)$$

in which K is the Kozeny-Carman constant
μ is the viscosity (dynamic)
S_o is the specific surface of clean filter media
ρ is the density of water
f is the porosity of filter media
v is the filtration rate

The total head loss through a filter can thus be written:

$$H = H_o + \int_0^L k \cdot \sigma_{ss} \, dL \qquad (6\text{-}17)$$

If the main part of the suspended particles is removed, the expression in Eq. (6-17) can be approximated to a linear relationship between the total head loss and the filter run time (Ives, 1963),

$$H = H_o + k \, v \, C_o \, T_f \qquad (6\text{-}18)$$

in which H is the total head loss in the filter
H_o is the initial head loss
k is a coefficient
v is the filtration rate
C_o is the concentration of suspended solids in the influent
T_f is the filter run time

This relationship (Eq. 6-18) was reported on the basis of empirical observations by Mints (1966).

If, in addition to filtration, straining takes place at the inlet face of the media or at the interface in dual media, additional head loss is incurred by the surface mat of deposition. This causes a logarithmic rise of head loss with time. It is an undesirable effect and can be obviated by either increasing the filtration rate (Cleasby and Baumann, 1962) or changing the filter design. The mathematical form of such an extra increase in head loss H_e, due to an extreme surface mat deposition can be written (Boucher, 1947):

$$H_e = k_1 e^{k_2 t} \tag{6-19}$$

in which k_1 and k_2 are coefficients

From an optimization point of view Eq. (6-18) gives a convenient expression for the slope $k\, v\, C_0$ of the linear relationship. It is thus possible to derive an expression of the coefficient k with respect to different conditions for a filter. The coefficient k is assumed to be a function of different variables:

$$k = f\text{ (particle characteristics, temperature)} \tag{6-20}$$

It has been reported (Ives, 1969) that the filter theory is useful in optimization studies as the filtrate quality can be continuously calculated at different head losses. The definition of an acceptable quality of water for public supply is, however, a task of great complexity.

The quality standards are often diffuse, especially concerning turbidity in water. Water quality goals adopted in 1969 by the American Water Works Association call for a finished water turbidity not exceeding 0.1 JTU. Normally, no attention has been paid to the changes in the distribution system. Thus, in order to meet the standards, it is supposed sufficient to determine a filtrate turbidity which is barely acceptable and to determine a maximum head loss which corresponds to the water quality limit. The maximum permissible head loss is normally not a subject of any thorough consideration in the design of filter operations but is chosen between one and two meters of water.

A somewhat more refined relationship ought to be sought in this investigation as the maximum permissible head loss $H_{f\,max}$ is dependent on several factors which may be expressed as follows:

$$H_{f\ max} = f\ (\text{temperature, particle characteristics, particle concentration, and filter design}) \quad (6\text{-}21)$$

The relationship expressed in Eqs. (6-18), (6-19), (6-20), and (6-21) can be combined with Eq. (6-2) which gives a basis for calculation of the amount of water produced per day under varying conditions during the year.

7 FILTRATION, EXPERIMENTAL INVESTIGATION

7.1 Objective of the experimental investigation

The objective of the investigation has been to quantify the significance of the different factors affecting the filtration operation in order to be able to mathematically describe the filter operation. The studies have mainly been focused on the change in head loss during the filter run time under varying conditions. Studies of the filtrate quality have been carried out in order to approximately calculate the effluent quality at one particular head loss or time and not in order to describe the particle removal as a function of time. This latter objective has been regarded as secondary as it does not affect the economic calculations.

7.2 Scope of the investigation

The following variables have been studied:

- Filter design
- Filtration rate
- Suspension characteristics
- Concentration of suspended solids
- Water temperature

The experiments have to a large extent been carried out comparatively with different filter designs, thus excluding some errors due to seasonal changes in the raw water. The filtration studies have been carried out in pilot plants of different sizes. The maximum flow of water through the pilot filter was 7.2 m^3/h. The pilot plants are described in detail in Appendix 4 (Pilot plant No. 1 and No. 4). Some results have been reported earlier (Hedberg 1969, 1974), and with these results as a basis together with complementary tests, the significance of the different filtration variables have been illustrated.

7.2.1 Filter design

FIGURE 7-1 shows the filter designs that have been investigated.

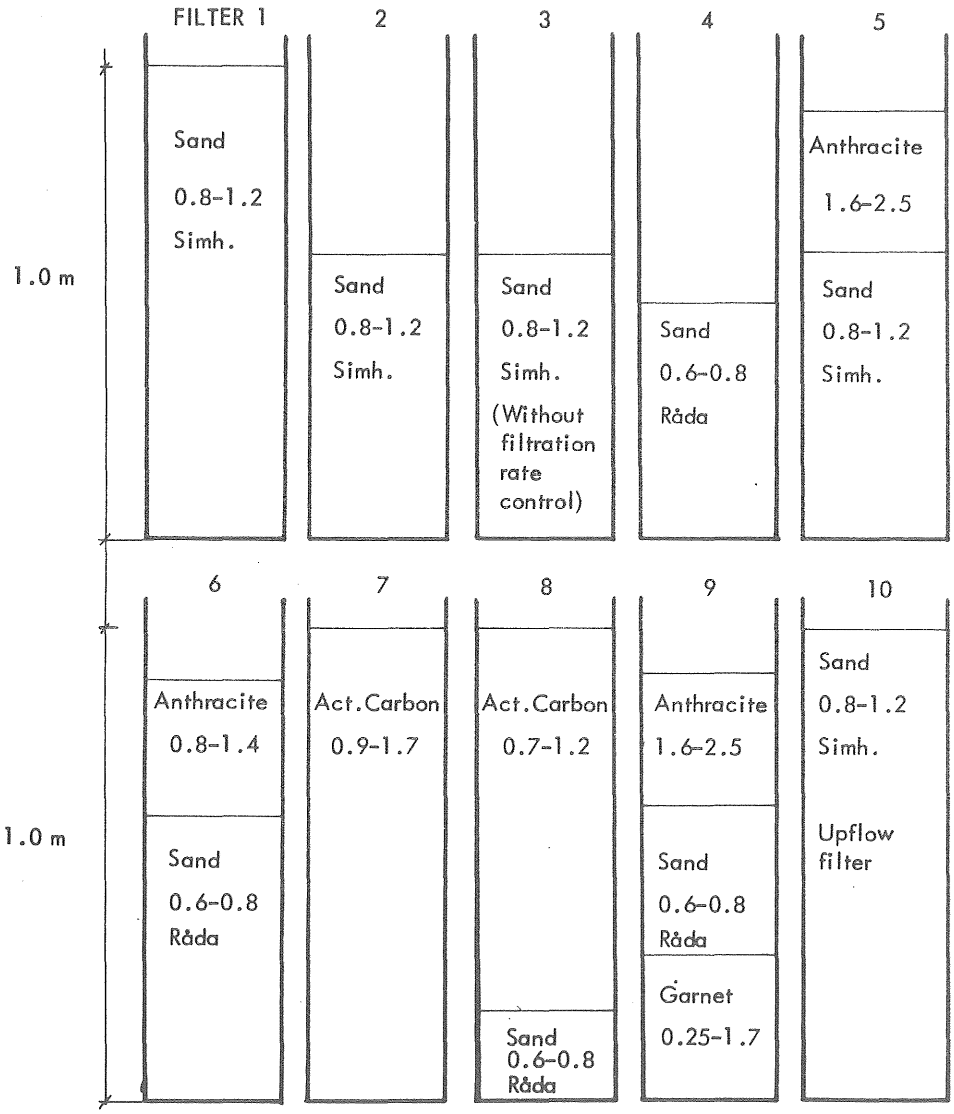


FIG. 7-1 Filter designs. Figures indicate grain size in mm

Quality mark:
Simh = Sand from Simrishamn
Råda = Sand from Råda
Anthracite = Hydroanthracite
Act. carbon = Filtrasorb

The different bed characteristics are shown in detail in Appendix 7.

7.2.2 Filtration rate

The filtration rate has varied up to 30 m/h, but filtration rates within the interval of 7-15 m/h have mainly been used. The filtration rate of 10 m/h has been used in all comparative tests.

7.2.3 Characteristics of the suspension

Different characteristics of the suspension have been obtained with a change in water characteristics because of changes in temperature and dosage of coagulant aid. The effect of various particle concentrations (measured in turbidity units) has also been studied.

7.2.4 Water temperature

The influence of water temperature on the filtration operation has been studied. In this report the water temperature has been used as a variable instead of the viscosity, because only natural water, which has different biological-chemical characteristics at various temperatures, has been investigated.

7.3 Results

7.3.1 Filter effluent quality patterns

As already discussed, different filter effluent patterns have been noticed in the filter research, and different hypotheses have been suggested to explain the different patterns which are:

1. Successively decreased filter efficiency
2. Initially improved, then constant or decreased filter efficiency

Both of these patterns have been observed in the investigations. The patterns are illustrated by some typical patterns for a conventional filter with a filter bed of 0.60 m of sand (eff. size 0.8 mm) at different filtration rates (FIG. 7-2 to FIG. 7-7).

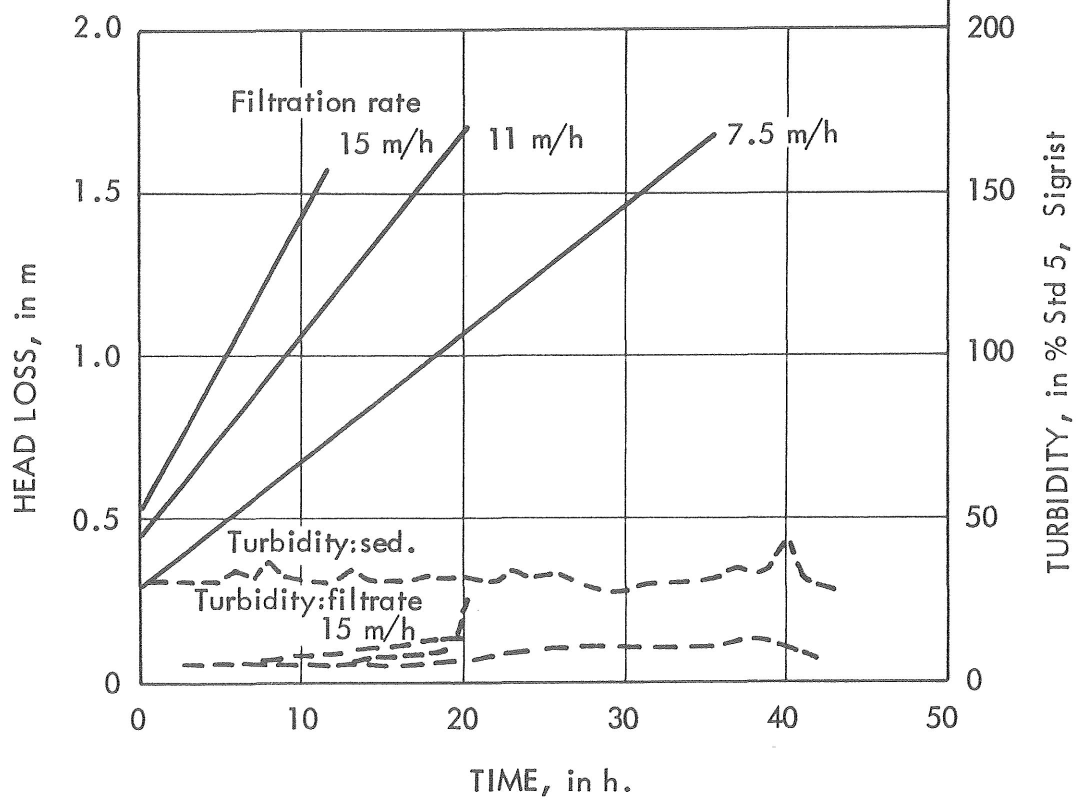


FIG. 7-2 Turbidity and head loss as a function of time. Activated silica 2.5 mg/l, temperature 3°C. No. 720314. Pilot plant No. 4.

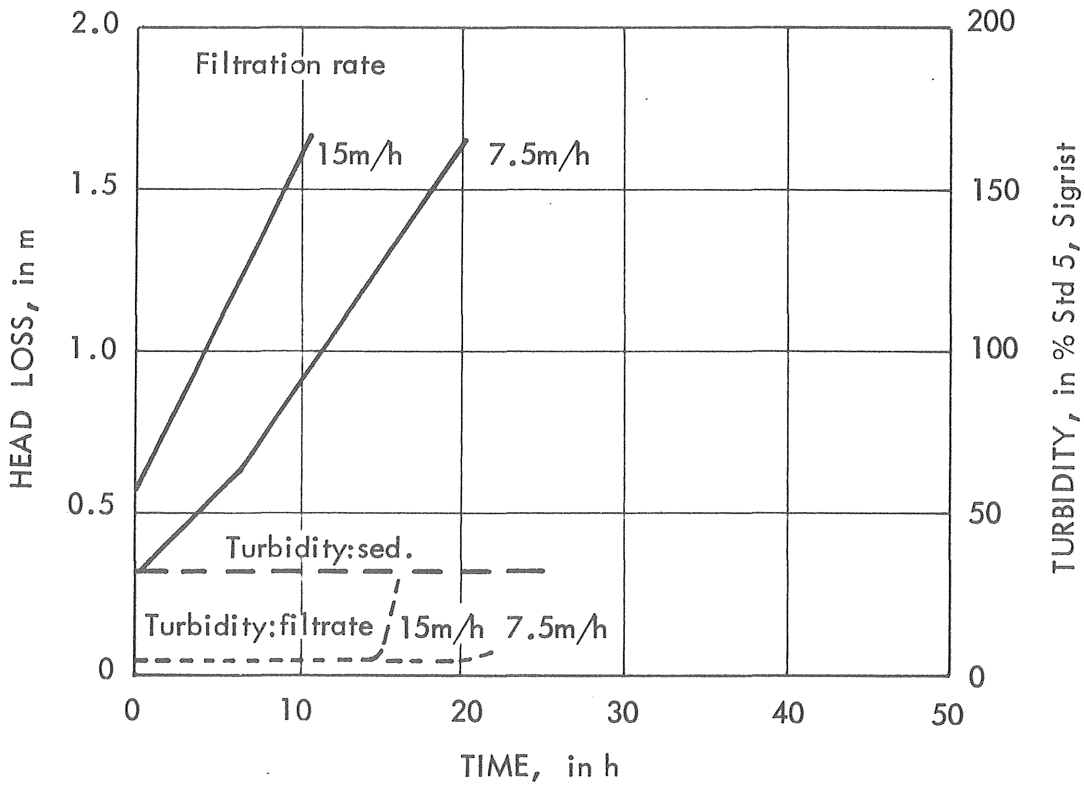


FIG. 7-3 Turbidity and head loss as a function of time. Activated silica 5.0 mg/l, temperature 2°C. No. 720306. Pilot plant No. 4.

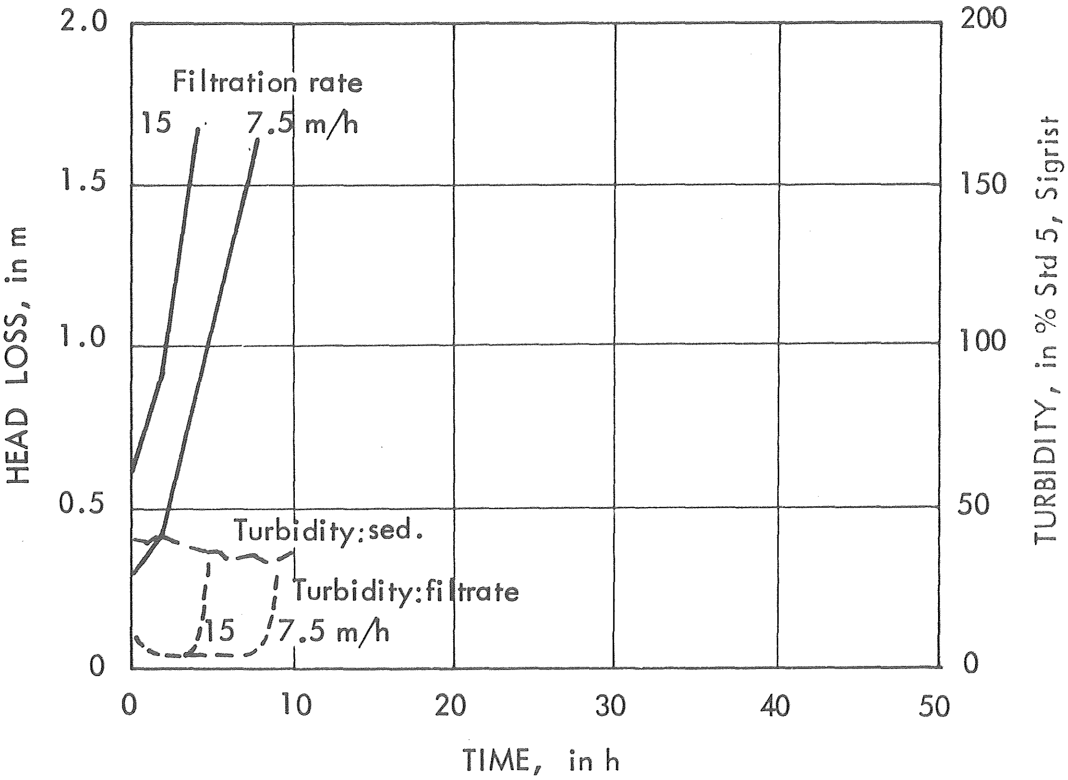


FIG. 7-4 Turbidity and head loss as a function of time. Activated silica 10.0 mg/l, temperature 3°C. No. 720313. Pilot plant No. 4.

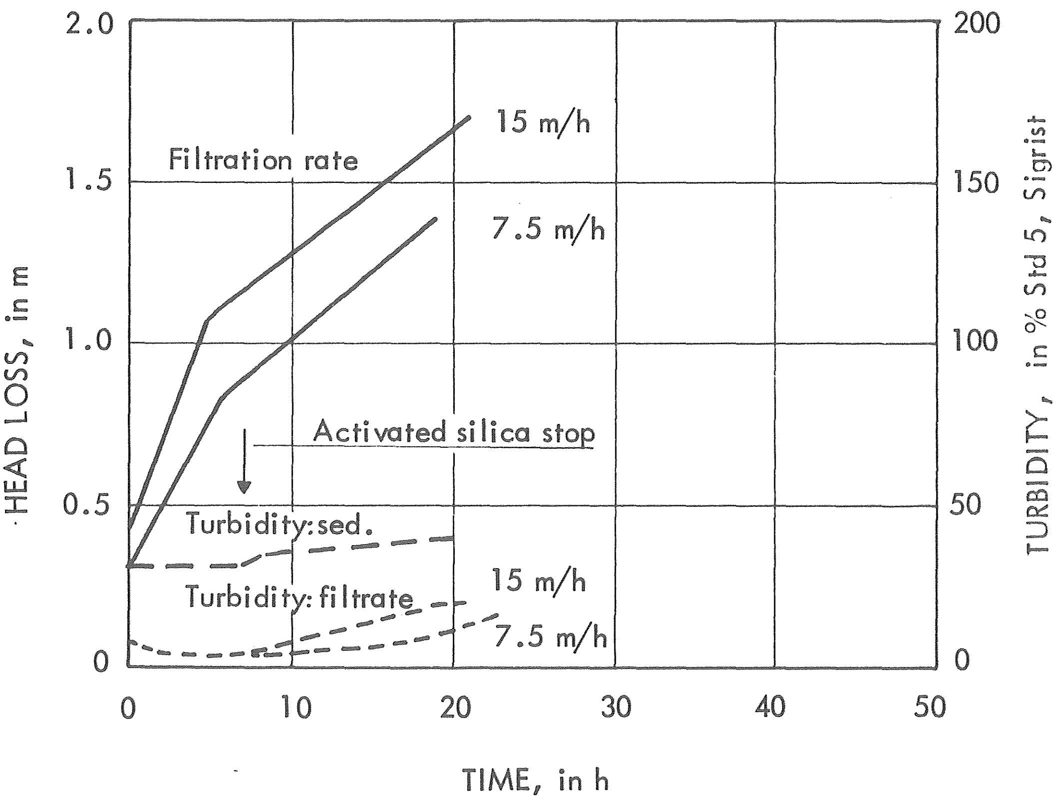


FIG. 7-5 Turbidity and head loss as a function of time. Activated silica 5.0 mg/l, temperature 2.5°C. No. 720307. Pilot plant No. 4.

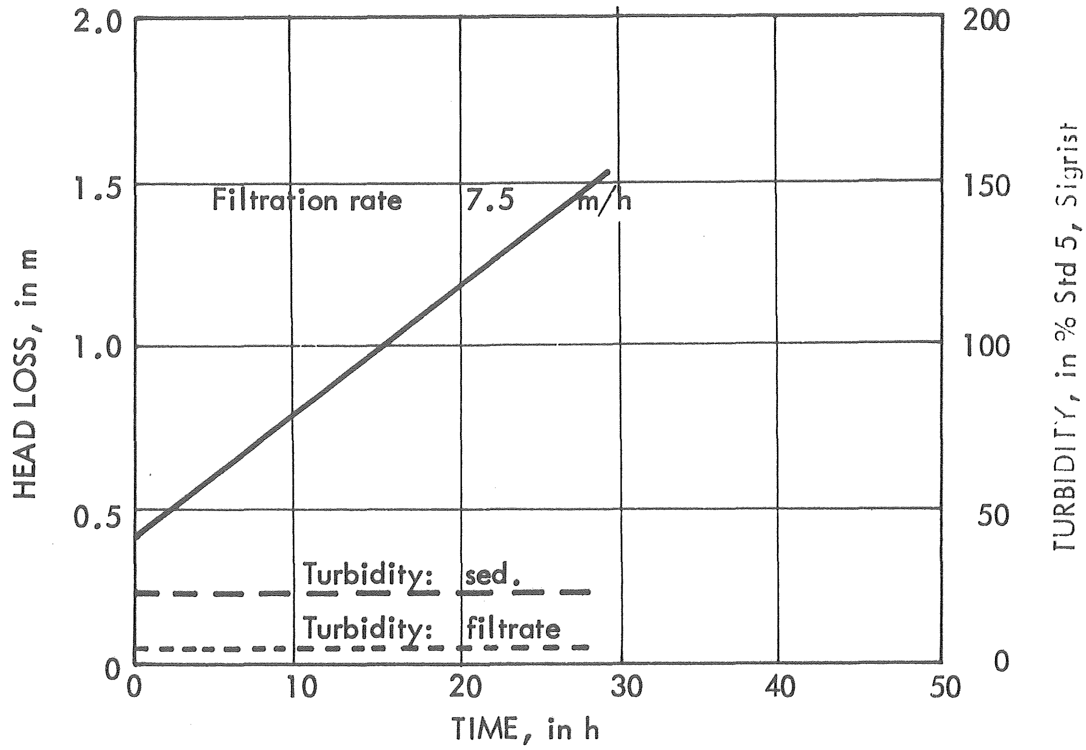


FIG. 7-6 Turbidity and head loss as a function of time. Activated silica 0 mg/l, temperature 18^{o}C. No. 720822. Pilot plant No. 4.

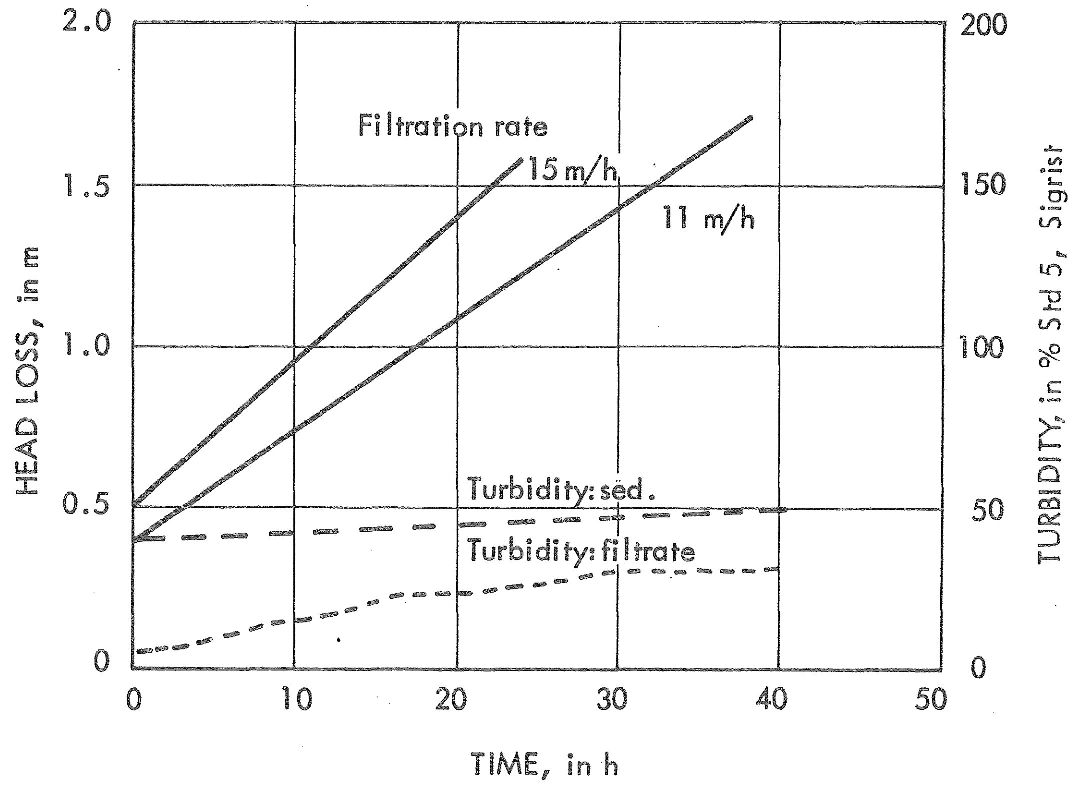


FIG. 7-7 Turbidity and head loss as a funtion of time. Superfloc 0.05 mg/l, temperature 2.5^{o}C. No. 720223. Pilot plant No. 4.

As seen in the figures, the filter efficiency or filtrate quality pattern is affected by the type and amount of coagulant aid added as well as by the temperature. Breakthrough – which means that the turbidity exceeds the quality limit – occurs at low head loss, at low temperature, or at an inappropriate or low dosage of coagulant aid. These tests support the hypothesis that the attractive forces between the deposited material and the suspended particles play an important role and are affected by the amount of coagulant aid. When the head loss increases, the shear stress is also increased and the removal efficiency decreases. Normal values of the head loss at breakthrough varied for the different filter designs. This will be discussed in Section 7.3.5.

The increase of head loss with time did, of course, differ with the different filter designs. A relationship was also observed between the increase in head loss and the maximum head loss permitted due to quality. The lower the rate of increase in head loss, the lower the maximum permissible head loss.

7.3.2 Characteristics of the suspension

Suspension characteristics vary with dosage of coagulant aid and change in water temperature. In connection with the flocculation studies it could be observed that the filtration characteristics of the particles were dependent on the energy input – velocity gradient – in the flocculation operation. However, this observation has not been quantified, and in order to obtain this quantification, one has to determine at least the particle size and shape. A turbidity measurement alone is not sufficient. One typical example of the significance of activated silica addition for the filtrate quality for a conventional filter is shown in FIG. 7-8.

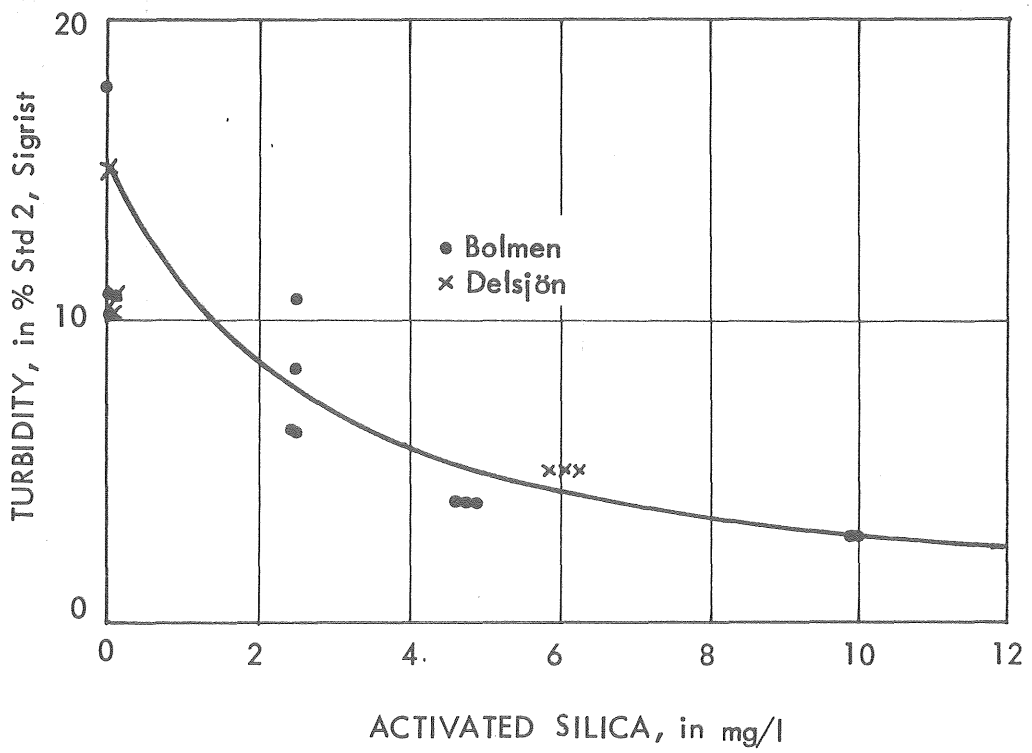


FIG. 7-8 Filtrate quality as a function of activated silica dosage. Filter design: 1.0 m sand (Filter No. 1). Filtration rate 10 m/h. Temperature 5°C. Head loss 1.5 m.

The relationship between filtrate quality C and addition of activated silica at a certain temperature and particle concentration C_0 in the influent can be expressed as:

$$C = \frac{a}{b + d \cdot A} \qquad (7\text{-}3)$$

in which a, b, d are coefficients dependent on the temperature and the particle concentration C_0

A is the addition of coagulant aid

It must be pointed out that the conditions during the preparation of the activated silica, as well as the point of addition and the mixing technique, is of fundamental importance to the result. The mechanism of the activated silica affecting the floc formation and aggregation has not been investigated. It has only been observed that the filtration characteristics of the particles are changed if the addition point is changed. Normally, activated silica is added to the water in the first flocculation tank a few minutes after the addition of alum.

In general, a lower turbidity in filtrate is obtained at higher temperatures, and the following relationship between the turbidity C in the filtrate and the temperature has been observed:

$$C = \frac{a_1}{b_1 + d_1 \cdot t} \tag{7-4}$$

in which a_1 b_1 d_1 are coefficients dependent on the dosage of coagulant aid and the particle concentration C_0

t is the water temperature

The variation of the concentration C in the filtrate is difficult to describe as temperature, activated silica addition, particle concentration, and filtration rate affect the result. If no activated silica is added, the filtrate quality is observed to be approximately proportional to the water viscosity. It is possible to counteract an unsatisfactory filtrate quality at a low temperature by adding activated silica. The equation describing the cooperative effect of temperature, activated silica, and particle concentration is given in Section 7.3.4

7.3.3 Filtration rate

It is a common opinion in the literature that the filtrate quality deteriorates with increasing filtration rate. In the experiments carried out it has been shown that the suspension characteristics and the water temperature are of importance for the change in filtrate quality as the filtration rate increases, (FIG. 7-9).

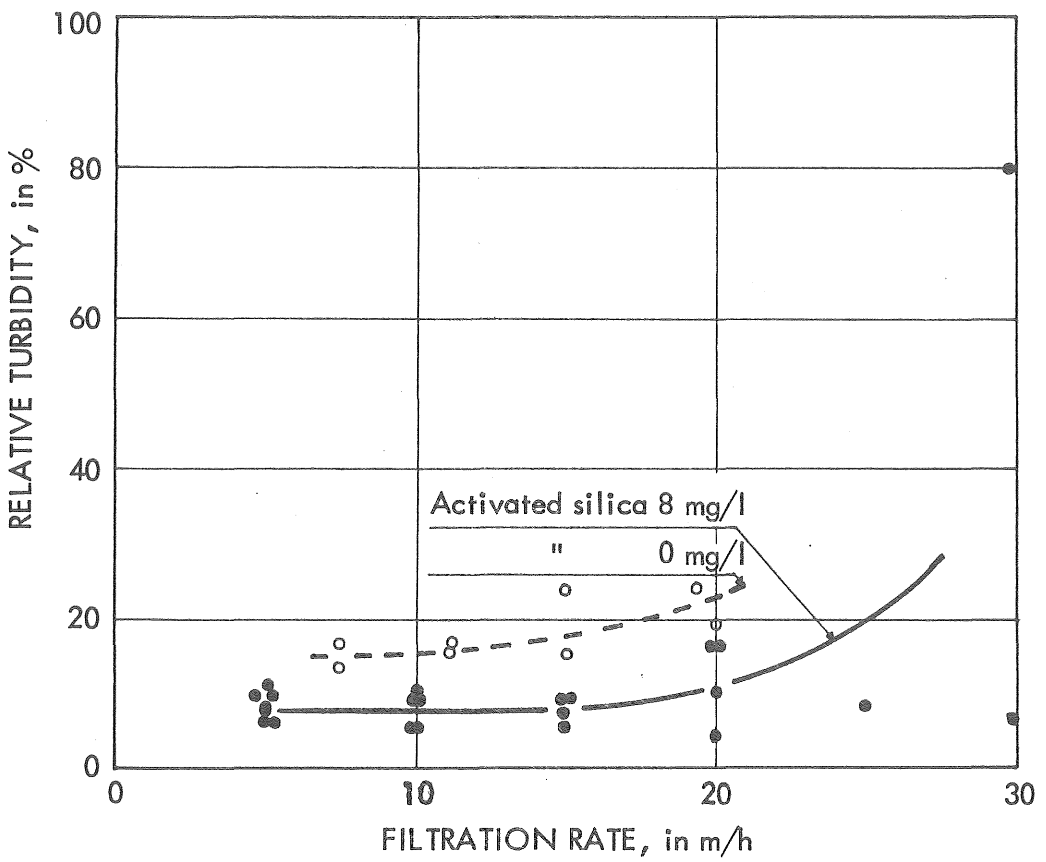


FIG. 7-9 Filtrate quality in turbidity units as a function of filtration rate. Sand filter 1. Pilot plant No. 1. Temperature 8-10^{o}C.

It can be seen that depending on the amount of activated silica added, the filtration rate is of importance only if it exceeds a certain value. If activated silica is added, filtration rates up to between 15 and 20 m/h may be used without any marked deterioration in the filtrate quality. Even higher filtration rates may be used under favorable conditions. Thus, water quality is not affected by the filtration rate in the water treatment plants, as the filtration rates normally are lower than 10 m/h. It is possible to formulate a mathematical expression of the quality deterioration due to the filtration rate. The factor of deterioration f_v may be written:

$$f_v = e^{av^b} \qquad (7\text{-}5)$$

in which f_v is the deterioration factor of the filtrate quality
v is the filtration rate
a, b are coefficients dependent on the amount of activated silica added, temperature, and filter design

Normally the filtration rate must be kept lower in an upflow filter because of the lifting forces, but in some designs a system of grids prevents the bed from lifting.

7.3.4 Particle concentration

In the optimization of the filtration operation, it is interesting to know how the filtrate quality changes with particle concentration of the influent. The particle concentration in the influent is changing due to increased or decreased residence time in the flocculation unit, or change in the overflow rate in the sedimentation unit. The value of the particle concentration may be exactly the same at a short flocculation time and low overflow rate in the sedimentation unit, as at a long flocculation time and a high overflow rate, but the particle characteristics may be different, which influences the filtration in different ways. Here it is assumed that the particle concentration measured in turbidity units can be used unless the origin of the particle is considered. This may be an object for further investigations. The approximation thus done may not affect the result unless an extreme situation is to be predicted. The filter efficiency F_{eff} can generally be expressed as:

$$F_{eff} = \frac{C_o - C}{C_o} \qquad (7\text{-}6)$$

in which C_o is the influent concentration
C is the effluent concentration

If F_{eff} is a constant for varying C_o, the filtrate quality deteriorates with increasing C_0-values. This is in agreement with Eq. (6-4) if the filter coefficient λ is constant. This has been observed to be the case when activated silica is not added. When activated silica is added, the filtrate quality is more independent of the influent concentration. FIGURE 7-10 illustrates this relationship.

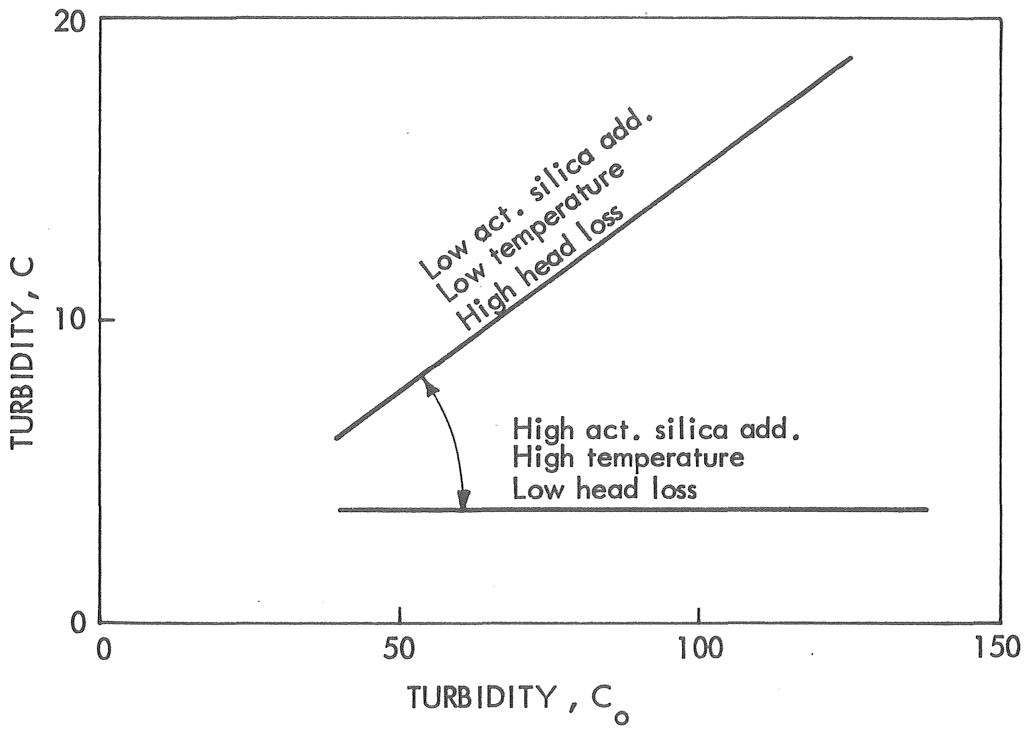


FIG. 7-10 Picture of the principle of the relationship between the filtrate quality and influent quality under varying conditions.

Results from experiments with various influent concentrations and activated silica dosages are shown in FIG. 7-11.

From FIG. 7-11 it can be seen that the filtrate turbidity decreases with the silica addition. At a certain activated silica addition it can be calculated that the removal efficiency increases with increased particle concentration. In FIG. 7-11 the values shown are valid for a head loss of 1.5 m of water. The tendency toward an increased removal efficiency at higher concentrations is more marked at lower head loss, i.e. at the initial stage of the filtration.

The filter efficiency may thus be expressed in mathematical form as:

$$F_{eff} = \frac{C_o}{\varepsilon + C_o} \qquad (7\text{-}7)$$

in which ε is a coefficient dependent on the coagulant aid, water temperature, and filter design

Combination of Eq. (7-6) and (7-7) yields:

$$\frac{C_o - C}{C_o} = \frac{C_o}{\varepsilon + C_o} \qquad (7\text{-}8)$$

or

$$C = \frac{C_o}{1 + \frac{C_o}{\varepsilon}}$$

This improved filtrate quality with increasing particle concentration might be due to the fact that the particle size is not constant but varies with the concentration which is measured in turbidity units. This is quite normal because when the overflow rate in the sedimentation unit is increased, larger and larger flocs are to be found in the effluent.

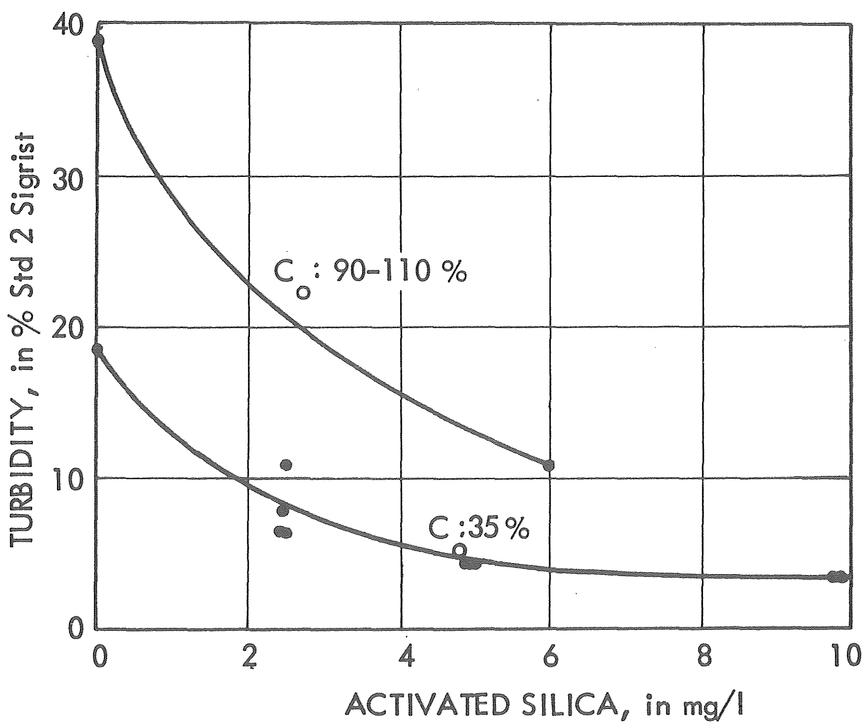


FIG. 7-11 Filtrate quality as a function of activated silica added

7.3.5 Filtrate quality model

It has been shown that several factors affect the filtrate quality which has been measured in terms of turbidity. As it is difficult to measure the turbidity in low concentrations, the degree of accuracy is not sufficient to evaluate any significant differences between the filters before breakthrough. At the present time only one equation describing the filtrate quality has been developed for all filters. As a basis for this equation the Eqs. (7-3), (7-4), (7-5), and (7-8) have been used. The filtrate quality model valid at the beginning of the filtration period can be written:

$$C = \frac{C_o}{1 + \frac{C_o}{a} \quad (1 + b\,A)\ (1 + dt)} \cdot f_v \qquad (7\text{-}9)$$

in which C is the turbidity in the filtrate at t=0
C_o is the turbidity in the influent
a, b, c, are coefficients
A is the activated silica dosage
t is the temperature
f_v is a factor for the filtration rate

Equation (7-9) has been tested and an agreement between the result and a theoretical calculation is shown in FIG. 7-12. The following variables have been studied.

Particle concentration C_o: 30-120 % Std 2 Sigrist.
(Normal values are 40-50 % Std 2)
Temperature: 4-18°C
Activated silica: 0-10 mg/l
The coefficients are: a=15, b=0.08, d=0.02

The experiments have been carried out in pilot plants No. 3 and No. 4.

FIGURE 7-12 shows relatively good agreement between theoretical calculations and results. In this investigation it is enough to approximately calculate the quality as it does not affect the economic calculation. Equation (7-9) gives the quality at t=0, but it is also desirable to calculate the filtrate quality at different head losses.

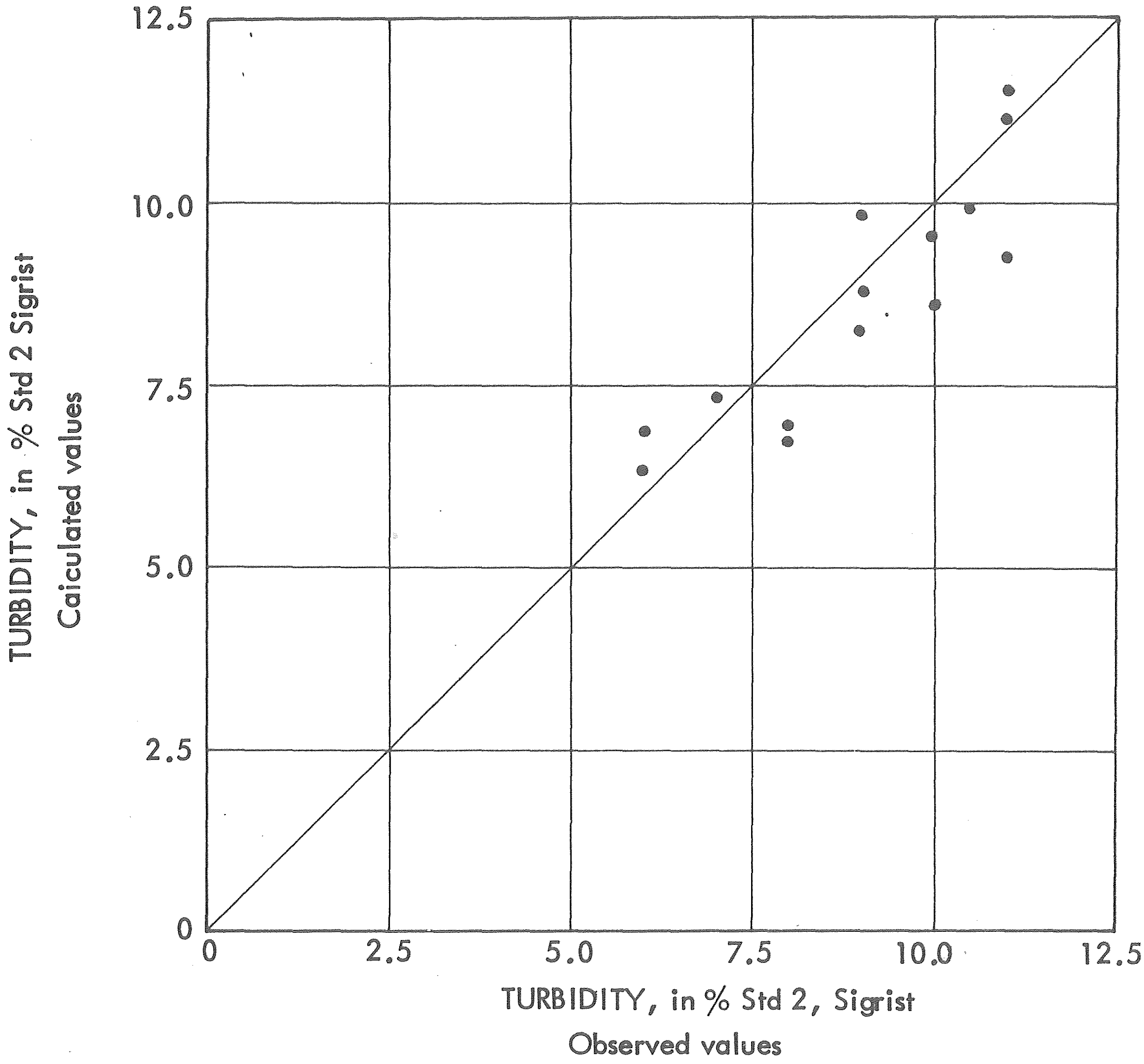


FIG. 7-12 Relationship between theoretical calculations and experimental results of initial turbidity in filtrate for filters at a filtration rate of 10 m/h (correlation coefficient r=0.80)

Since the filter equations already given (Eq. 6-8) are quite different and valid for uniform filters only, it has been considered incorrect to use them. Therefore, in this investigation another method has been used; where at a given filtrate quality C_{crit}, the maximum permissible head loss $H_{f\ max}$ is determined, which in principle has been expressed in Eq. (6-21).

7.3.6 Maximum head loss considering the filtrate quality

With guidance of results obtained and earlier reported results

Eq. (6-21) can be rewritten as:

$$H_{f\ max} = f(F_{type}, A, t, C_o) \qquad (7\text{-}10)$$

in which $H_{f\ max}$ is the maximum permissible head loss at a given filtrate quality C_{crit}

F_{type} is the filter design

A is the coagulant aid dosage

t is the temperature

C_o is the turbidity in the influent to the filter

In practice, the maximum permissible head loss $H_{f\ max}$ is routinely chosen and the value of the head loss in meters of water is normally constant during the year.

In order to determine the Eq. (7-10) for the different filters, one can plot the turbidity versus the head loss caused by the deposited particles. In FIG. 7-13 some typical curves are shown in order to describe the relationships for the filter designs studied in this investigation.

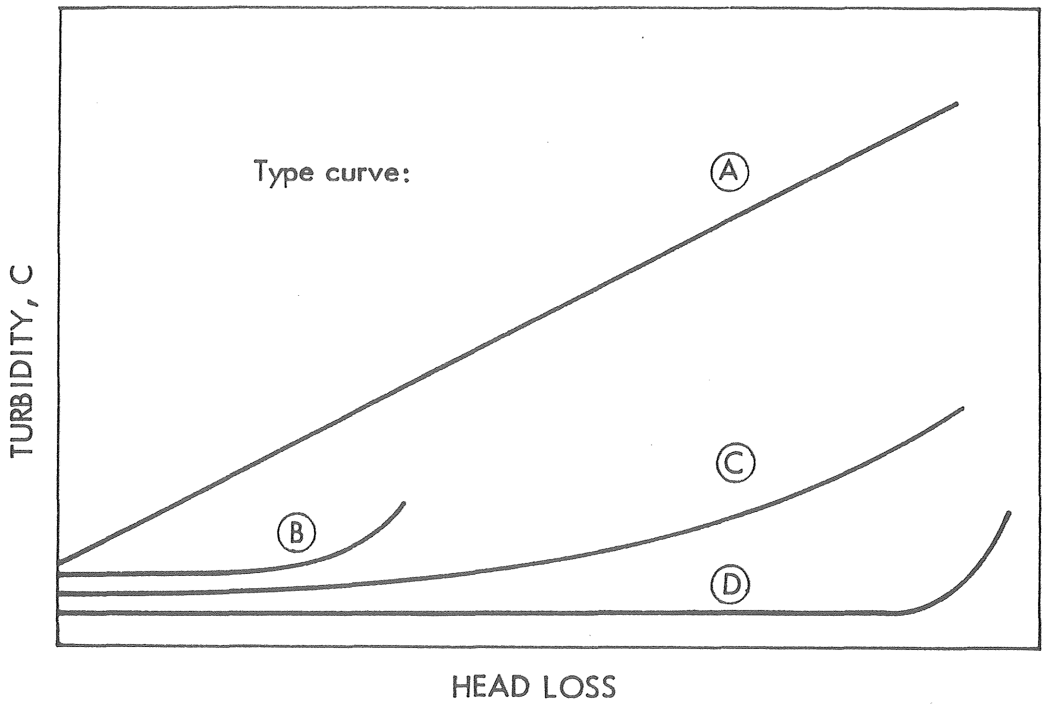


FIG. 7-13 The turbidity in the filtrate as a function of head loss.

In FIG. 7-14 and 7-15 two examples, one conventional filter and one sand-anthracite filter, are shown in accordance with the principle given in FIG. 7-13. Other filters are described by the typical curves illustrated in FIG. 7-13.

Filter 1. Sand filter 100 cm, d_{eff} = 0.87 mm

At a low water temperature and with no addition of activated silica, the filtrate quality deteriorates linearly with the head loss (type A). At higher temperatures and with addition of activated silica, the removal efficiency is constant up to a head loss of more than 2.0 m of water.

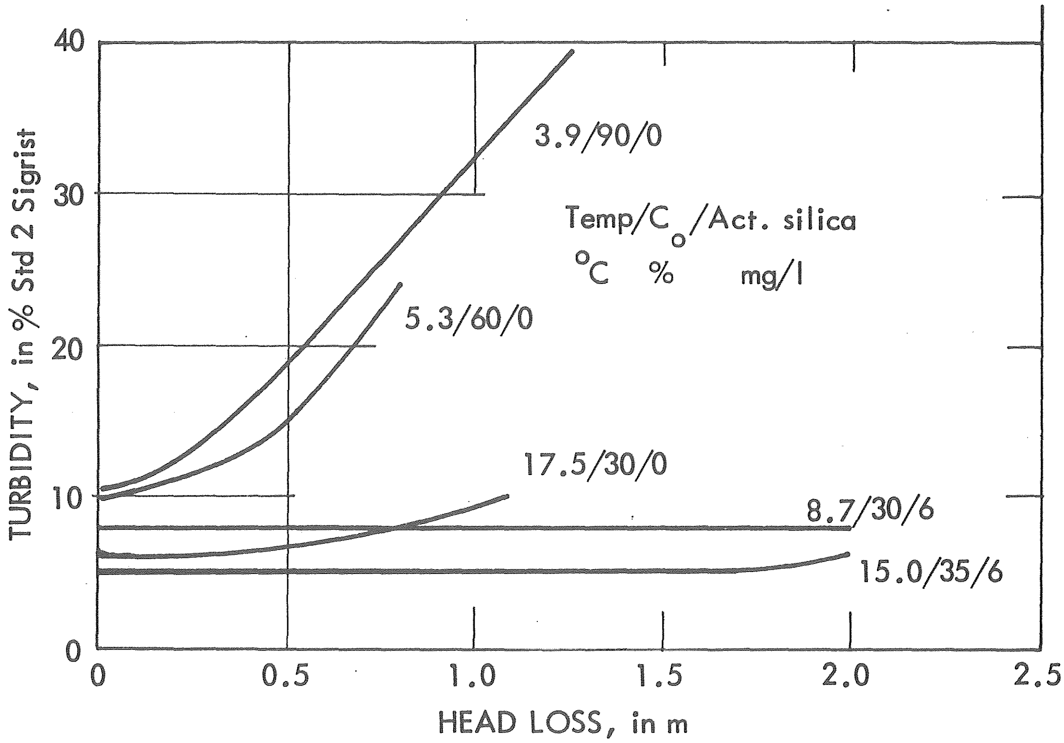


FIG. 7-14 Filtrate quality as a function of head loss. Filter 1. Filtration rate 10 m/h

Filter 2 and 3. Sand filter 60 cm, d_{eff} = 0.87 mm

Similar to filter 1 at normal conditions

Filter 4. Sand filter 60 cm, d_{eff} = 0.65 mm

In principle similar to filter 1 at normal conditions

Filter 5. Sand-anthracite I, 60 cm, d_{eff} = 0.87 mm-
30 cm, d_{eff} = 1.6 mm

In principle similar to sand filter 1 but it has a somewhat higher rate of increase in turbidity at low temperatures. In this type of filter there is a risk that solids, probably smaller particles which have not been intercepted by already removed particles, penetrate deeply into the bed or even through it, causing a deterioration of the filtrate quality. This has been shown by Hedberg (1974).

Filter 6. Sand anthracite II, 60 cm, d_{eff} = 0.65 mm-
30 cm, d_{eff} = 0.95 mm

The removal efficiency is constant at low temperatures up to a head loss of about 0.4 m of water and then rapidly decreases (type B in FIG. 7-13). At other temperatures and with the addition of activated silica, the removal efficiency is constant at least to 1.5 m of water (FIG. 7-15).

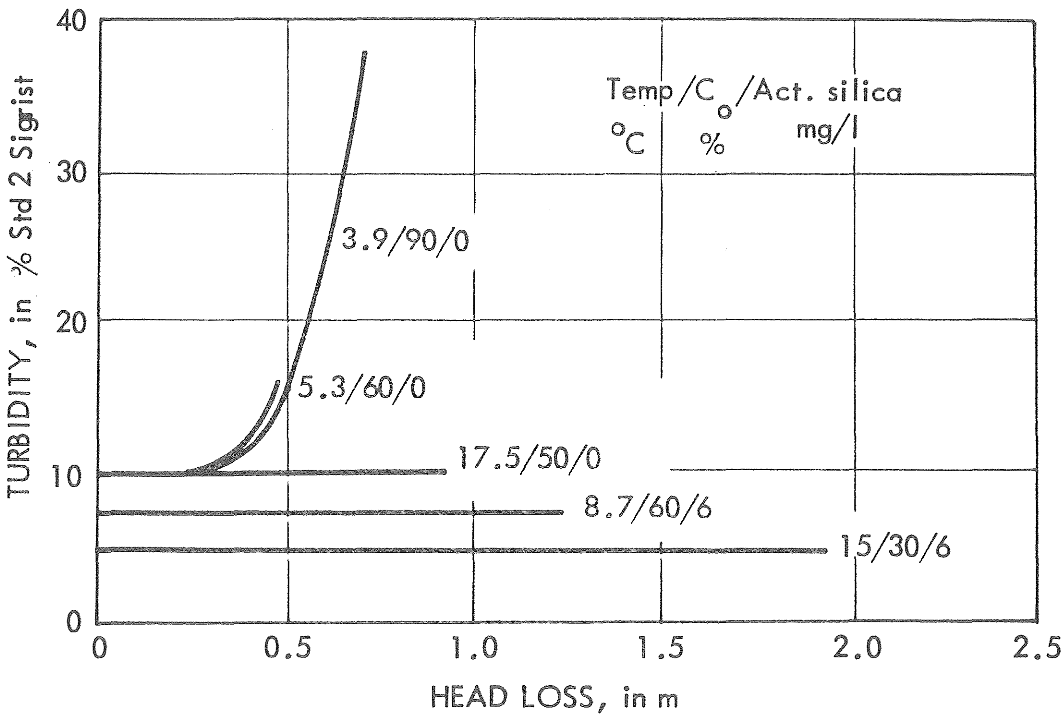


FIG. 7-15 Filtrate quality as a function of head loss.
Filter type Sand-anthracite II
Filtration rate 10 m/h

Filter 7. Activated carbon 100 cm, d_{eff} = 1.0 mm

It is possible to reach higher head losses in this filter than in a sand filter with the same effective grain size.

Filter 8. Sand-activated carbon 20 cm, d_{eff} = 0.65 mm-
80 cm, d_{eff} = 0.67 mm

This filter has a high removal efficiency even at a low temperature and without addition of activated silica up to a head loss of about 0.5-0.6 m of water (type C). This is probably due to the layer of fine sand at the bottom of the filter. The difference in density and grain size of the two media is such that a sharp interface is developed.

Filter 9. Anthracite-sand-garnet 30 cm, d_{eff} = 1.6 mm-
30 cm, d_{eff} = 0.67 mm-
30 cm, d_{eff} = 0.30 mm

In this three-media filter a certain mixing at the interface occurs. This makes the filter very tight, which means that the pore volume is rather low over a relatively long part of the bed. The head loss through the clean filter is higher than those of the other filters. The removal efficiency is high and similar to that of filter 8 – activated carbon-sand filter.

Filter 10. Upflow filter 100 cm, d_{eff} = 0.87 mm

This differs from the other filters, and because of the lifting forces is is more affected by the actual diameter size of the pilot filter than the other down-flow filters. It is doubtful if the results concerning the ability of the filter to resist a breakthrough at a high head loss is relevant to conditions in full scale operations. Tests in larger pilot filters than used in this investigation are necessary. However, several tests have been carried out, and valuable information on the filtrate quality has been obtained. In the pilot plant with a filter diameter of 0.1 m, a total head loss of about 1.5 m of water has been reached without a breakthrough.

In summarizing the different quality-head loss patterns, the breakthrough seems to occur more suddenly and rapidly the longer the removal efficiency stays constant. Also, the "tighter" the filter, the better it resists a breakthrough. It can be concluded that the maximum head loss is dependent on filter design, particle concentration, coagulant aid, and temperature. Before some equations are developed, the porosity of the filters and the head loss distribution in the filter bed should be discussed.

7.3.7 Porosity and head loss in the filter bed

Valuable information as to filter requirements for efficient removal of suspended particles can primarily be obtained by determination of the grain size of the media, the grain size distribution, and the shape of the grains; and secondly, after testing, by determination of the head loss in the filter bed. The media is normally classified by the following parameters:

Effective grain size d_{10} (sieve size on the sieve which 90 % of the media passes)

Nonuniformity $\frac{d_{60}}{d_{10}}$

Shape of grain, Sphericity ψ (ψ = 1 for a sphere)
Porosity f
Pore size f_d
Degree of packing β

In FIG. 7-16 the tightest and the loosest packing of spheric grains with uniform grain size are shown and also an expression of the diameter of the pore f_d.

The porosity for a bed of spheric grains with uniform grain size is 0.38 and 0.40-0.48 for crushed or irregular grains such as anthracite. The pore size in a filter bed is dependent on the grain size and degree of packing. The pore size can be estimated for a bed of uniform grains to be about $0.25 \cdot D$.

In order to complete the picture of the removal efficiency for the different filters, one can study the results from porosity

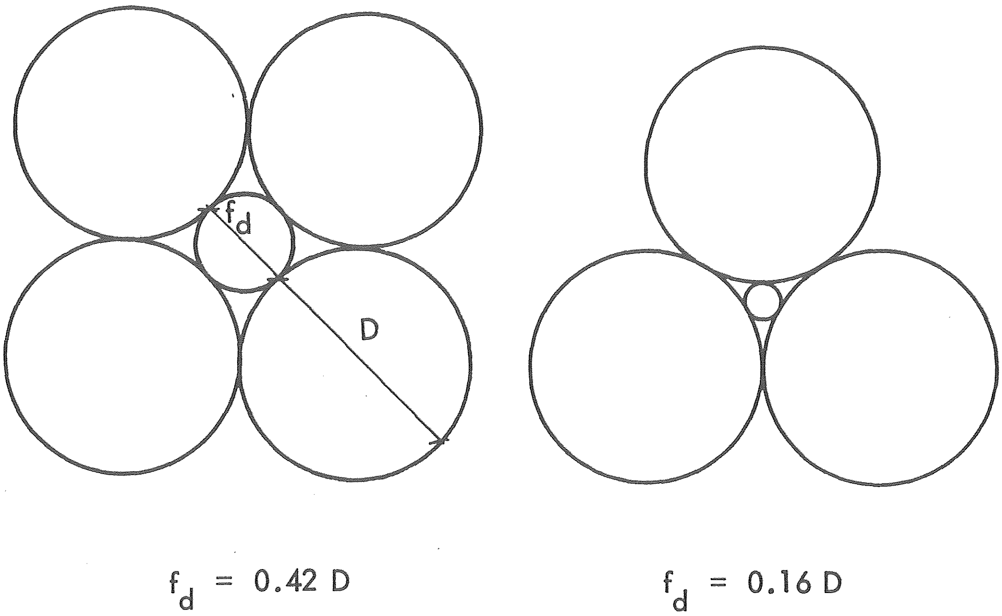


FIG. 7-16 Different packing of spheric grains

measurements and head loss measurements. These results are shown in FIG. 7-17 through 7-23.

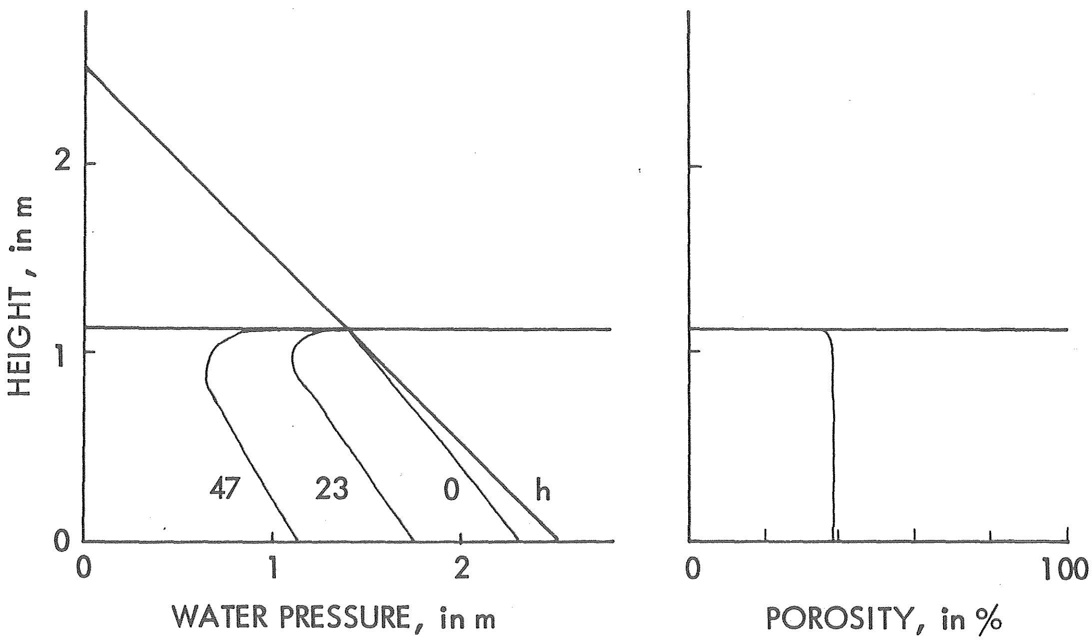


FIG. 7-17 Sand filter, filter 1
Head loss and porosity in the filter bed,
Test 750707, Pilot plant No. 3

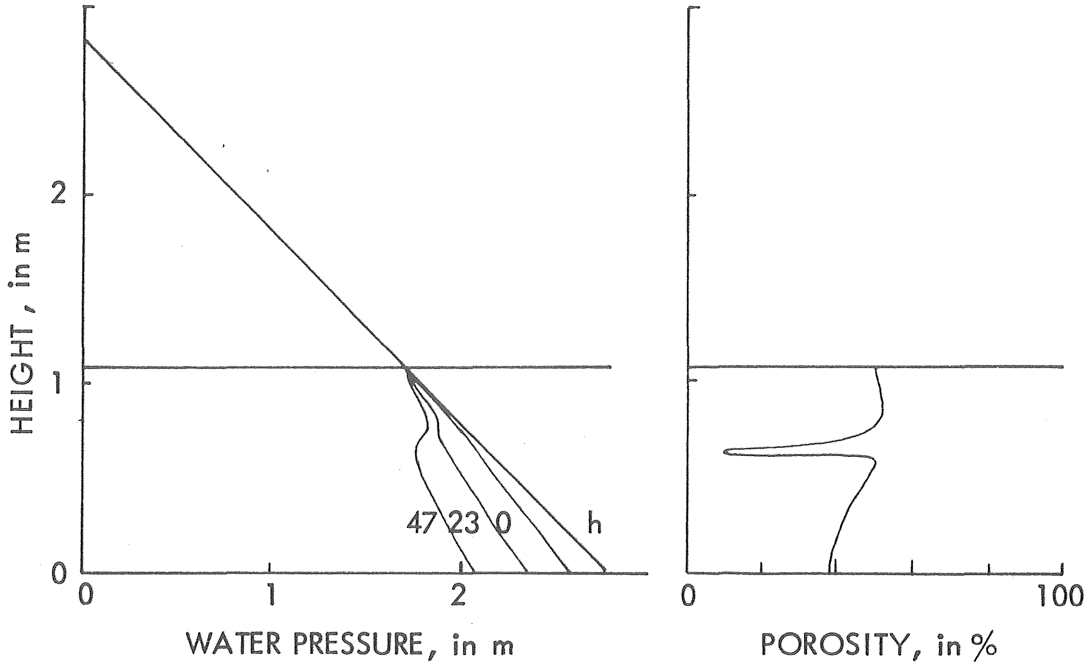


FIG. 7-18 Sand-anthracite I filter, filter 5
Head loss and porosity in the filter bed.
Test 750707, Pilot plant No. 3

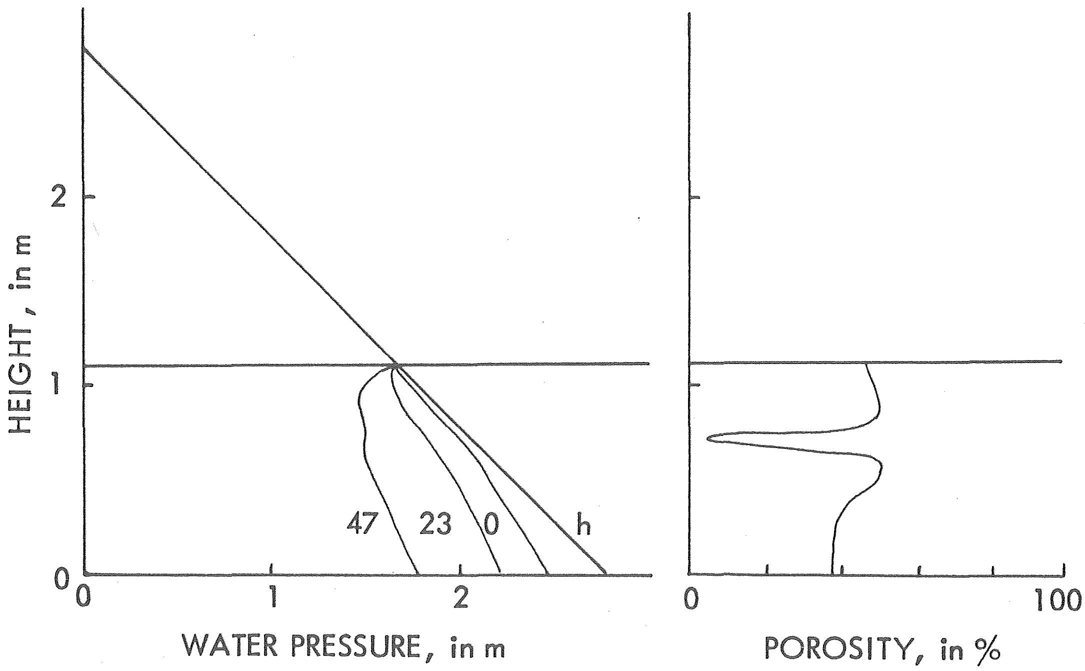


FIG. 7-19 Sand-anthracite II filter, filter 6
Head loss and porosity in the filter bed.
Test 750707, Pilot plant No. 3

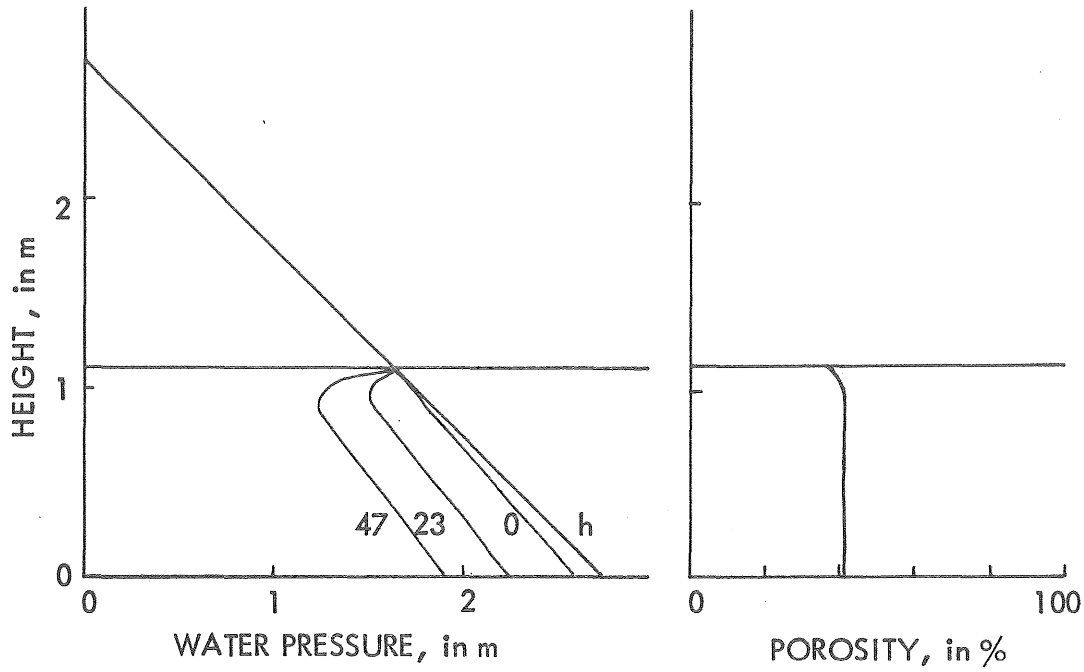


FIG. 7-20 Activated carbon filter, filter 7.
Head loss and porosity in the filter bed.
Test 750707, Pilot plant No. 3

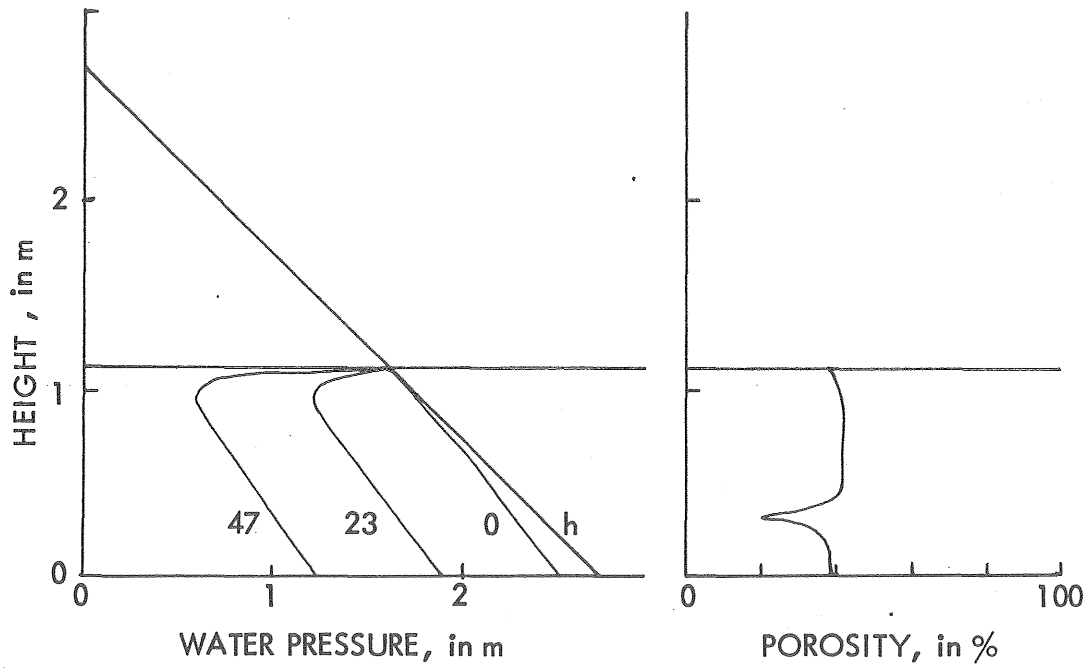


FIG. 7-21 Sand-activated carbon filter, filter 8.
Head loss and porosity in the filter bed.
Test 750707. Pilot plant No. 3.

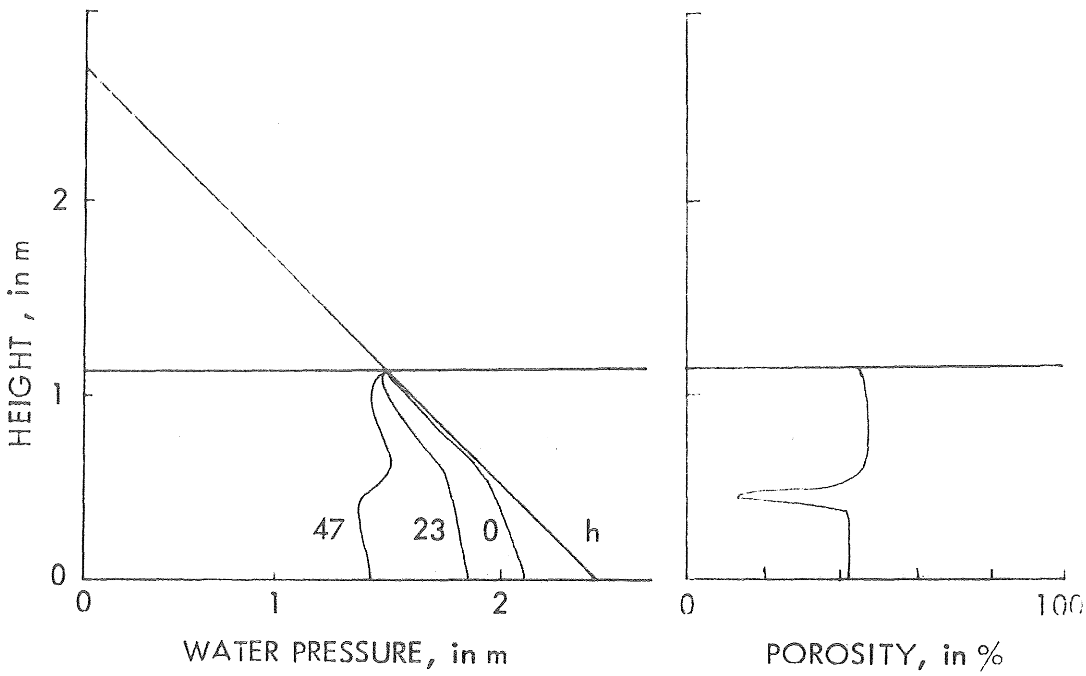


FIG. 7-22 Three-media filter, filter 9.
Head loss and porosity in the filter bed.
Test 750707. Pilot plant No. 3

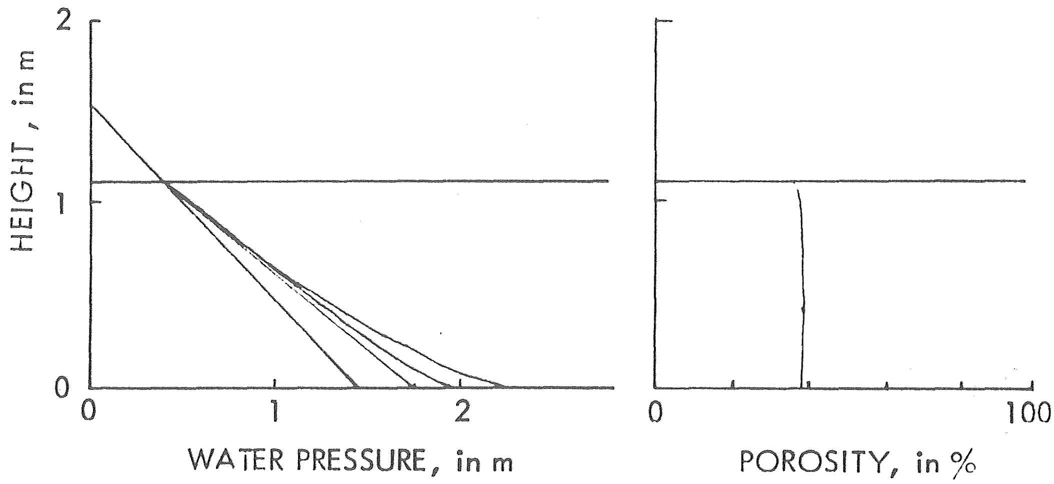


FIG. 7-23 Upflow filter, filter 10.
Head loss and porosity in the filter bed.
Test 750707. Pilot plant No. 3

From the figures it can be seen that a certain degree of mixing between the different media in the dual or in three-media filters occurs as the porosity is markedly changed. The porosity was measured by means of pouring water in the dry filter bed. The formation of the mixing zone at the interfaces between the different media is dependent on the media characteristics. For example, filter 5, FIG. 7-18, is very tight at a certain level. From a removal efficiency point of view it is of course advantageous, but this tight zone must be placed at the correct level in the filter bed in order not to reduce the depth of penetration in the bed. Studies carried out by Cleasby and Sejkora (1975) also show a high removal efficiency if an intermedia mixing zone is developed. Photographs of the different filters are shown in FIG. 7-24.

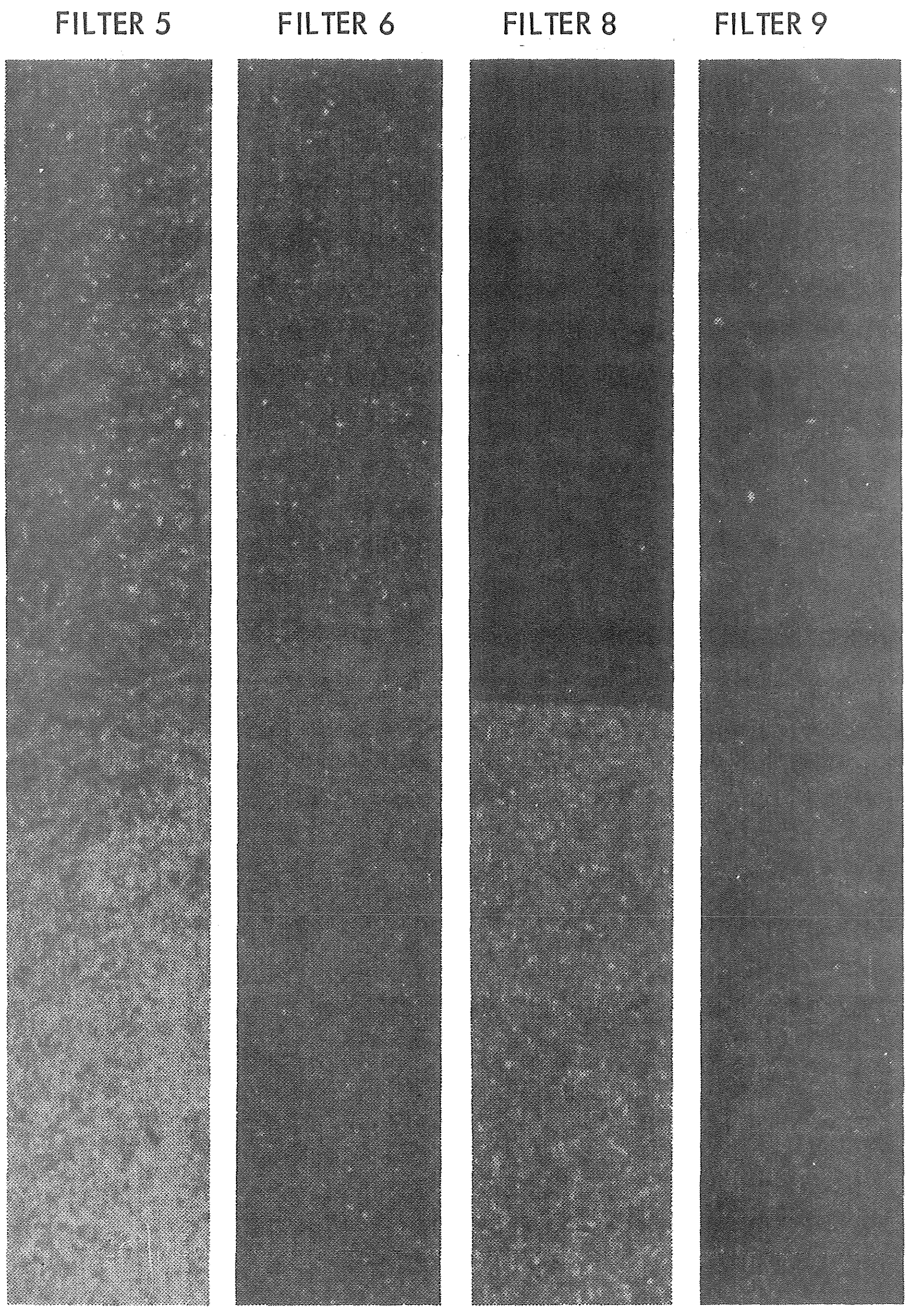


FIG. 7-24 Photographs of the mixing zone in the different filters

In addition to the size and density of the grains, the sphericity of the grain affects the degree of intermixing. In filter 5 and 6 and filter 9 two different types of anthracite have been used, (FIG. 7-25).

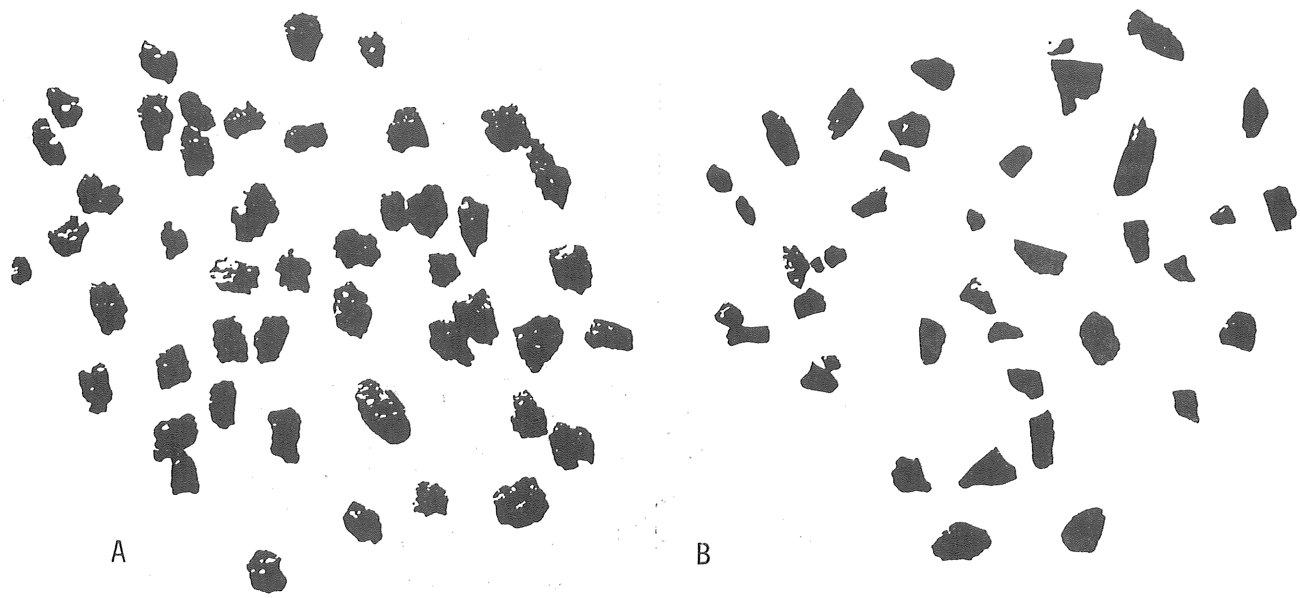


FIG. 7-25 Anthracite grains
A: Hydroanthracite
B: Anthracite

Type A anthracite has been used in the two-media filters and type B anthracite has been used in the three-media filter. Because the type A anthracite grains used in the two-media filters are more spheric, the porosity is higher than in the three-media filter.

A tight zone within the filter bed can under certain conditions – high water temperature, high dosage of activated silica – be dominant and completely decisive for the filter run time. One example of this is shown in FIG. 7-26.

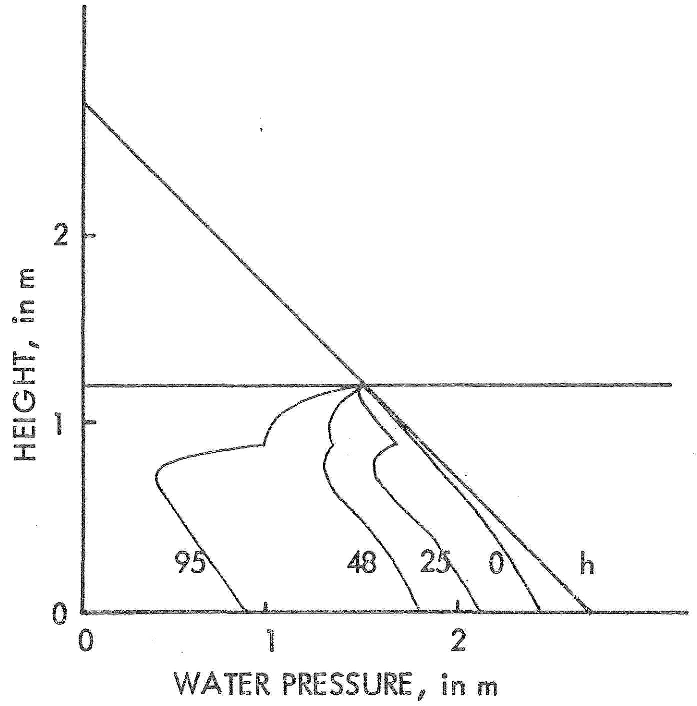


FIG. 7-26 Head loss in the filter bed. Sand-anthracite filter, filter 5, Pilot plant No. 3

In a filter with a tight layer the change in head loss as a function of time may deviate from a linear relationship and become an exponential function of time

An upflow filter is a design which combines high efficiency with maintained high porosity in the entire filter bed. This is also possible to obtain to a certain degree at least, by use of several media, selected in order to have a controlled mixing at the interfaces. Such a filter has been developed for experimental purposes, and the grain size in the top layer D_o has been chosen to have a certain relationship to the grain size in the lower layer D_u. The relationship was $D_o/D_u = 2.5$. The settling velocities of the different grains which are important for the settling after bed expansion at the back washing have also been considered. The experimental filter consisted of seven layers with grain sizes between 0.5 and 3.0 mm. The filter media were basically made up of a plastic material with the addition of various amounts of minerals of a high density. The head loss and the porosity in the filter bed are shown in FIG: 7-27.

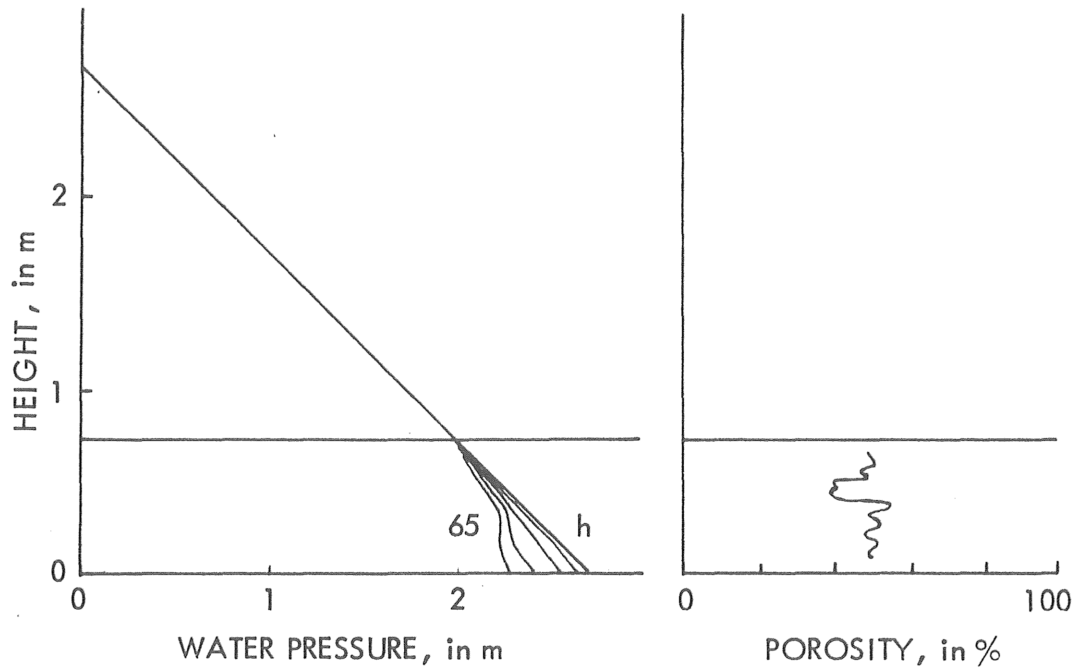


FIG. 7-27 Multimedia filter. Head loss distribution and porosity in the filter bed.

This figure shows that it is possible to maintain high porosity throughout the filter bed, and to avoid tight layers within the filter bed by a proper choice of grain characteristics.

The result with respect to porosity and head loss in the filter bed explains the different filtrate quality patterns, which have been shown in principle in FIG. 7-10. With the guidance of the result obtained, it is possible to obtain a maximum head loss in order not to exceed a given turbidity value in the filter effluent. In practice, the maximum head loss is more or less routinely chosen and the value is normally constant during the year and independent of, for instance, temperature and chemical dosage. In optimization of the filter operation it is desirable to develop a more refined method in order to decide which filter designs are the most favorable.

7.4 Model for the maximum head loss considering the water quality

The maximum permissible head loss $H_{f\ max}$ has been assumed to be a function of temperature, activated silica dosage, concentration of particles, and filter design. The relationship between the head loss, temperature and activated silica dosage for each filter is assumed to be described by a linear expression (FIG. 7-28).

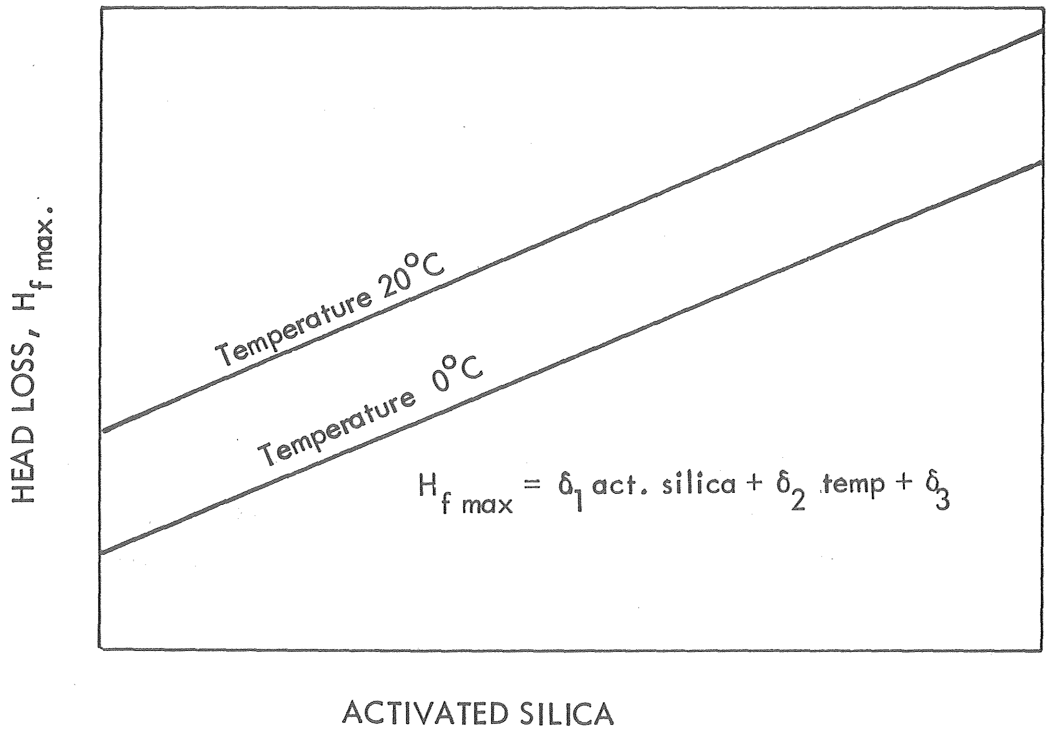


FIG. 7-28 Relationship between maximum permissible head loss, activated silica, and temperature.

A relationship of this type gives only an approximation of the situation, but it may be true within certain limits. Further research is needed to verify the suggested linear relationship.

The filtration experiments have shown that the maximum permissible head loss is dependent upon the particle concentration, measured in turbidity units. An increase in turbidity is normally caused by an increase in the overflow rate of the sedimentation unit, and this in turn results in a simultaneous change in the particle size distribution in the influent to the filter. It has earlier been stated by Hannah *et al.*, (1967) that the strength of the flocs is high if the floc size is small and low if the floc size is large. Therefore it is logical that an earlier breakthrough occurs at higher turbidity of the influent. This effect can be eliminated, at least to a certain extent, by an increase of the activated silica addition. The relationship shown in FIG. 7-28 may thus be corrected with regard to particle size or turbidity and addition of activated silica, and an empirical relationship between the maximum permissible head loss $H_{f\,max}$ and the variables mentioned can be suggested:

$$H_{fmax} = (\delta_1 A + \delta_2 t)\left(\frac{\delta_3}{\delta_4 + C_o}\right)\left(1 + \frac{A}{\delta_5}\right) + \delta_6 \qquad (7\text{-}11)$$

in which A is the dosage of activated silica

t is the temperature

δ_1 to δ_6 are coefficients, different for the various filters

It is possible to determine the different coefficients for each filter, but it is first necessary to decide the maximum permissible turbidity in the filtrate. The goal set up by the American Water Works Association in 1969 was a turbidity value of 0.1 JTU which corresponds to about 10 % Std 2 Sigrist.

In this investigation a value of 15 % Std 2 Sigrist has been chosen as a maximum value of the turbidity in the effluent.

In FIG. 7-29 and FIG 7-30 two examples are shown of the agreement between theoretical calculations and test results.

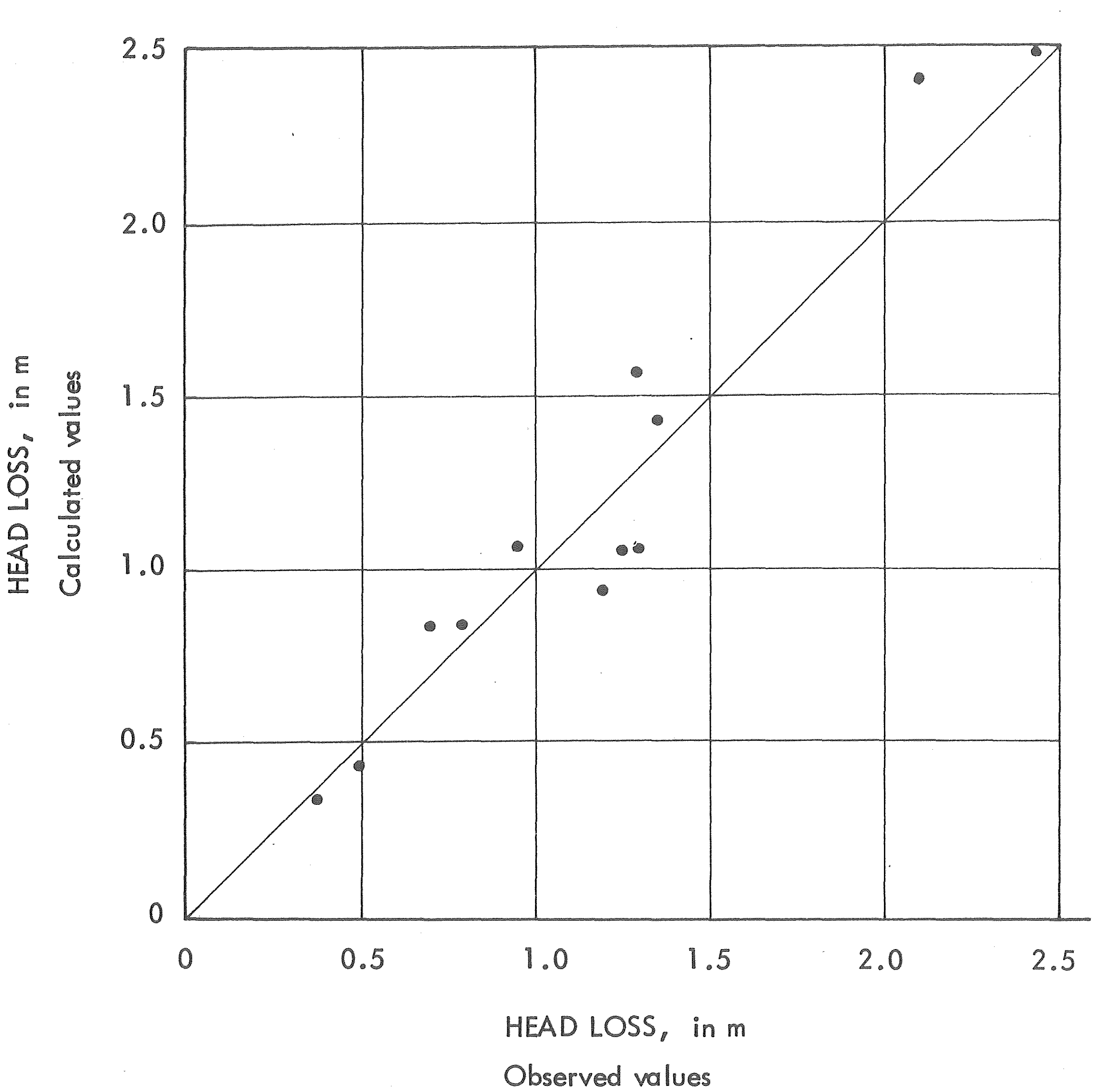


FIG. 7-29 Correlation between observed and calculated values of the maximum permissible head loss $H_{f\ max}$, Sand filter, Filter No.1 (r=0.96)

The other filters show similar relationships, and the different equations are shown in Appendix 9. A study concerning maximum permissible head loss ought to be performed in the future with filters technically equipped to increase the head loss until breakthrough of the filter. As a general conclusion it can be stated that the finer the media, the higher the influence of temperature on maximum permissible head loss.

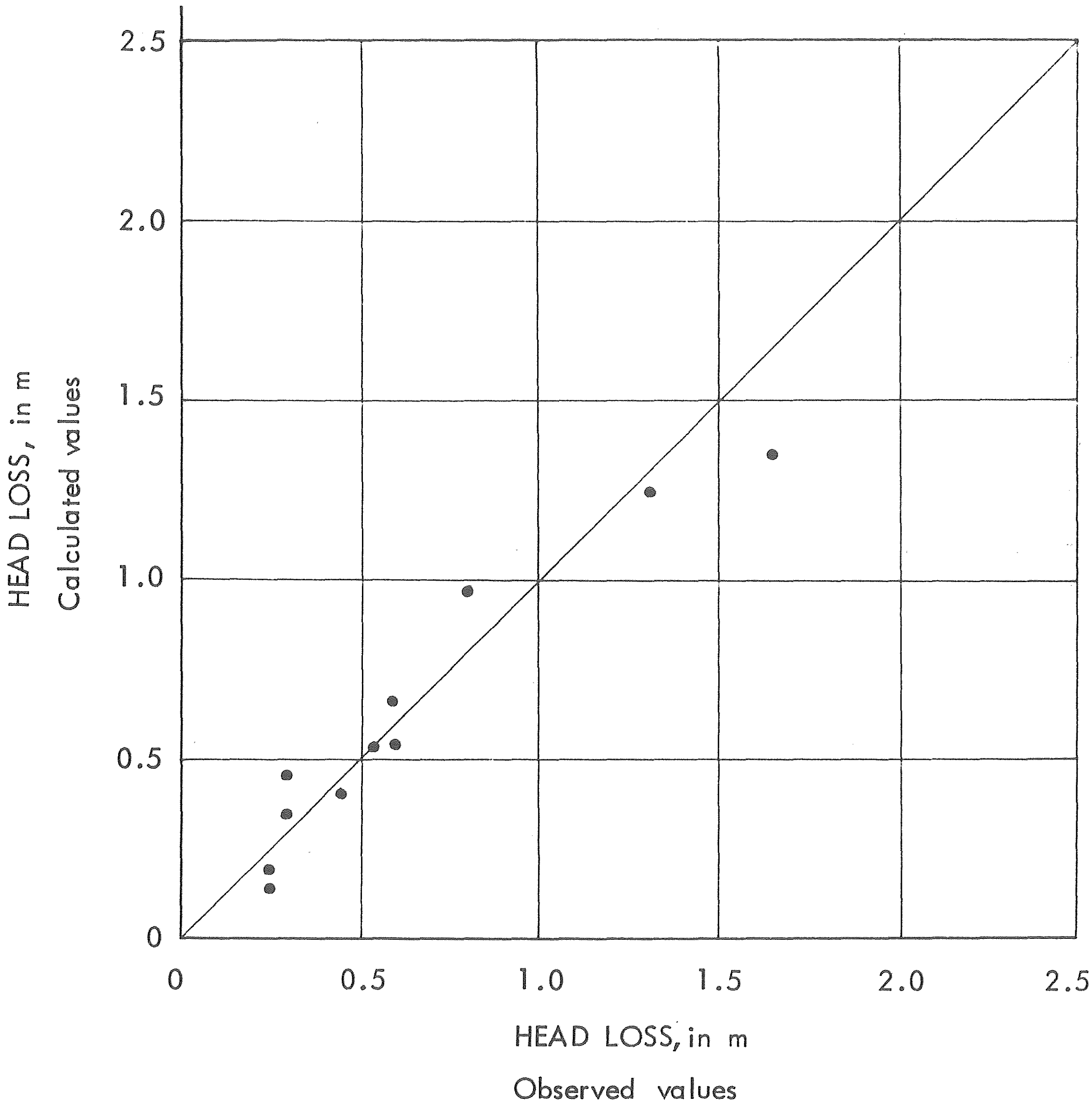


FIG. 7-30 Correlation between observed and calculated values of the maximum permissible head loss $H_{f\ max}$, Sand-anthracite filter, Filter No.5 (r=0.95)

7.5 Head loss caused by the removed particles

In Eqs. (6-18) and (6-19) approximate expressions of the total head loss have been given. The head loss H_f, caused by the removed suspended solids can be expressed:

$$H_f = k \cdot C_o \cdot v \cdot T_f \tag{7-12}$$

or

$$H_f = k \cdot C_o \cdot v \cdot T_f^{\nu} \tag{7-13}$$

in which k is a coefficient which is dependent on the actual filter design and the characteristics of the particles and the raw water quality

C_o is the turbidity of the influent

v is the filtration rate

T_f is the filter run time

ν is an exponent ≥ 1

In the case of very fine media filters, the development of the head loss is rapid and follows an exponential curve. This has also been observed when using dual media in which some intermixing of the two media occurs. In a filter with an effective grain size of about 0.8 mm, the increase in head loss with time is almost linear.

Several factors affect the change in head loss with time as can be seen from Eq. (7-12). The change is proportional to the filtration rate, concentration of particles, and a coefficient k. In order to determine the coefficient under different conditions, we have carried out experiments at varying temperatures and dosages of activated silica. In FIG. 7-31 and FIG. 7-32 the results for a conventional sand filter (filter No. 1) and a sand-anthracite filter (filter No. 5) are illustrated. In the figures all the data available are not represented. Further data are used as a verification of the models developed from the experiments. The relationships illustrated by lines in the figures should be regarded as hypotheses.

FIGURE 7-31 is typical for a "surface filter" in which the removal mainly takes place at the top layers. FIGURE 7-32 is typical for a "volume filter" in which the material penetrates more deeply into the bed. Thus, depending upon the filter design, the absolute value of the change in head loss per unit of time is considerably higher for a surface filter such as a conventional sand filter, than for that of a volume filter such as a coarse sand-anthracite filter or an upflow filter. The dependence on temperature is greater for a sand filter than for a sand-anthracite filter. This has also been observed in other investigations (Hedberg, 1974). According to theory the change in head loss with time is directly proportional to the particle concentration, but

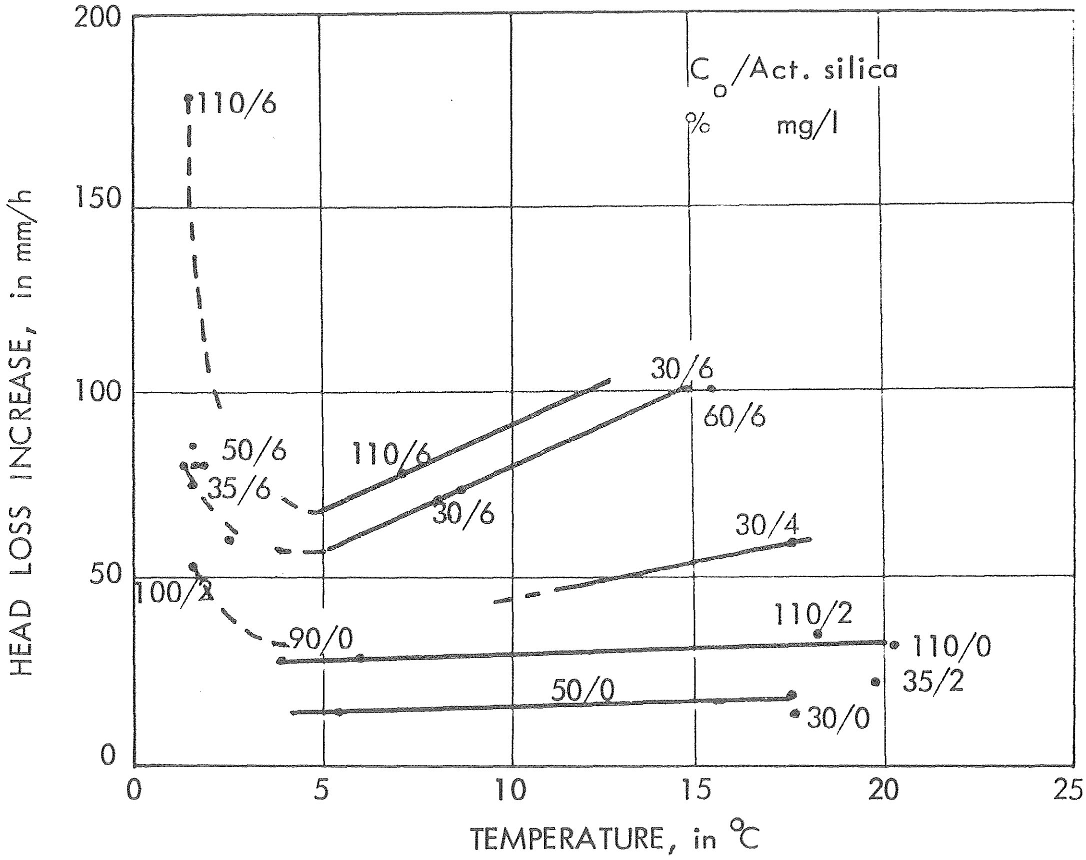


FIG. 7-31 Change in head loss per unit of time as a function of temperature. Sand filter, filter No. 1. Different turbidity C_o (% Std 2 Sigrist) in influent and different activated silica dosages A (mg/l). Filtration rate 10 m/h. Pilot plant No. 3). The dashed lines represent the conditions at low temperatures but these conditions are not considered in the filter models.

this does not seem to be true when activated silica is added. The concentration dependency seems to decrease with increasing addition of activated silica. This observation is valid to a greater extent for the surface than for the volume filters. The explanation for this fact may be that the shape and the size of the particles are changed. It can be assumed that the particle size is increased when activated silica is added. This affects the "surface filters" to a greater extent than the two-media filters as the width of the pores is smaller in a fine filter than in a coarse one.

The studies of the filtration operation at an extremely low temperature, below 2^{o}C, showed a remarkable change in the head loss increase for all the filters, but the head loss increase was most drastic for the "surface filters". The cause of this is

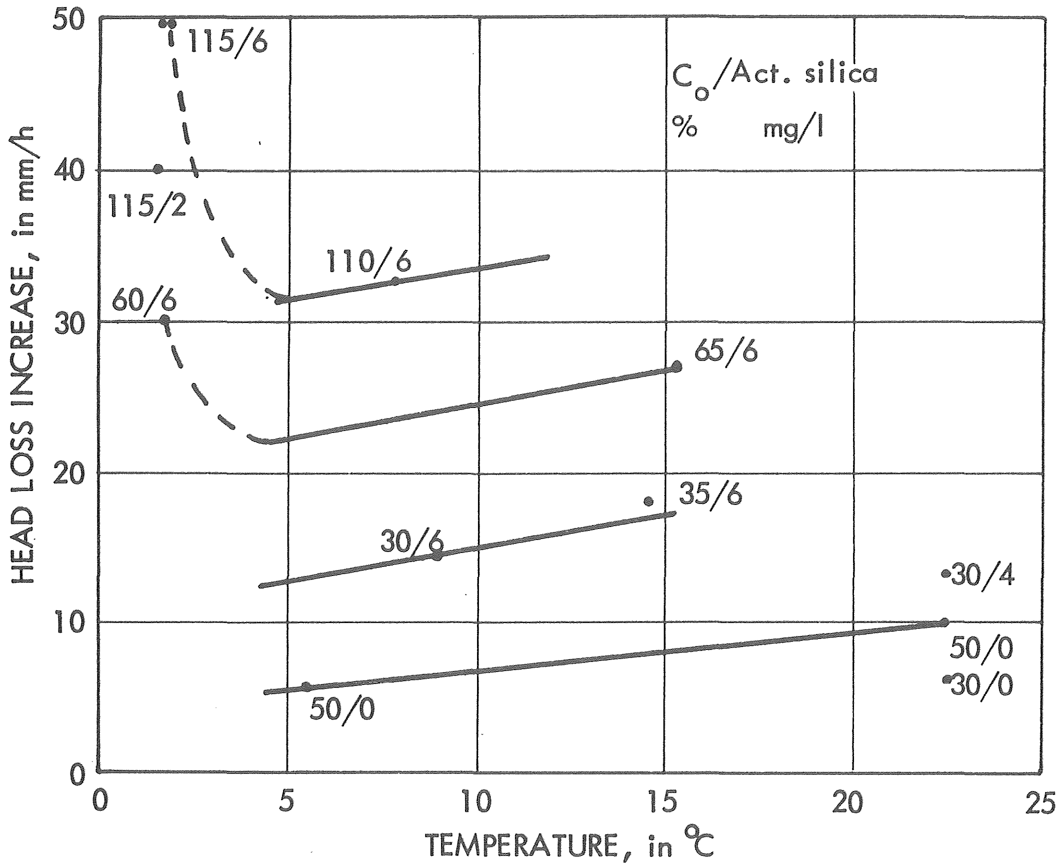


FIG. 7-32 Change in head loss per unit of time as a function of temperature. Sand-anthracite filter, filter No. 5. Different turbidity C_0 (% Std 2 Sigrist) in influent and different activated silica dosages A (mg/l). (Pilot plant No. 3).

probably to be found in the destabilization step and in the early steps of flocculation. The molecular movement, which is the driving force for the perikinetic flocculation, decreases with temperature. In water the hydrogen bonds are more dominant at lower temperatures. This may also affect the coagulation.

The temperature influence upon the filtration performance of the various filters can be shown by comparing the ratio between head loss increase at a high and low turbidity of the influent. This ratio has been plotted as a function of the effective media size, and the result is shown in FIG. 7-33.

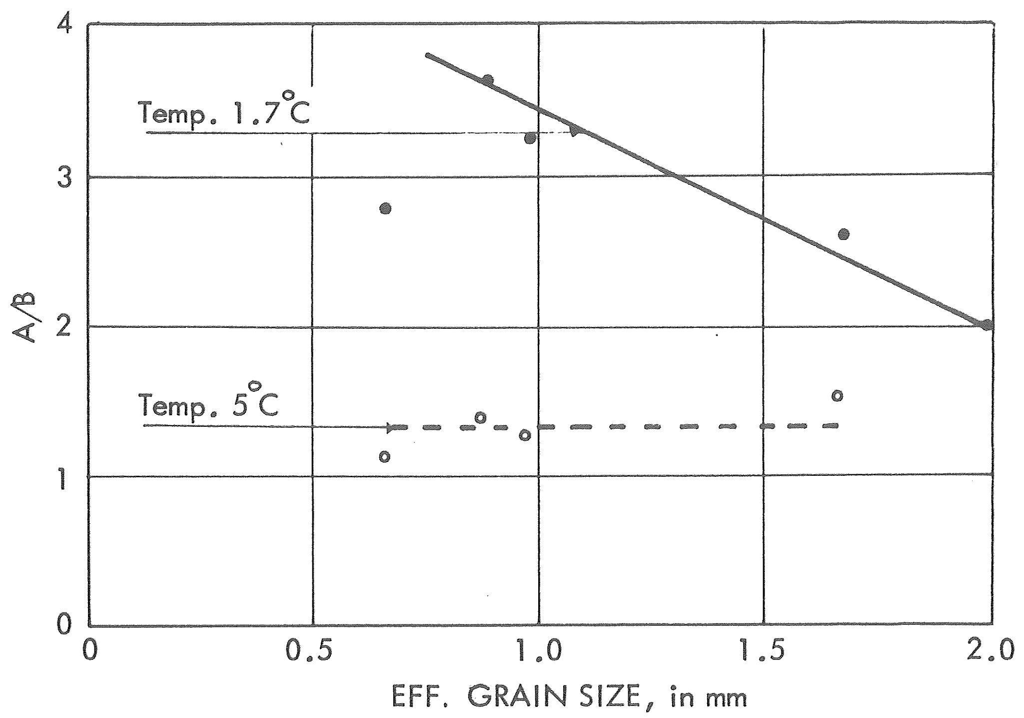


FIG. 7-33 The ratio: $\frac{\text{Head loss increase, high turbidity (A)}}{\text{Head loss increase, low turbidity (B)}}$ as a function of the effective grain size. Activated silica dosage, 4 mg/l

FIGURE 7-33 shows that when the turbidity is increased in the influent due to a high overflow rate in the sedimentation unit, the head loss increase for a conventional sand filter at water temperatures below 2^{0}C is drastically changed. A two-media filter with a coarse layer of anthracite is not affected to the same extent. At temperatures above 5^{0}C a change in turbidity affects all the filters to an equal degree.

It may be concluded that the particle characteristics change at an extremely low water temperature and that, depending on the filter design, the removal mechanism may also change. In order to avoid a too rapid head loss buildup, one can use coarser media or lower addition of activated silica. The overflow rate in sedimentation should, if possible, be kept low. When raw water is taken from a lake, the intake arrangements may be placed at such a depth that water with a temperature of less than 3^{0}C is avoided.

The increase in head loss with time was in most cases linear, but some observations indicate a slight exponential increase in the head loss with time. This was observed for the two-media filter, especially when activated silica was added.

In the optimization study it is necessary to describe the filtration operation under different conditions, and it is thus important to develop an expression for the coefficient k in Eq. (7-12), an expression in which temperature and activated silica addition are considered for each filter design. In the simpliest case the change in head loss per unit of time can be expressed as a linear function of temperature – within the interval $3^{o}C$ to $25^{o}C$ – as:

$$\frac{dH_f}{dT_f} = (k_t \cdot t + l_t)\ v \qquad (7\text{-}14)$$

in which T_f is the filter run time
t is the temperature
k_t and l_t are coefficients dependent on the particle concentration measured in turbidity units and on the activated silica dosage
v is the filtration rate

The coefficient k_t can be shown to be dependent on the activated silica dosage at a particular turbidity according to a relationship of the form:

$$k_t = (a \cdot A)^{\alpha} + b \qquad (7\text{-}15)$$

in which a, b are coefficients
α is a positive exponent

The head loss increase per unit of time at a given temperature should, according to the theory (Eq. 7-12), increase in direct proportion to the particle concentration. In the experiments it was, however, observed that the influence of the turbidity decreases with increasing addition of activated silica, which may be expressed mathematically:

$$k_t = (\{a \cdot A\}^{\alpha} + b) \cdot (d \cdot C_o)^{\frac{1}{1+f \cdot A}} \qquad (7\text{-}16)$$

in which a, b, d, f are coefficients

C_o is the turbidity of the influent

It has been observed that the dependence of activated silica addition and particle concentration on the intercept, the term l_t in Eq. (7-14), may be described by an equation similar to Eq. (7-15). The interdependence of the particle concentration and activated silica addition, which can be expressed by the coefficients "a" and "b" in an equation like Eq. (7-15) was observed to be quite different for the various filters. It is mainly the coefficient "a" which varies. Thus, it has been necessary to develop two particle or turbidity dependent terms, one for typical "surface filters" and the other for typical "volume filters". The expression for the term l_t as a function of activated silica addition and turbidity is:

"Surface filters"

$$l_t = \left(\frac{g \cdot A}{1 + h \cdot C_o}\right)^{\beta} + (j \cdot C_o)^{\gamma} \qquad (7\text{-}17)$$

Volume filters

$$l_t = (p \cdot C_o^{\delta} \cdot A)^{\beta} + (j \cdot C_o) \qquad (7\text{-}18)$$

in which g, h, j, p are coefficients

β, γ, δ, are positive exponents

The difference between Eq.(7-17) and Eq. (7-18) can be explained in the following way. Surface filters are more independent of an increase in turbidity with activated silica addition than volume filters, for which also the turbidity is of importance at high dosages of activated silica.

The equations expressing the change in head loss per unit of time due to particle concentration, temperature, and addition of activated silica are based on the results obtained. A detailed explanation and general description of the mechanism of the activated silica addition demands more specific studies. The following general observations have been made concerning the different filter designs:

The head loss increase per unit of time has been found proportional to the addition of activated silica raised to two, ($C\alpha=2$ in Eq.(7-15) and (7-16)).

The increase in head loss is almost proportional to the particle concentration measured in terms of turbidity in cases when activated silica is not added, ($C\alpha<1.5$ in Eq. (7-17) and (7-18)).

The values of the different coefficients for the various filters are presented in Appendix 9.

7.6 Initial head loss

The initial head loss through a clean filter bed has previously been discussed and is expressed in Eq. (6-16). Most of the filters studied in this investigation have, however, a relatively complex design, and it is difficult to theoretically determine the initial head loss H_o. The initial head loss has been empirically determined at a certain temperature and filtration rate. Thus, H_o can at any temperature and filtration rate be calculated by the following equation:

$$H_o = k_{H_o} \cdot \frac{v}{10} \cdot \frac{60}{(1.8\ t + 42)} \qquad (7\text{-}19)$$

in which k_{H_o} is a constant (determined at a filtration rate of 10 m/h and at a temperature of 10^oC)
t is the temperature
v is the filtration rate

7.7 Model for calculation of the total head loss in a filter as a function of time

The main equation describing the total head loss in a filter is:

$$H = H_o + H_f \qquad (7\text{-}20)$$

in which H_o is the initial head loss
H_f is the head loss caused by deposited suspended particles

The head loss H_f may be limited by the water quality (Eq. 7-11) or it may be fixed at any routinely chosen value. Equations (7-19) and (7-12) referring to the head losses H_o and H_f can be combined in accordance with Eq. (7-20):

$$H_o = k_{H_o} \cdot \frac{v}{10} \cdot \frac{60}{(1.8\ t + 42)} + k\ C_o v\ T_f \quad (7\text{-}21)$$

By combining Eqs (7-12), (7-14), (7-16), (7-17), (7-18), and (7-19) it is possible to calculate the filter run time T_f at a given head loss H, at a constant filtration rate V, (the integration limits in Eq. (7-14) are T_f:0 and T_f; H_f:0 and $H_{f\ max}$)

$$H = v\ \frac{78.5}{1.8\ t+42} + T_f(t\{a \cdot A\}^{\alpha}+b)\ \{d \cdot C_o\}^{\frac{1}{1+f \cdot A}} + \{\frac{g \cdot A}{1+h \cdot C_a}\}^{\beta} + $$
$$+\{p \cdot C_o^{\delta} \cdot A\}^{\beta} + \{j \cdot C_o\}^{\alpha})\Big| \qquad (7\text{-}22)$$

in which v is the filtration rate
t is the temperature
a,b,d,f,g,h,j,p are coefficients
$\alpha, \beta', \beta'', \gamma, \nu$ are positive exponents

Equation (7-22) is primarily based on theory, but partially the equation is based on empirical results. Thus, the equation is not general, and its use must be limited to the interval of the variables studied in this report.

Equation (7-22) has been tested by experiments performed on water from lake Bolmen. This lake has a similar water quality to that of lake Delsjön. The filter designs particularly studied in the investigation were conventional sand filters, (filter 1, 2, and 3) and sand-anthracite filters (filter 5 and 6). The coefficients in Eq. (7-22) have been determined during the experiments carried out at Lackarebäck (Pilot plant No. 3). The following variables and the variation interval have been studied.

Temperature, t: $1.4^{o}C - 20^{o}C$
Activated silica dosage , A:0-6 mg/l
Turbidity, C_o: 30 % - 120 % Std 2 Sigrist
Filtration rate, v: 10 m/h

FIGURES 7-34 and 7-35 illustrate the agreement between observed and calculated values for a conventional sand filter and a two-media filter.

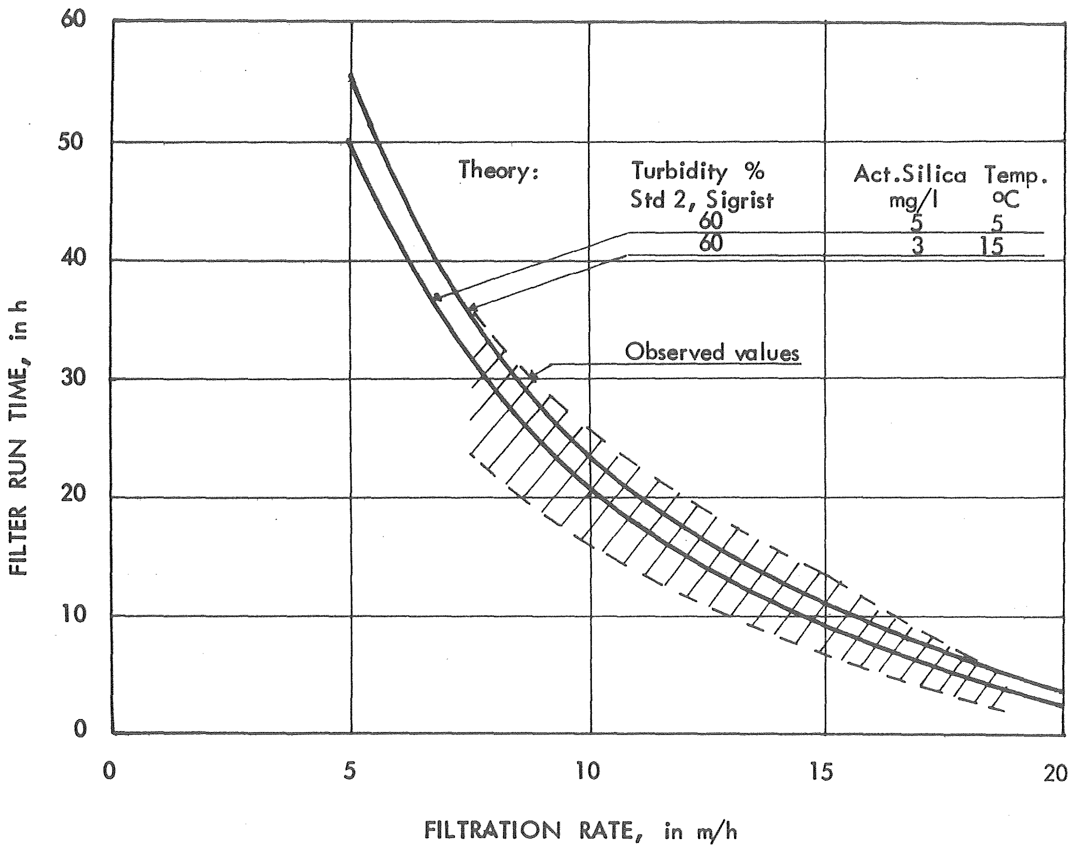


FIG. 7-34 Sand filter, filter No. 1

Agreement between observed and calculated values of the filter run time until a total head loss of 1.5 m is reached.

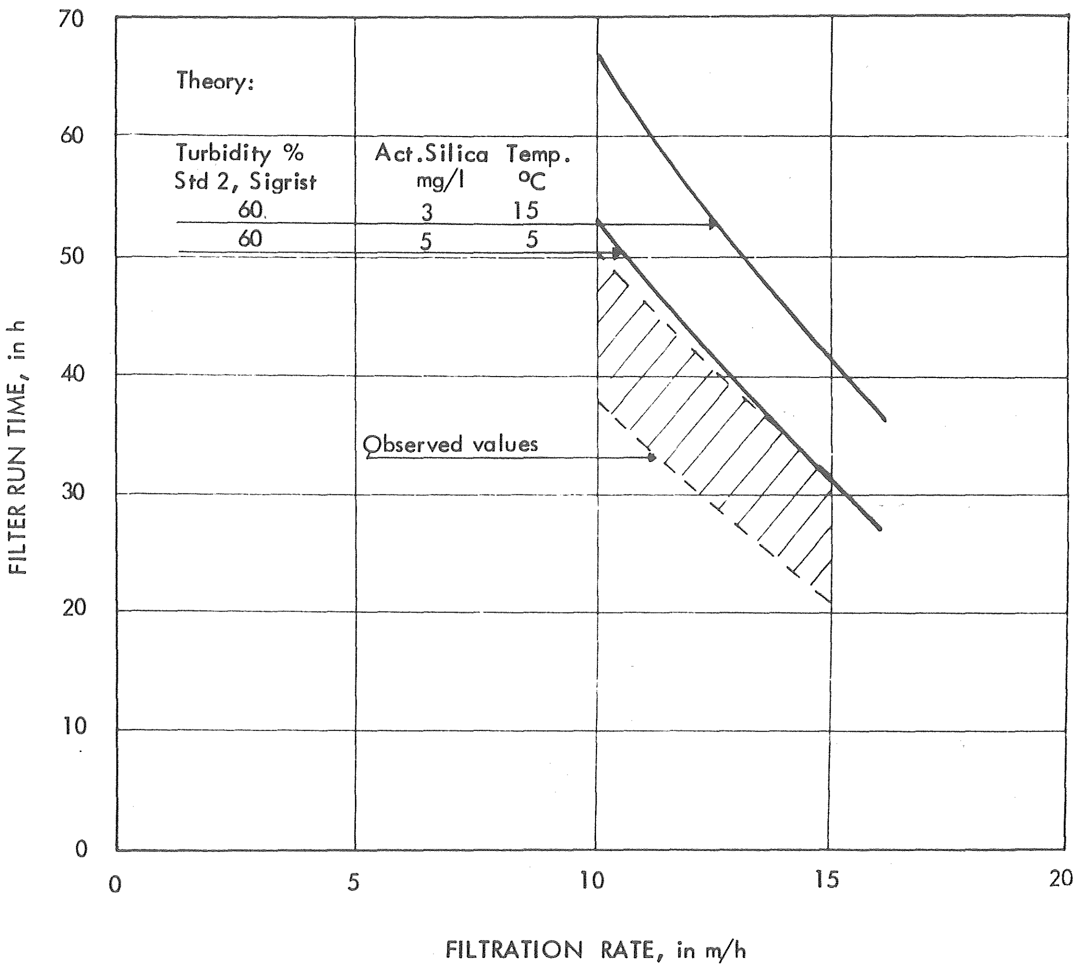


FIG. 7-35 Sand-anthracite, filter No. 5

Agreement between observed and calculated values of the filter run time until a total head loss of 1.5 m is reached

From FIG. 7-34 and FIG. 7-35 it can be seen that the agreement between observed and calculated values is satisfactory. There is, however, a certain discrepancy between predicted values and observed values for the sand-anthracite filter. This can be explained by the fact that the anthracite quality was different in the two pilot plant filters. The effective grain size was slightly smaller in pilot plant No. 4 than in pilot plant No. 5 in which the result was used as a basis for the empirical filter equations. The investigation was carried out over a period of years, but an effort was made to keep the different variables the same for all the tests. It was observed that the filters were very sensitive to changes in many of the variables. It is, for example, difficult to attain uniform quality of filter media in different filters even if they are taken from the same delivery. Good agreement in results from one pilot plant to another has been obtained. It is also interesting to compare pilot plant filters with a full scale filter , but this is more difficult because many factors vary with time. At the Water Treatment Plant at Lackarebäck a full scale test was made, and the filtration rate was raised from about 5.5 m/h up to 8 m/h. The results from this test have been summarized in TABLE 7-1 in which the increase in head loss per unit of time has been calculated.

TABLE 7-1 Results from the Water Treatment Plant at Lackarebäck, mean values.
Filter design: Activated carbon (filter No.8 without sand)
Turbidity in the influent 90-100 % Std 2 Sigrist
Activated silica dosage 2 mg/l. Temperature 10^oC

Numbers of filters in operation	Filtration rate m/h	$(\frac{\Delta H_f}{\Delta T_f})$ test mm/h	$(\frac{\Delta H_f}{\Delta T_f})$ theory mm/h
14	5.6	25	24.5
14	6.1	28	26.2
12	6.7	32	28.3
11	7.3	37	31.8
10	8.0	42	36.4

This summary shows a relatively good agreement between observed and calculated values of the change in head loss with time in full scale operation. The deviation may be due to different pH-values or particle characteristics.

The filters at the Water Treatment Plant were earlier conventional sand filters, and these filters were specifically studied in 1971 over a three month period (June 1 - August 23, 1971). The following operational data can be presented for the filter during this period:

Temperature t: 11.8 - 14.5^oC
Turbidity in the influent C_0: 70 % Std 2 Sigrist
Activated silica A: 6-8 mg/l
Filtration rate: 4-5.7 m/h. Mean value 4.9 m/h
Filter period 19-28 h. Mean value 23 h

A theoretical calculation based on a water temperature of 12.5^oC, a turbidity of the influent of 70% Std 2, an activated silica addition of 6 mg/l, and a filtration rate of 5 m/h, gives a filtration time of 19 hours to a total head loss of 1.2 m of water, which is in good agreement with the values obtained in practice.

7.8 Discussion of the model for total head loss as a function of time

The model is based on some assumptions which can be regarded as general; the change in head loss in a filter is directly proportional to the filtration rate and particle concentration. The particle concentration in the previous unit operations varies with capacity changes and chemical dosage. However, during such variations, which from an optimization point of view are essential, variables, other than particle concentration, such as particle size distribution and particle characteristics are changed. As the particle concentration in this investigation has been measured in turbidity units, and as this gives information on the product of the particle surface and the number of particles, the chosen method of sampling is of great importance for the determination of the coefficients in the model, since the particle characteristics may change in the sampling procedure.

7.9 Backwashing of filters

Rapid filters must be washed to restore their capacity when the effluent quality becomes unacceptable or when the pressure drop reaches a predetermined value. Several different techniques are used for the backwashing:

1. Fluidization and bed expansion with water
2. Washing with water and air
3. Washing with water and air, followed by bed expansion

Several different washing routines are applied at the water treatment plants. In this investigation the backwashing techniques have not been studied, but some important facts, assumptions and approximations used in this report should be mentioned:

1. Bed expansion is necessary when a multi-media or an upflow filter us used
2. A filter medium with a low density can be washed by a lower washwater speed than a medium with a high density
3. The consumption of washwater is independent of the density of the medium
4. The depth of penetration of the suspended particles is important for the total consumption of washwater and thus the filter design is essential for the total consumption of washwater

The first two statements are self-explanatory. A low backwash speed, which can be used when the grain size is small or when the grain density is low, does not significantly contribute to a reduced backwash water consumption. The porosity of the sand is about 40 %, and the type of flow in the sand at the backwashing is supposed to be almost equal to plug flow. In the water above the sand the flow is supposed to be almost completely mixed. In a conventional filter there is normally a certain depth of water above the sand to avoid negative head loss in the filter bed. It is necessary to change this water volume 3-4 times before the water is clean. In the sand, on the other hand, a small quantity of water is needed due to the hydraulic conditions. The consumption of washwater can be calculated as:

Washwater in the bed incl. 10 % expansion	Q_{s1}	1 m^3/m^2
Washwater in the water volume above the filter bed	Q_{s2}	3-4 "
Total consumption of washwater	Q_s	4-5 "

A magnitude of 4-5 m^3 of backwash water per m^2 surface area of the filter may be considered as normal. In the case of a sand filter with an effective grain size of about 0.8 mm, it is necessary to wash the filter at a rate of 50 m/h, which corresponds to an effective washing time of 6 minutes. If the grain size is smaller, the rate is lower, but the time required to change the water volume is longer and thus the consumption of washwater is the same. A small grain size or a low grain density contributes, however, in reducing the dimension of the washwater pipes and pumps.

In order to decrease the washwater consumption, one can lower the depth of water above the filter bed,or one can change the design of the washwater outlet, for example by a movable outlet level to empty the water above the sand during backwashing.

The fourth statement can be explained by the fact that a certain depth of water is needed in order to avoid the negative head loss during filtration. This water depth must be higher for "surface filters" than for "volume filters". The necessary water depth H_v can be calculated (FIG. 7-36) as follows:

$$H_v = H_f - \frac{H_b - H_o}{H_b} \cdot D_p \qquad (7\text{-}23)$$

in which H_f is the maximum head loss due to the deposited particles

H_o is the initial head loss

H_b is the depth of the filter bed

D_p is the penetration depth of the suspended particles

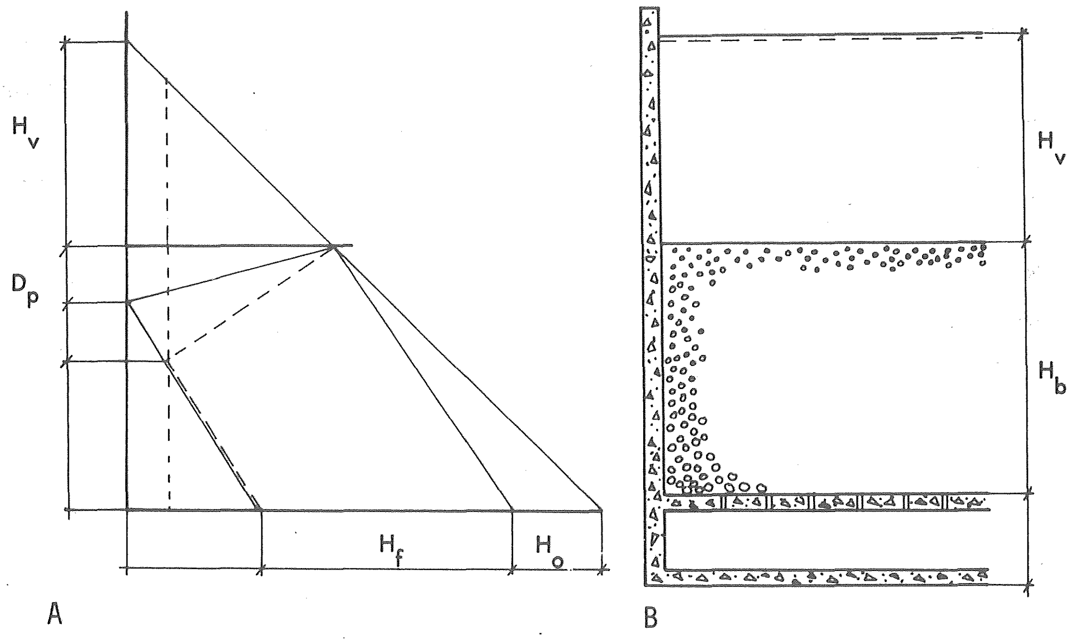


FIG. 7-36 A. Head loss distribution in the filter
B. Cross-section of a filter

Equation (7-23) also shows that the penetration depth is of importance for the water depth above the media, which in turn affects the consumption of backwash water. Equation (7-23) has in this study also been used in the calculation of the total height of the different filters.

The washwater consumption can roughly be calculated as follows:

$$Q_s = (3 \text{ to } 4) \cdot H_v + H_b \qquad (7\text{-}24)$$

in which Q_s is the washwater consumption in m^3/m^2
H_v is the depth of water above the media
H_b is the depth of the filter bed

The equation is valid as long as no special arrangements have been made in the washwater outlet design.

7.10 Safety factor

In the filter design the filtration rate should be as high as possible considering the filtrate quality and economic conditions. As a basis for the selection of a suitable filtration rate, the relationship between the amount of water produced per day ($m^3/m^2 \cdot d$) and the filtration rate can be useful. One example is illustrated in FIG. 7-37

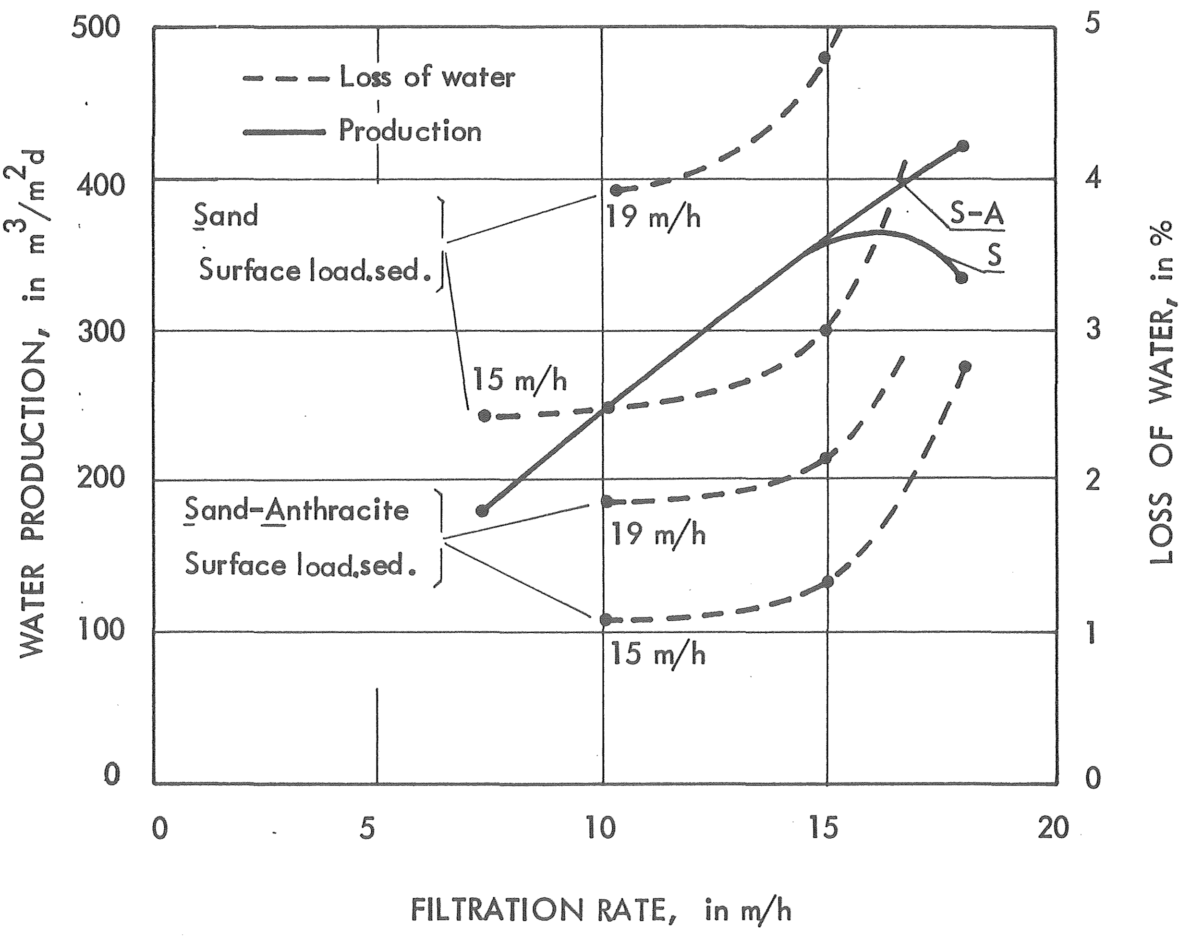


FIG. 7-37 Production of water and the loss of water as a function of filtration rate.

From FIG. 7-36 it can be seen that the production of water in cubic meter per square meter of filter area per day increases with the filtration rate until a particular value is reached, and then the production decreases. Simultaneously, the loss of water during the backwashing increases and under varying conditions an optimum filtration rate may be obtained. However, the range of variation in the factors affecting the filter run time, and thus indirectly the loss of water, must be considered. The variation in turbidity in the influent and the addition of activated silica are very important. The filtration rate, calculated under optimum conditions, may be reduced to some extent because of variations in the vital factors affecting the filtration performance. If the filter performance has been considered under different circumstances the safety factor may be slightly above unity.

8 CONCLUSIONS OF THE TECHNICAL INVESTIGATION

A summary of the theory of flocculation kinetics is presented, and based upon this theory a new flocculation principle has been suggested. In this system the residence time in the primary reactors is shorter than in the terminal reactors. The matematical model describing flocculation performance in reactors in series has been based on the work of others (Argaman and Kaufman, Parker, Kaufman and Jenkins). This model is based upon the assumption that the breakup of flocs is caused by floc surface erosion. It has been improved to some extent in order to be able to use the actual velocity gradient in a single reactor instead of using the mean velocity gradient of the whole flocculation system. An empirical model has been derived, describing the settling properties of the flocs, based on the flocculation performance.

The sedimentation models derived are based on the Hazen/Camp ideal sedimentation theory for discrete particles. Two different models, one for a conventional sedimentation basin and the other for a lamella sedimentation unit, have been suggested.

The four models – the flocculation model, the settling characteristics model, and the two sedimentation models – make it possible to predict the concentration of the suspended solids in the effluent from a conventional as well as from a lamella sedimentation unit at different water temperatures, dosages of coagulant aid, flocculation parameters, and sedimentation designs.

Several different models for the filtration operation have been described in the literature. The different models are developed for ideal conditions and are supposed to predict the removal efficiency as well as the head loss in a filter for any given period of time. The different models derived are more or less special cases of a model suggested by Ives, who also has proposed an approximate filtration model describing the change in head loss with time. The model used in this investigation is based on that approximate model, but it has been necessary to modify it empirically, due to wide variation in the filtration variables. An empirical model has been developed to describe the water quality in the filtrate at a given head loss. Another empirical model has been made for the maximum permissible head loss through a filter considering the water quality. The different variables studied and the various models are summarized in TABLE 8-1.

From the studies of the flocculation operation the following conclusions may be drawn. The flocculation process is affected by temperature, especially when the residence time is short and when no activated silica is added.

An increase of the floc volume fraction increases the flocculation performance. The optimum pH-value is influenced by temperature. The optimum pH-value at a temperature of 15 to 16^{o}C is 6.2-6.3 and 6.4-6.6 at a temperature of 4^{o}C.

Several studies were performed to examine the hydraulic conditions in a reactor. The geometrical design was found to be unimportant at least in the relatively normal units used in these investigations.

Tracer studies carried out in order to determine the residence time distribution showed that the energy input (velocity gradient) affected the time distribution. At higher velocity gradients the reactor was almost completely mixed and at lower velocity gradients the reactor was characterized by a lower degree of dispersion .

The residence time distribution in a reactor was also influenced by the rate of flow through the reactor. The amount of dispersion in the reactor decreased with an increase in the rate of flow.

The installation of a baffle in a reactor had, of course, a marked effect on the residence time distribution.

The direction of the flow through the reactor was not important for the flocculation performance.

Different paddle designs were briefly studied, and it was observed that the paddle design was important. A paddle design factor was developed to characterize the energy distribution in

TABLE 8-1 Matematical models and system variables

VARIABLES				
Flow of water	Chemicals	Flocculation	Sedimentation	Filtration
Q (m^3/h)	Act.silica (mg/l)	Temperature (^oC) Flocculation system Number of reactors Velocity gradient (s^{-1}) Residence time (s) Settling characteristics of the flocs.	Temperature (^oC) Conventional basin Lamella sedimentation unit Lamella length (m) Lamella width (m) Lamella distance (m) Overflow rate (m/h)	Temperature (^oC) Filter design Filtrations rate (m/h)
MATHEMATICAL MODELS				
		Residul turbidity - residence time during flocculation Equation (3-3) Residual turbidity - settling characteristics Equation (2-38) Equation (3-5) Equation (5-7)	Turbidity in effluent - surface loading Equation (4-4) Equation (5-4) Equation (5-6)	Filtrate quality Equation (7-9) Permissible head loss Equation (7-11) Head loss - filter run time Equation (7-22) Water production Equation (6-2)

the reactor. A paddle with a relatively large cross sectional area with several paddle blades was found to be very effective. The floc size distribution was uniform, and the overflow rate in the sedimentation unit could be increased.

A new type of compartmentalization of the total flocculation volume was tested. Instead of a series of reactors of equal size, the primary reactors in the series were smaller than the terminal ones. The results of these tests showed that it was possible to increase the flocculation performance by using this new design.

The studies of the settling characteristics of the flocs performed in a settling column in which the flocs were allowed to settle, showed that the depth of sedimentation was important, both for the mean settling velocity and the standard deviation of the settling velocity of the flocs.

Several different designs of lamella sedimentation units have been tested and the results can be summarized as follows:

The experimental result agrees with the theory, showing that an increase in settling velocity of the flocs causes a corresponding increase in the overflow rate. According to the theory the surface rate Q/A is increased in proportion to the relationship between the length of the lamella and the horizontal distance between the lamellae (L/S-ratio). This has been confirmed by the investigation.

Further, the investigation shows that the removal efficiency slightly increases with increasing L/S-ratio.

The overflow rate Q/BL is independent of the horizontal distance between the lamellae but there exists a critical distance depending on the thickness of the sludge sliding down on the lamella. The critical distance is dependent upon the volume concentration of the sludge. The critical distance for the particular sludge studied in this investigation was estimated at 0.04 m. The corresponding critical rate of flow was calculated to be 0.005 m/s.

More favorable and constant hydraulic conditions are developed in the lamella sedimentation unit in comparison with the conventional basin. Reynold's number is less than 100 which means that the flow is laminar.

It has also been observed that the settling properties of the flocs are changed in the lamella sedimentation unit. The settling velocity distribution of the flocs is uniform.

The surface loading is dependent upon the design of the inlet and outlet of water in the lamella unit. A surface loading as high as 40 m/h has been reached in a unit with an ideal inlet and outlet of water. In the full scale unit at the Water Treatment Plant at Lackarebäck a surface loading of 17 to 18 m/h has been reached. This lamella sedimentation unit has a lamella length of 3.0 m and a horizontal distance between the lamellae of 0.08 m. The corresponding surface loading of the conventional Lovö-basin is about 3 m/h.

The filtration operation has been studied under various conditions and the investigation has included many filter designs.

All the downflow filters may be considered representative, and the results obtained may be used as a basis for full scale design. The upflow filter is, however, due to the principle, dependent on the scale and the actual design of the system, which prevents the bed from lifting. Thus, even if several investigations have been carried out with upflow filters with promising results, it is doubtful if they can be used to predict the result in a full scale plant. Several conclusions may be drawn from the investigation, and some of the most important ones are summarized in the following paragraphs.

In terms of turbidity at the initial stage of the filter period, the filtrate quality measured is almost equal for the different filters. The filtrate quality pattern during the filter run time is, however, different for the various filters, and the quality pattern is strongly dependent on the addition of activated silica and the temperature.

The filtrate quality has proved to be almost independent of filtration rate at least up to a rate of 15 to 20 m/h.

It has also been shown that an increase in turbidity in influent water causes an increase in removal efficiency.

The maximum permissible head loss through the bed, considering the filtrate quality has empirically been expressed in a mathematical form. The maximum permissible head loss generally increases for all the filters with an increase in temperature, and dosage of activated silica. The maximum permissible head loss is higher for filters with finer media than for filters with coarser media. Especially high head loss can be obtained for the finer dual-media filter or the three-media filter. In the two- and three-media filters the different layers mix at the interfaces, which in general leads to a high removal efficiency and resistance against a breakthrough due to a high head loss. In the mixing zone the pore width is small, which causes a rapid increase in head loss, and it is thus necessary to be able to control the degree of mixing by choosing the right media.

The increase in head loss with time is dependent upon temperature, dosage of activated silica, particle concentration,and filter design. The conventional filters with only one medium are in general more dependent on temperature and the amount of activated silica added than the two- or three-media filters, which are more dependent on the particle concentration. In general, the rate of head loss increase for a coarse dual-media filter is about half of that of a conventional sand filter. At extremely low temperatures a very rapid increase in head loss has been observed, especially when activated silica is added to the conventional filters. This is probably due to a change in particle characteristics originating from a change in the perikinetic flocculation caused by a decrease in the molecular movement. This emphasizes the necessity of placing the raw water intake under the stratification layer when a lake is used as a water source. The coarse dual media filters were not affected to the same degree,and thus it might be considered that the filtration mechanism was changed and straining probably became the dominant mechanism.

The filter run time is dependent upon the permissible head loss. If the available head is fixed to a certain value, for instance 1.5 m of water, the filtration time for a conventional sand filter can be estimated at 20 hours and 10 hours at the filtration rates of 10 m/h and 15 m/h, respectively. Corresponding values for the coarse sand-anthracite filter was about 45 hours and 25 hours, respectively.

The washwater consumption for the different filters has been assumed to be almost independent of the media size. The consumption of washwater for the filter is dependent on several factors, but in general the washwater consumption for a conventional sand filter and a sand-anthracite filter may be estimated at 2-3 % and 1-2 %, respectively.

The results concerning the conventional sand-filters and activated carbon-filters were all in agreeement with the results obtained in full scale testing at the Water Treatment Plant at Lackarebäck.

Based on the technical investigation alone, it is impossible to optimize the water treatment process.

9 COMPARATIVE COST OF TREATMENT

9.1 General

The cost of any system of water treatment is made up of different categories of costs such as capital charges, chemical costs, and labour costs. The types of costs will vary depending on the location of the plant. In this investigation the cost calculation is based on experience from the Water Treatment Plant at Lackarebäck in Göteborg, and it considers mainly those costs linked to the different unit operations that are affected in size and design by the optimization. Different types of costs have been put together in groups which in principle are dependent on horizontal surfaces, vertical surfaces, and volumes. This means relatively great freedom in the choice of detail designs within the limit of the stated cost.

9.2 Relative costs

Relative costs mean the total costs related to the effective volume of reactors, basins, and channels in which the operation is performed. The different costs which have been considered in the relative costs are investment, capital, and maintenance costs, and they include the following:

Chemicals: Price, delivery, and handling.

Flocculation: Raw water distribution system
Mixing equipment
Reactor volume
Stirring equipment
Baffles
Foundation
Building
Ventilation, heating
Maintenance

Sedimentation: Water distribution system
Sedimentation basin
Lamella sedimentation equipment
Sludge removal equipment
Foundation
Building
Ventilation, heating
Maintenance

Filtration: Water distribution system
Filter basin
Filter media
Gallery and control equipment
Foundation
Building
Ventilation, heating
Maintenance

Most of the costs are easy to calculate in an exact manner, but some have to be estimated. The following approximations have been made.

> The cost of the stirring equipment is assumed proportional to the reactor volume.
>
> The cost of the gallery and the filter control equipment is assumed proportional to the number of filters.
>
> The cost of the maintenance is assumed inversely proportional to the flow of water.

In addition to the above mentioned costs, investment in and maintenance of the chemical dosing equipment also affect the relative costs, but not to an equal extent, and are therefore not considered in the calculation.

9.3 Design principles

The design of the water treatment plant affects the different costs. A conventional treatment process design is illustrated in FIG. 9-1.

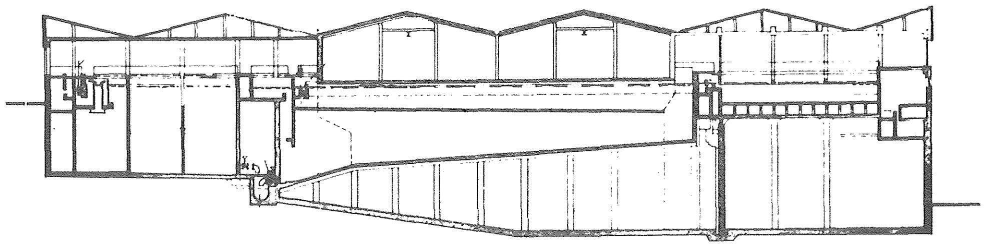


FIG. 9-1 Longitudinal section through the Water Treatment Plant at Lackarebäck.

In addition to a conventional design, a design with a lamella sedimentation unit has also been studied (FIG. 9-2).

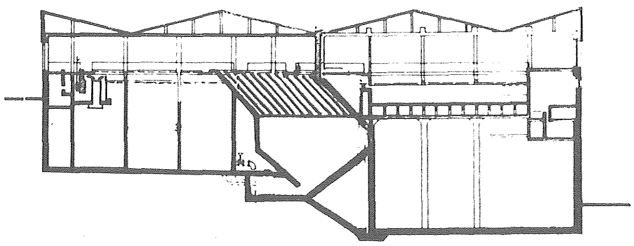


FIG. 9-2 Longitudinal section through the treatment plant with a lamella sedimentation unit.

A detailed section and plan of the lamella sedimentation is shown in FIG. 9-3.

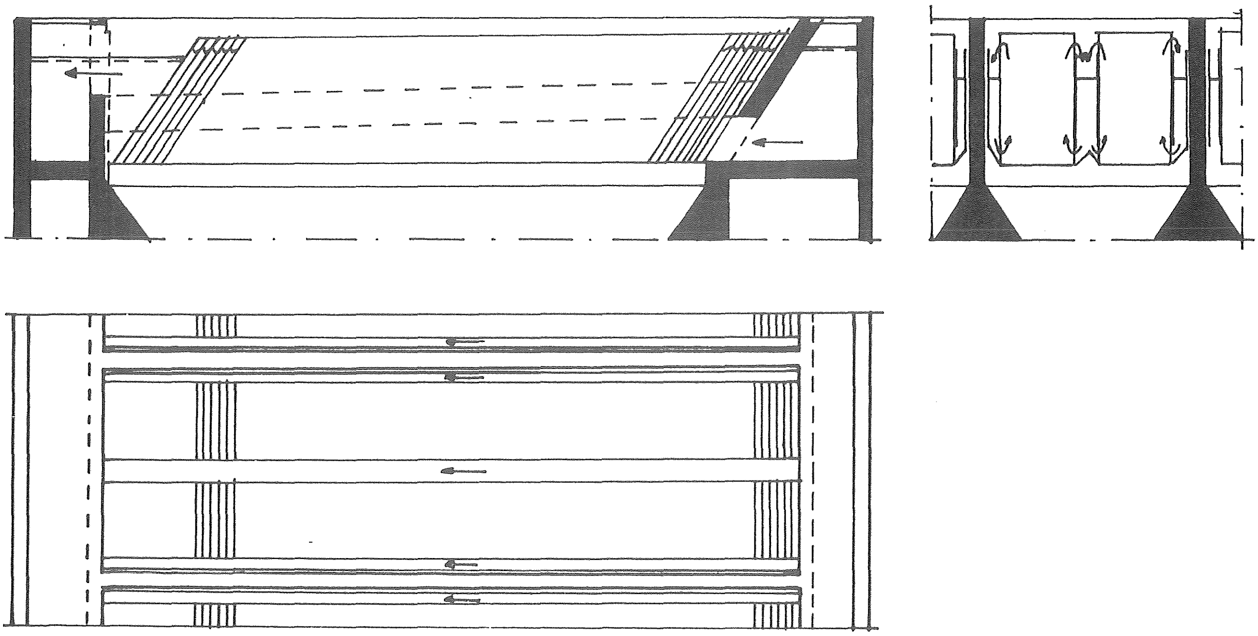


FIG. 9-3 Section and plan of a lamella sedimentation unit.

Several designs, different from conventional ones, can be suggested for flocculation-sedimentation linkage. Various possibilities are illustrated in FIG. 9-4.

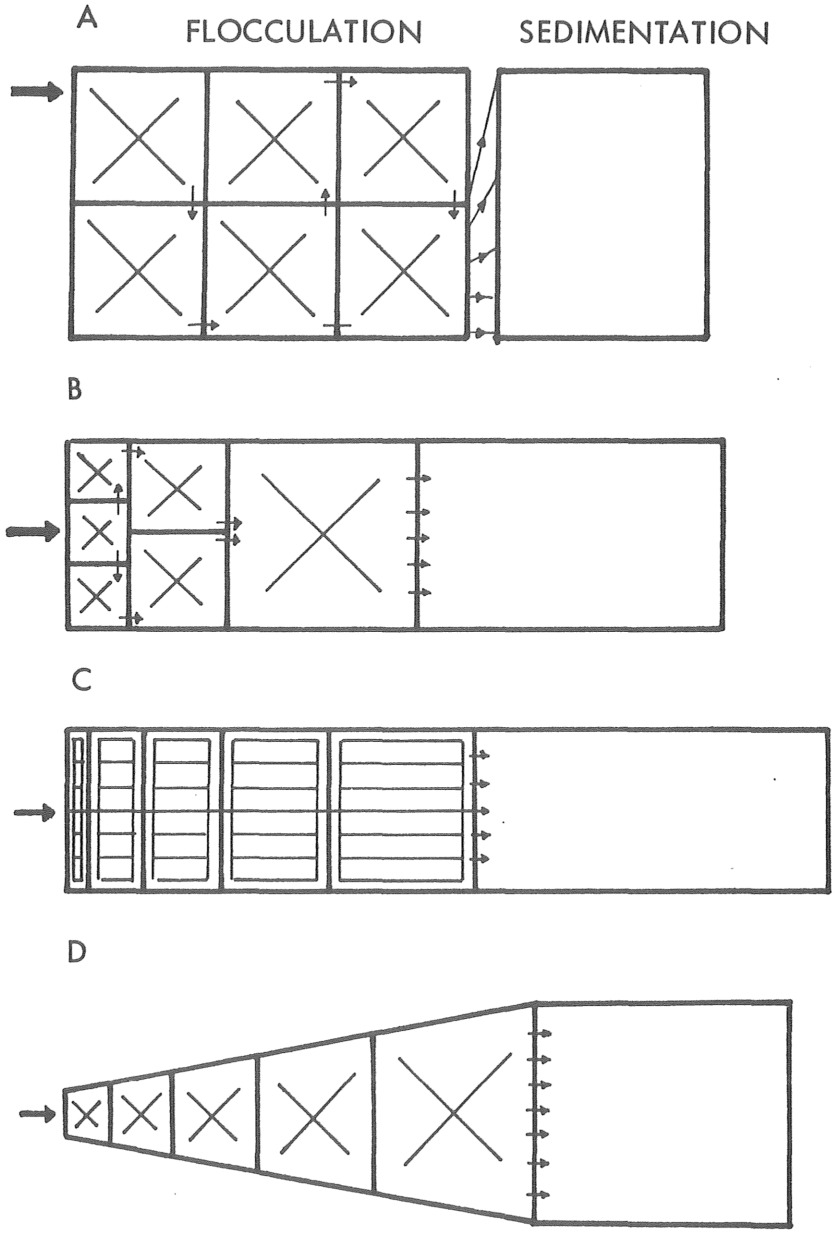


FIG. 9-4 Designs of flocculation and sedimentation systems

The conventional design of the flocculation unit usually consists of 4 to 6 reactors in series. Generally, the water is distributed in channels to the separation unit. This design should be avoided due to undesirable sludge settling in the distribution channels. By applying the designs shown in FIG. 9-4, b,c,d, one can avoid this problem. In the design shown in FIG. 9-4 b, the water is led to a common reactor and is then divided into two lines, which are brought together in the last reactor. Another technique is to use horizontal stirring equipment so that the reactor volume can easily be divided as desired. The design shown in FIG. 9-4, c, offers great advantages for larger treatment plants as a centralized chemical handling and dosing system is obtained. This design requires that the water treatment plant has a circular form. Suggested designs may be modified and adapted to particular requirements.

In the optimization, any value may be chosen for lengths, widths, and heights, but in order to avoid unreasonalbe designs, one can use the limitations shown in FIG. 9-5.

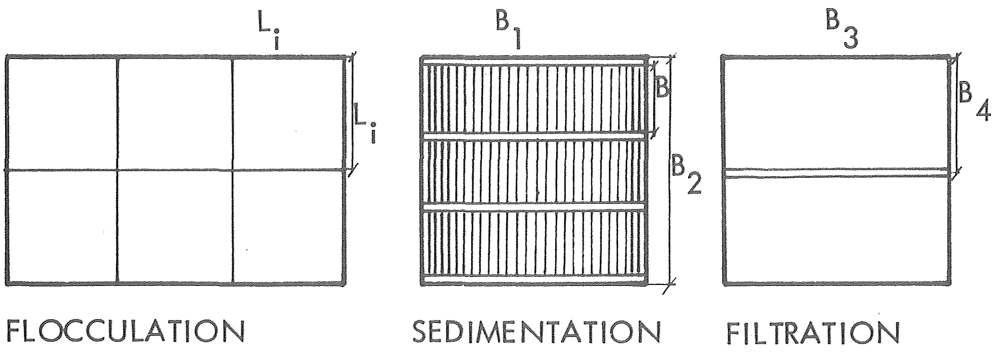


FIG. 9-5 Design principles

The last reactor of the flocculation unit must have an equal width, or half the width, of the sedimentation unit ($L_i = B_1$ or $2 L_i = = B_1$).

In the lamella sedimentation unit, the relationship $B_2/1.2$ B must be a whole number. This means that the width of the water distribution channels between the lamella packages is between 0.3 and 0.5 m for normal values of the width of the lamellae. The ratio between the length (B_4) - and the width (B_3) of a filter is important in order to obtain as low a gallery cost as possible. According to Weber (1972), the relation B_4/B_3 ought to be 3.

There are many principles of filtration rate control. In this investigation a system with a constant filtration rate and a constant water level in the filter unit has been studied.

The unit operations are affected by temperature, and it is necessary to consider this fact when designing a water treatment plant. But even the water consumption varies with temperature and studies by Wunsch et al (1959), Asemann and Wirth (1973), and Rechard (1971) show an increase in water consumption with temperature. The value of the increase seems to be dependent on several factors. The investigations show that there is a tendency of a greater increase in the water consumption with temperature in smaller communities than in larger ones.

In Göteborg the water consumption does not seem to increase with temperature at all but instead decrease , due to the fact that in the summer people move to the vacation areas surrounding the city. The water consumption in Göteborg during 1974 and 1975 is shown in FIG. 9-6.

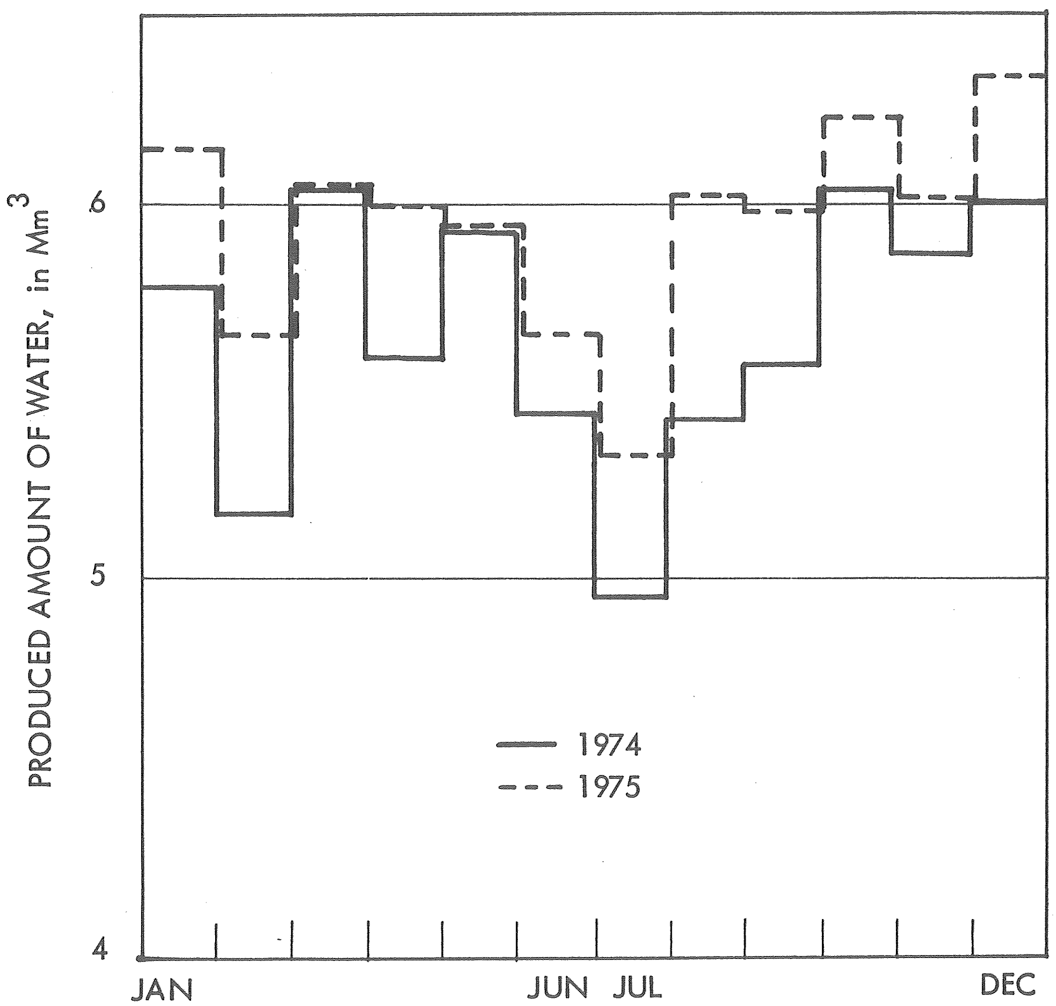


FIG. 9-6 Water consumption in Göteborg in 1974 and 1975.

10. RESULT OF THE TECHNICAL AND ECONOMIC ANALYSIS

10.1 Scope of the study.

The construction of the water treatment plant model, shown in principle in FIG. 10-1, makes it possible to study almost any water treatment plant design. Several parameters can be studied independently of each other.

A complete optimization of a water treatment plant has not been carried out in this investigation. Instead, a limited analysis of the flocculation, sedimentation, and filtration operation has been made in order to find the most economical solutions and to show some of the possibilities of how to use the water treatment plant model. Some designs of water treatment plants have been studied at the "design water temperature" which in this case has been chosen to be 5° C. Some of the most interesting combinations of flocculation, sedimentation, and filtration units have then been studied at different temperatures and chemical dosages thus simulating the plant operation during the year.

In the following the result of the technical and economic calculations are shown in tables and graphs. The economic calculation has been based on 1975/1976 prices (in Sw.cr.) in Sweden. As not all the details in a water treatment plant have been considered, the economic calculations must be regarded as comparative.

Three different types of costs, investment, running, and total costs, have been calculated. The investment cost has been expressed in Sw.cr. per cubic meters of building volume (Sw.cr./m^3) The capital cost, the maintenance, and the total cost have been expressed in öre per cubic meters of water (öre/m^3).

The different basic prices can easily be changed in the program setup. But in this report no variation in the different costs has been made. The following flocculation, sedimentation, and filtration units have been analyzed (TABLES 10-1, 10-2 and 10-3).

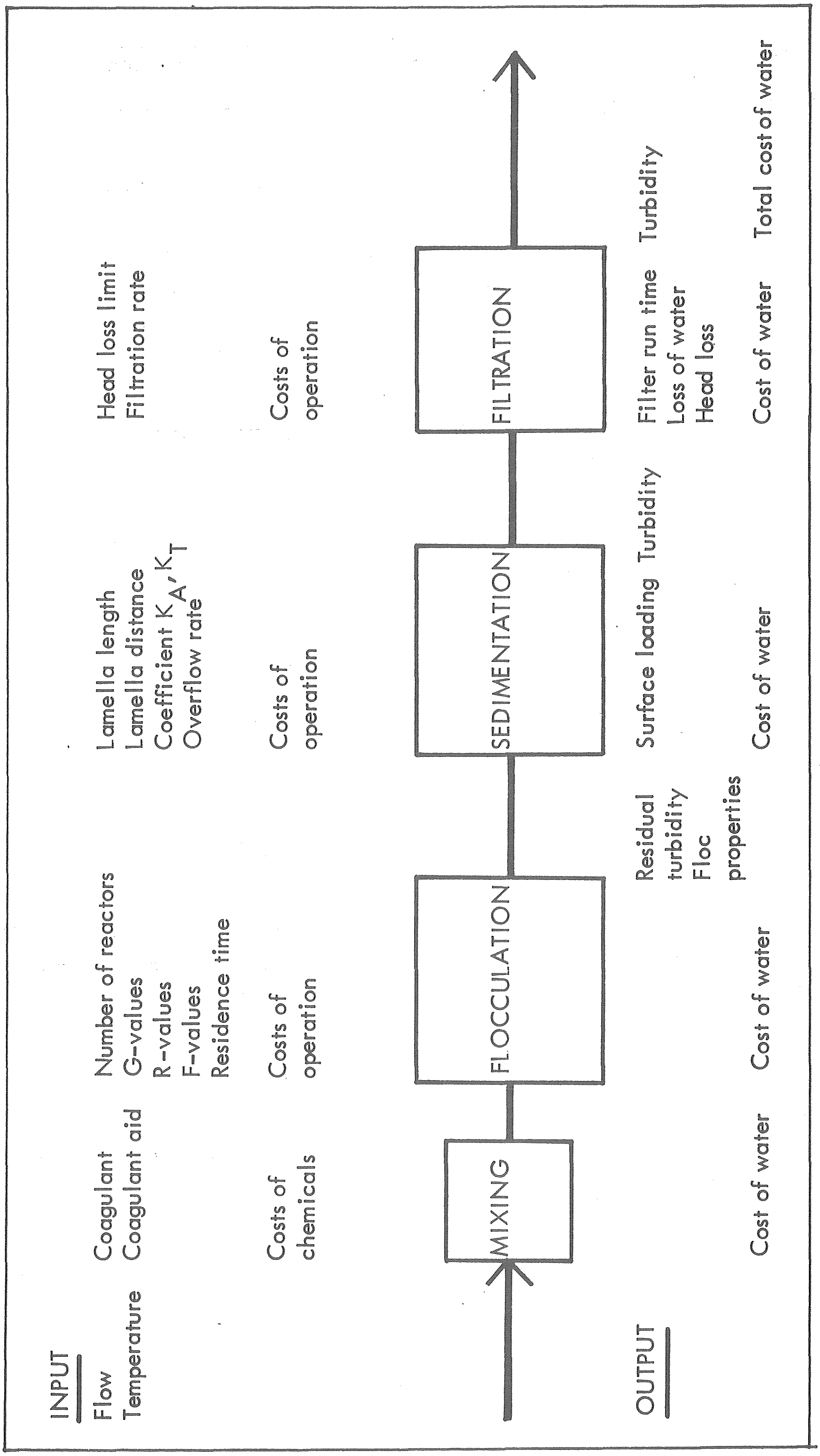


FIG. 10-1 The water treatment plant model

TABLE 10-1 Flocculation units (see Eq.3-3)

System	Number of reactors	G-values s^{-1}	R-values	F-values
1	2	75,15	0.5,0.5	1,1
2	2	75,15	0.5,0.5	2,1
3	3	100,50,10	0.33,0.33,0.33	1,1,1
4	3	300,75,10	0.167,0.167,0.666	1,1,1
5	3	100,75,10	0.33,0.33,0.33	2,2,1
6	4	125,75,15,5	0.25,0.25,0.25 0.25	1,1,1,1
7	4	400,100,255	0.1,0.1,0.4,0.4	1,1,1,1
8	4	125,75,15,5	0.25,0.25,0.25, 0.25	2,2,1,1
9	5	125,75,25 10,5	0.2,0.2,0.2 0.2,0.2,0.2	1,1,1 1,1
10	5	500,150,50 15,5	0.07,0.07,0.3 0.3,0.3	1,1,1 1,1
11	5	125,75,25 10,5	0.2,0.2,0.2 0.2,0.2	2,2,1 1,1
12	6	125,75,50 25,10,5	0.16,0.16,0.16 0.16,0.16,0.16	1,1,1 1,1,1
13	6	600,300,150 75,25,5	0.055,0.055,0,22 0.22,0.22,0.22	1,1,1 1,1,1

Note

Detention times in the flocculation units: 15, 30, 45, 60, and 15 min.

G-values have been chosen mainly based on results from tests in pilot plants. (A theoretical optimization of the flocculation performance model has not been made.)

R-values have been chosen to obtain proper design.

F-values have in most cases been chosen to be 1. An F-value of 2 indicates that there is a baffle in the reactor.

TABLE 10-2 Sedimentation units (see Eq. 5-6 and Eq. 5-8)

Lamella sedimentation units		
Lamella length	m	2.0, 3.0
Lamella distance	m	0.03, 0.04, 0.06, 0.08, 0.09
Lamella width	m	1.0 to 1.25
Inclination angle	α	55
Coefficient K_A		0.65, 0.80
Overflow rate	m/h	1.00, 1.50, 1.75, 1.80, 2.00, 2,10, 2.25, 2.50
Lovö basin		
Overflow rate		1.00, 1.50, 1.75, 1.80, 2.00, 2.10, 2.25, 2.50

TABLE 10-3 Filter units

Sand filter (filter 1)
Sand-anthracite filter, coarse media (filter 5)
Sand-anthracite filter, fine media (filter 6)
Activated carbon filter (filter 7)
Sand-activated carbon filter (filter 8)
Garnet-sand-anthracite filter (filter 9)
Filtration rate: 10, 15, 20 m/h Head loss a) Quality limit (15 % Std. 2 Sigrist) b) Total head loss limit (1.5 and 1.75 m of water or quality limit)

10.2 Flocculation

10.2.1 Costs

A presentation of the investment cost for the different flocculation units is made in FIG. 10-2 in order to show the magnitude of the cost of the reactor volume used in this investigation.

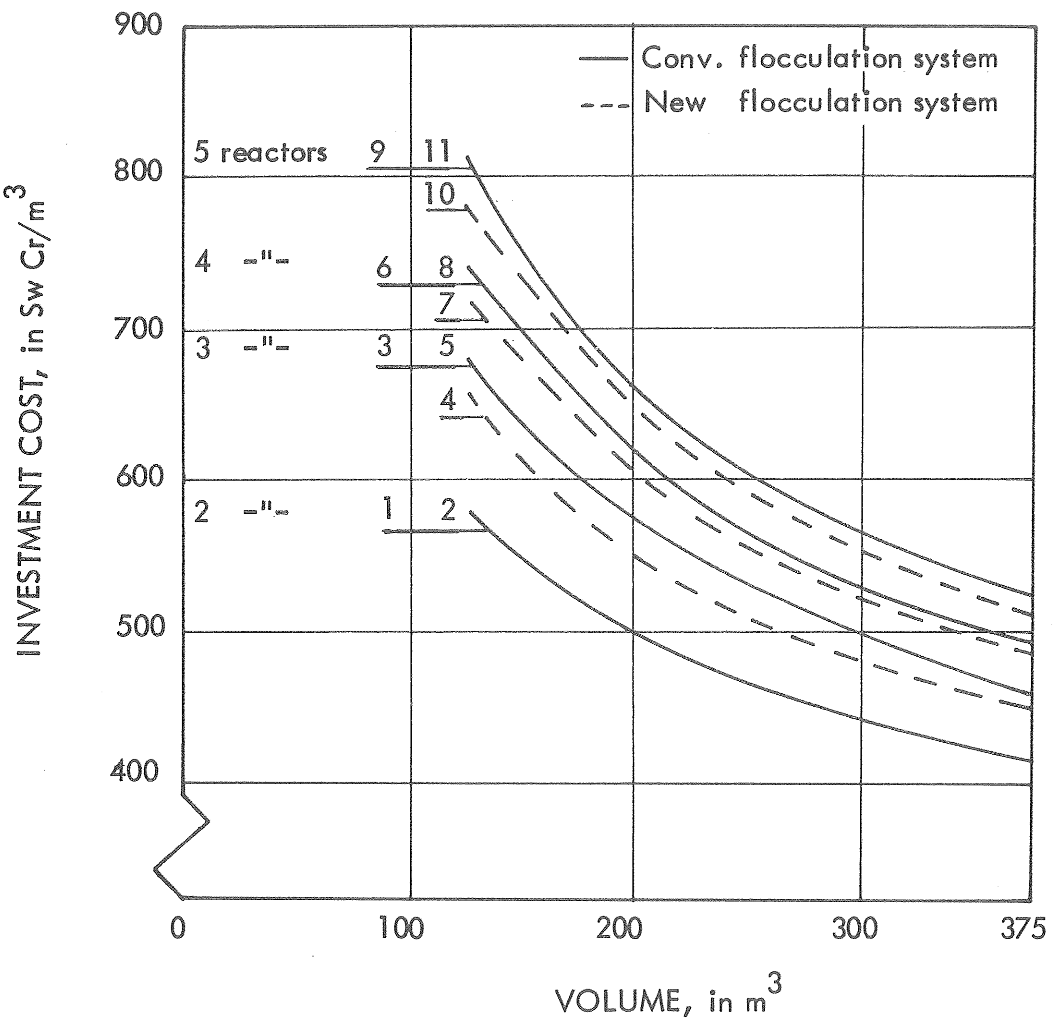


FIG. 10-2 Investment cost expressed in SW.cr./m^3 of basin volume for the flocculation units as a function of the basin volume. (The number refers to TABLE 10-1)

FIGURE 10-2 shows that the investment cost increases with decreasing reactor volume and with increasing number of reactors in series. It can also be seen that the new flocculation system with a shorter detention time in the first two reactors than in the following ones is somewhat cheaper than the conventional system.

The greatest part of the cost of the flocculation unit is contributed by the concrete basin followed by the cost of the stirring equipment and the roof of the building.

The maintenance cost for the flocculation operation has been estimated at 10-20 per cent of the total maintenance cost of the plant studied.

10.2.2 Results.

It is not correct to consider a flocculation unit separately without considering the subsequent unit operations, as the result obtained in the flocculation unit affects the overall treatment process. It has been observed that especially the filter operation is important in the minimization of the cost of treatment. Important factors for the filter operation are the permissible head loss and the price of the backwash water. Therefore the performance of the flocculation units presented in TABLE 10-1 have been examined under the following conditions:

Case 1. Permissible head loss in the filters considering a certain water quality limit. High price of the filter backwash water.

Case 2. Permissible head loss in the filters considering a certain water quality limit. Low price of the filter backwash water.

Case 3. Permissible head loss in the filter limited to 1.5 m. High price of the filter backwash water.

Case 4. Permissible head loss in the filters limited to 1.5 m. Low price of the filter backwash water.

In the analyses of the cases listed above the overflow rate in the sedimentation unit has been constant (1.0 m/h).

Case 1.

All the flocculation systems have been analyzed at a water temperature of 5^{o} C and a dosage of silica of 4 mg/l.

In FIG. 10-3 and FIG. 10-4 two different principles of flocculation units are shown. The total cost of the water treatment in öre per cubic meter is plotted against the residence time in the flocculation unit. Two different filters (filter 1 and filter 9) with a filtration rate of 10 m/h have been chosen to show the result of this analysis.

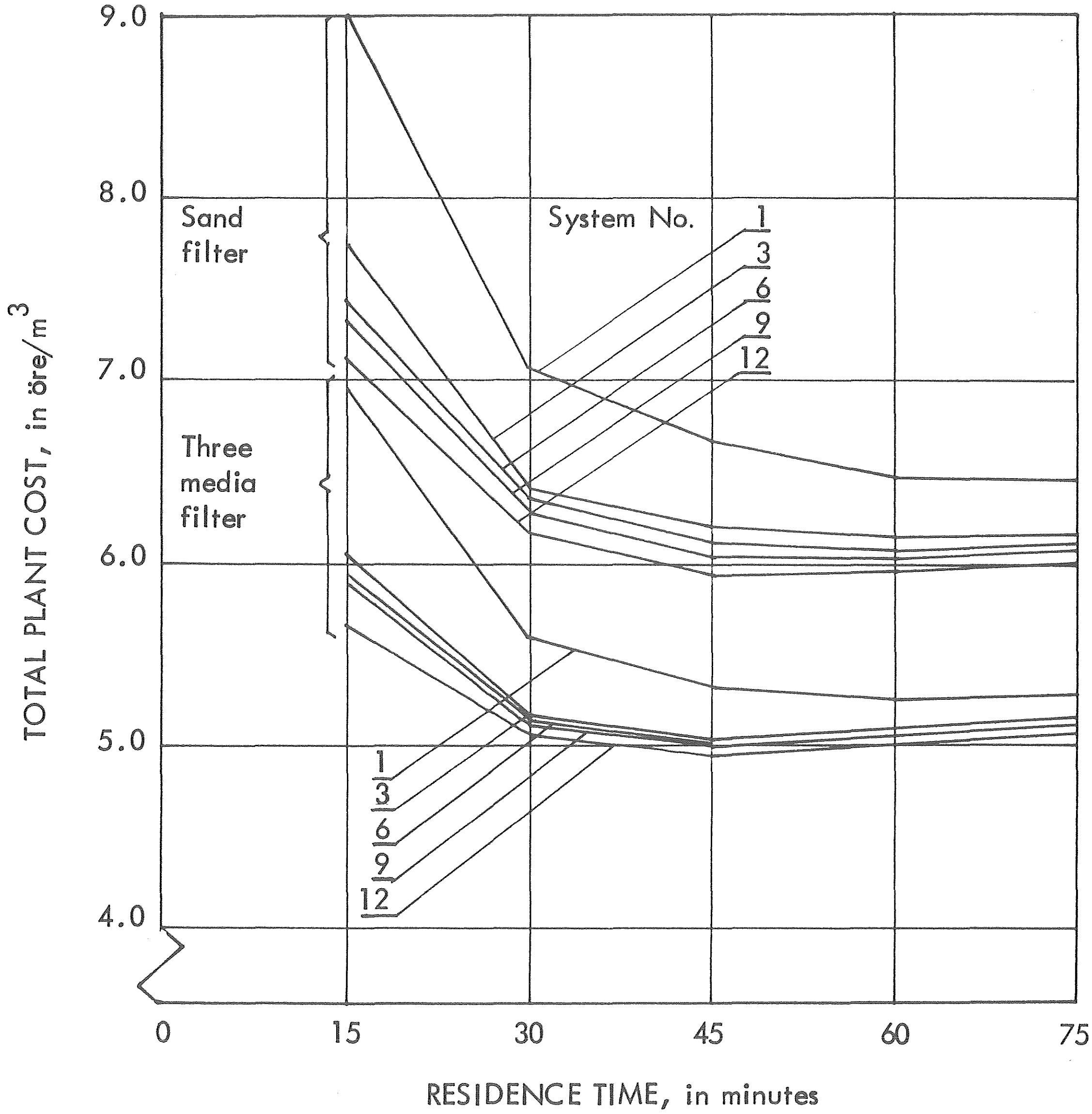


FIG. 10-3 The comparative cost of the plant as a function of the residence time in the conventional flocculation units, systems No. 1,3,6,9, and 12 in TABLE 10-1.

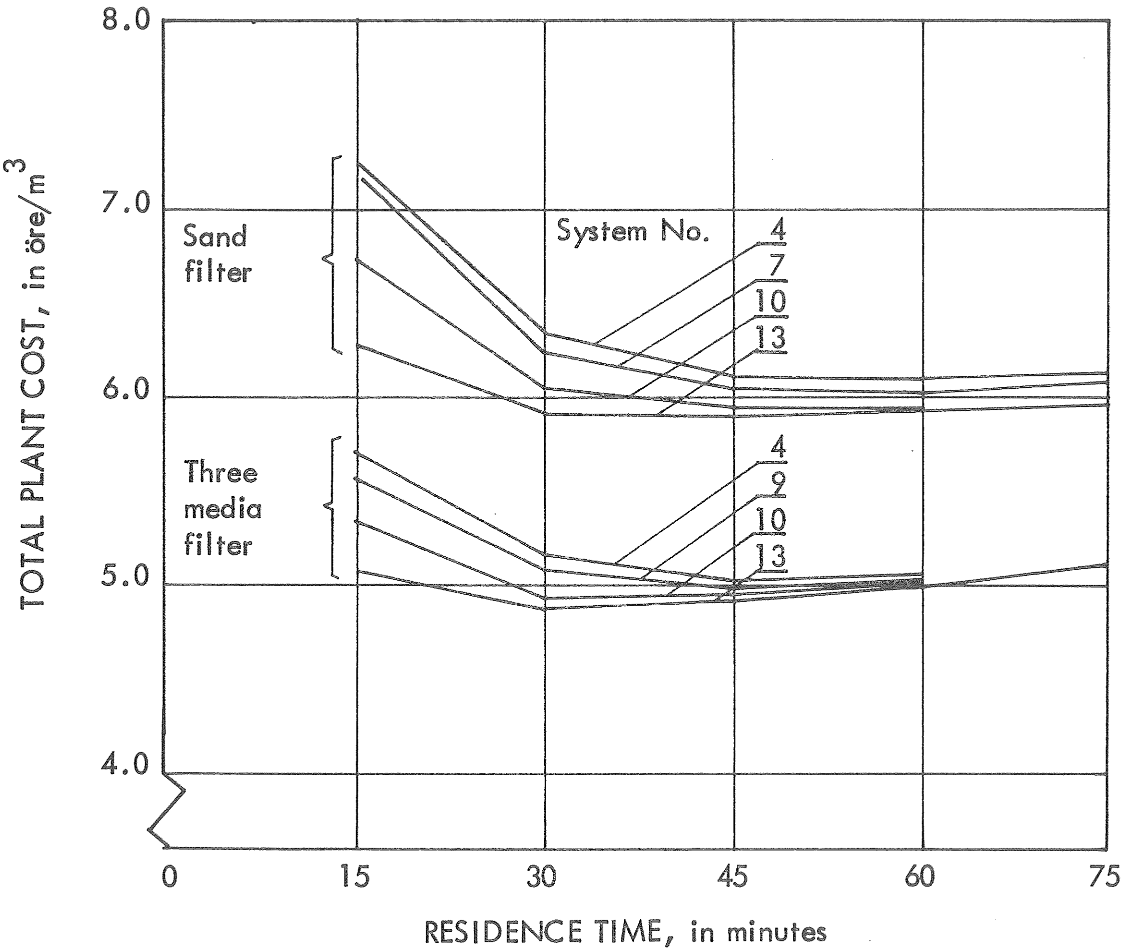


FIG. 10-4 The comparative cost of the plant as a function of the residence time in the new flocculation units, systems No. 4, 7, 10, and 13 in TABLE 10-1

FIGURE 10-3 and FIG. 10-4 show that there is no great difference in the minimum cost in öre/m^3 of water for the two flocculation systems. The water cost is slightly lower for the new flocculation system than for the conventional one. The detention time at the minimum cost is shorter in the new flocculation system than in the conventional one. The following comparison between the systems can be made, TABLE 10-4.

TABLE 10-4 Comparison of flocculation systems

	Detention time in minutes at minimum cost for each system.	
	Sand filter	Three media filter
Conv. system No. 1	75	60
3	75	40
6	55	40
9	50	40
12	45	40
New system No. 4	60	45
7	60	40
10	50	35
13	40	35

TABLE 10-4 shows a shorter optimum detention time for the new system compared to that of a conventional system.

Case 2.

In this case where the price of the backwash water is five times lower than that in Case 1, it is of less importance to have a low turbidity of the filter influent as the running costs of the filter operation are markedly reduced. The detention time in the flocculation unit at minimum cost is shorter than in the previous case as a consequence of the low price of the backwash water.

The optimum detention time for the conventional flocculation system (No. 12) is about 40 min. instead of 45 min. as in Case 1 and the optimum detention time for the new system (No. 10) is about 30 min. instead of 50 min. The total cost of the treatment plant, expressed in öre/m^3 of water is of course lower than in Case 1. The result is shown in FIG. 10-5.

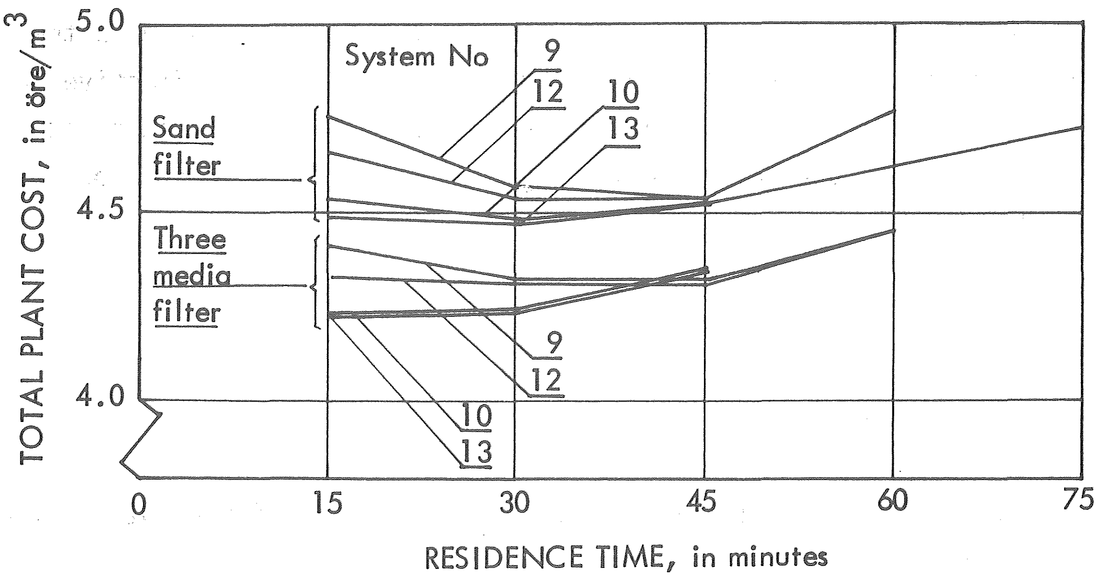


FIG. 10-5 The comparative cost of the plant as a function of the residence time in the flocculation units, systems No. 9,10,12, and 13 in TABLE 10-1.

Case 3.

In this case, where the permissible head loss is limited, one must consider the filtration rate since the filter run time can be very short at high filtration rates, especially for filters with a high initial head loss. The short run time in turn increases the loss of water, and as the price of the backwash water is high, the water cost becomes high. For low filtration rates, about 10 m/h, this case is very similar to Case 1.

Case 4.

The same is true for the filtration rate in this case as in the previous one. At filtration rates of about 10 m/h, this case is very similar to Case 2.

From the above,one can conclude that the filter operation has a heavy influence on the design of the flocculation unit.

The suggested new flocculation unit principle seems to be less expensive than that of a conventional unit.

Installation of a baffle in the first reactors of the flocculation unit does not increase the cost of the unit but according to the theory, a compartmentalization by means of baffles is effective in reducing the turbidity. As a consequence of this the running cost of the filter operation and also the cost of the sedimentation are reduced.

10.3 Sedimentation

10.3.1 Costs

FIGURE 10-6 presents the investment cost for the different lamella sedimentation units.

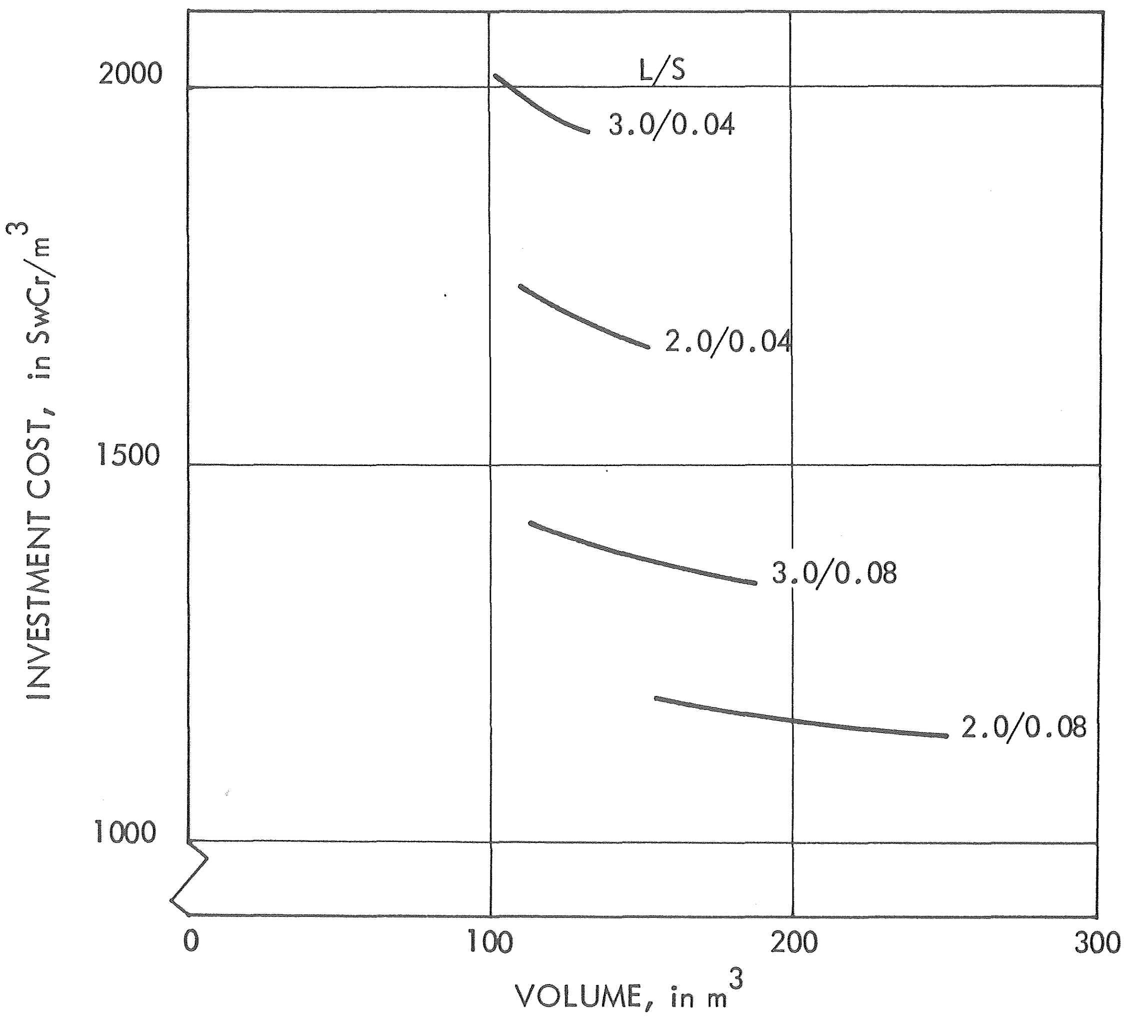


FIG. 10-6 Investment cost of the basin, expressed in Sw.cr./m^3 of basin volume for some lamella sedimentation units, as a function of basin volume. (B = 1.0 to 1.25 m.)

The greatest part of the cost of the lamella unit is contributed by the distribution channels for the influent and effluent, followed by the cost of the lamella plates. The material in the channels and the plates is stainless steel. The cost of the concrete basin is of less importance. In the conventional basin, however, the cost of the basin is the greatest cost contribution.

The maintenance cost for the lamella sedimentation operation has been estimated to about 10 % of the total maintenance cost of the plant studied. For the conventional sedimentation operation the corresponding value is about 25 %. The high maintenance cost of the conventional basin is mainly due to the sludge handling.

10.3.2 Results

Several different sedimentation units have been analyzed. In the study of the sedimentation operation a conventional flocculation system (system No. 6 in TABLE 10-1) has been used for production of flocs. The lamella sedimentation model has been used to calculate the breakup curves shown in FIG. 10-7.

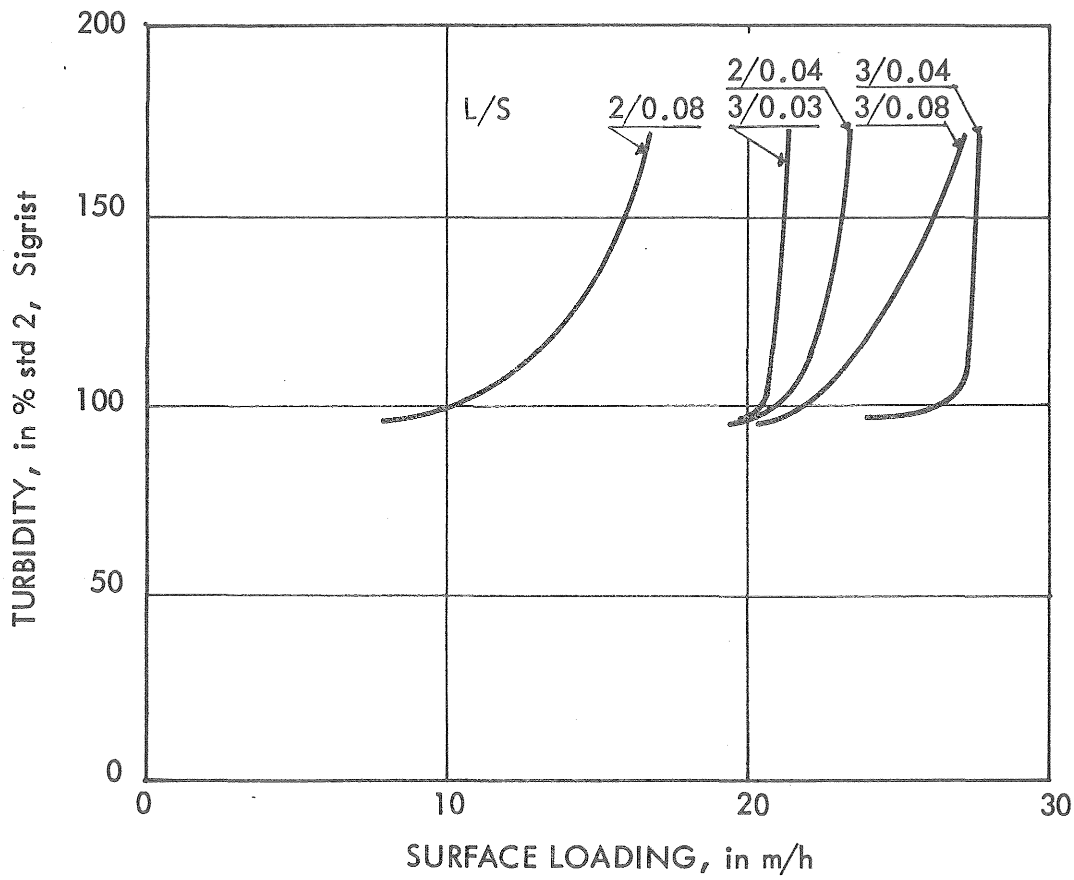


FIG. 10-7 Turbidity in effluent as a function of surface loading of the lamella sedimentation units. Temperature 5°C, activated silica dosage 4 mg/l.

The highest surface loading was reached in a lamella unit with a lamella length of 3.0 m and a lamella distance of 0.04 m. This corresponds to an L/S-ratio of 75. The surface loading increases with increasing L/S-ratio according to Eq. (4-17).

However, a further increase of this ratio to 100 by a change in the lamella distance (L=3.0 S=0.03) resulted in a decreased surface loading, which is in agreement with results obtained in pilot plants. The horizontal distance between the lamellae is too small and the sludge is eroded.

The highest surface loading does not, however, mean the minimum cost. A calculation of the operation cost of the units expressed in öre/m^3 of water at the breakup point is shown in FIG. 10-8.

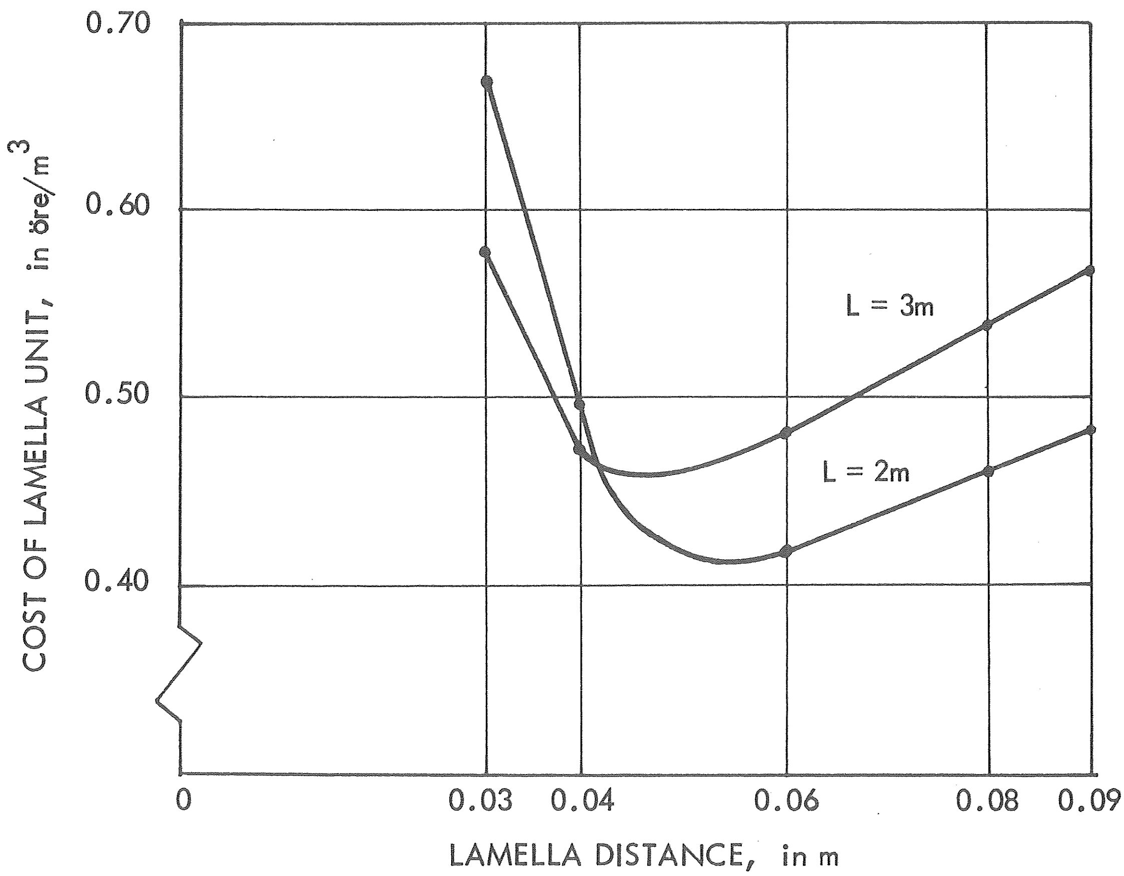


FIG. 10-8. The comparative costs of the lamella units. Investments and running costs, as a function of the horizontal distance between the lamellæ.

The minimum cost of a lamella unit with a lamella length of 3.0 m is lower than that of a unit with a lamella length of 2.0 m. The most economical horizontal distance is between 0.05 m and 0.06 m.

The type of lamella sedimentation unit examined in this investigation has been designed in order to simplify the construction and at the same time obtain satisfactory performance (see FIG. 4-5). The inlet and outlet system is, however, of great importance for the utilization of the entire lamella plate for separation of the flocs. This investigation shows that it can be economical to improve the inlet and outlet system. For example, in a lamella unit with a lamella length of 3.0 m and a lamella distance of 0.08 m, the cost of the inlet and outlet system is allowed to increase from 10 % of the price of the lamella plate to about 40 % provided that this increase in cost leads to an increase of the performance of about 20 %.

In a lamella unit with a very small, less than 0.05 m, lamella distance, the breakup curve is very steep. For this reason a horizontal distance between the lamella of between 0.06 m and 0.08 m ought to be used in practice. This means that for a lamella unit with a lamella length of 3.0 m, a surface loading between 18 and 24 m/h can be reached. This value is valid when a conventional flocculation system is used and the water temperature is low.

The result of the analysis of the Lovö basin is shown in FIG. 10-9.

The investment cost of a Lovö basin is about 40 per cent higher than that of a lamella unit with a lamella length of 3.0 m and a lamella distance of 0.06 m. The maintenance cost of a Lovö basin is also higher than that of a lamella unit. Based on the costs given in this report, one can calculate that the running costs of a Lovö basin are about 80 per cent higher than those of a well-designed lamella unit.

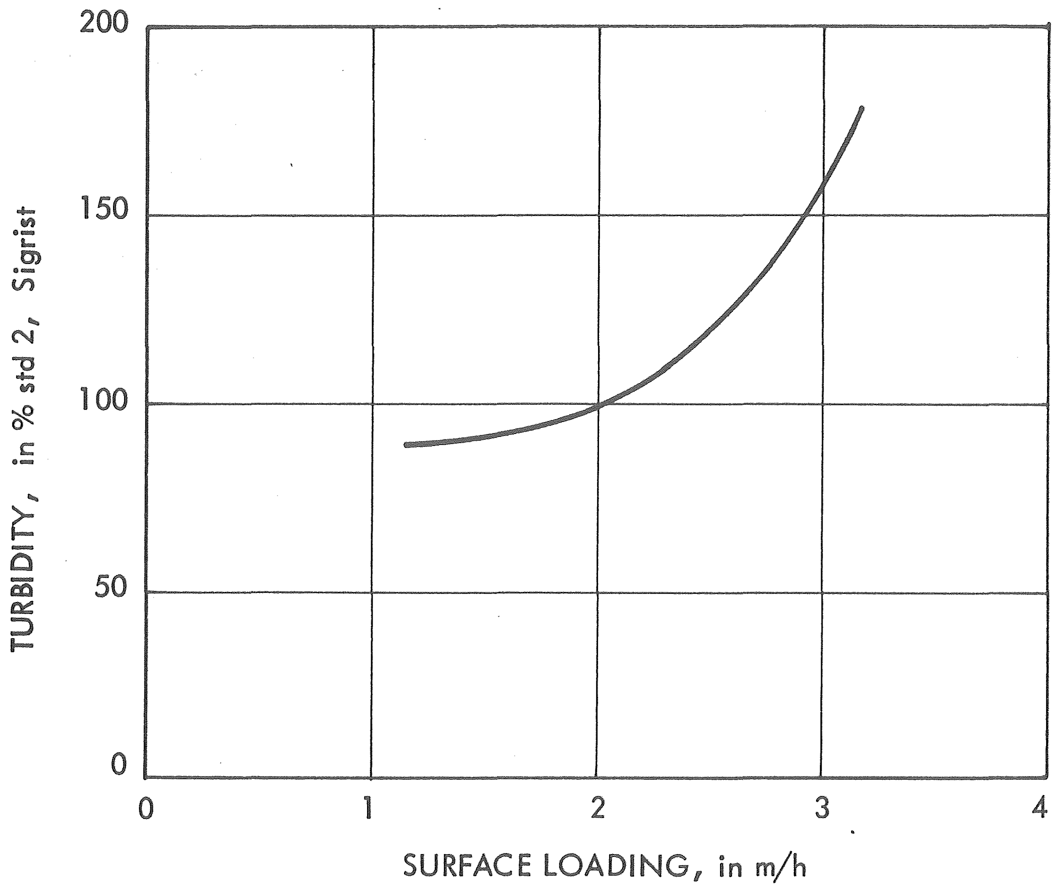


FIG. 10-9 Turbidity in effluent as a function of surface loading of the Lovö basin. Temperature 5°C, activated silica dosage 4 mg/l.

10.4 Filtration

10.4.1 Costs

The investment cost for the filter units varies not only with the surface loading as for the previous unit operations but also with the permissible head loss of the filter. This head loss can be chosen on the basis of the strength of the suspended particles in relation to the size of the filter media. The permissible head loss varies with temperature, particle concentration, and dosage of activated silica, which has to be considered when determining the available head for the filter in question. A comparison of the investment cost for the different filters is shown in TABLE 10-5. The total height of the various filter designs is different depending on the penetration depth and the head loss in the bed (Eq. 7-23). The depth of the filter bed is constant.

TABLE 10-5 Investment cost of the filter units.

Filter design	Investment cost, in Sw.cr./m^3 basin volume		
	Filtration rate, in m/h		
	10	15	20
Sand filter	900	1000	1200
Sand-anthracite (coarse medium)	1300	1400	1500
Sand-anthracite (fine medium)	1050	1200	1300
Activated carbon	1900	2000	2100
Sand-act. carbon	1800	1900	2000
Garnet-sand-anthracite	1050	1150	1250

The maintenance cost for the filter operation has been estimated to be about 10 per cent of the total maintenance cost. The running costs for the filter operation are heavily influenced by the price of the backwash water. The backwash water can either be treated at the plant and recycled to the intake, or it can be discharged to a sewage treatment plant. In the latter case the price of the backwashwater may be relatively high in comparison with the price of internal treatment, which may be very simple. In this study the influence of different prices of the backwash water on the cost of the treatment has been examined (1 Sw.cr. per cubic meter of backwash water and 0.2 Sw.cr. per cubic meter of water).

10.4.2 Result

First, all the filters have been analyzed at a water temperature of 5°C, varying particle concentrations, and different dosages of an activated silica. The dosage of activated silica added in the flocculation unit has an effect on both the permissible head loss (Eq. 7-11) and the increase in head loss per unit of time (Eq. 7-22). A variation of the dosage of activated silica of course course affects the water cost. It is therefore necessary to consider these facts when comparing the filters.

In the analysis of the filters it is necessary to separate the following cases.

Case 1. Permissible head loss considering a certain water quality limit. High price of the backwash water.

Case 2. Permissible head loss considering a certain water quality limit. Low price of the backwash water.

Case 3. Permissible head loss limited to 1.5 m of water or considering a water quality limit. High price of the backwash water.

Case 4. Permissible head loss limited to 1.5 m of water or considering a water quality limit. Low price of the backwash water.

The four cases have been studied at the filtration rates of 10,15, and 20 m/h. The flocculation unit used in this analysis has been a conventional system (System No. 6 in TABLE 10-1). The overflow rate in the sedimentation units - Lamella unit and Lovö basin - - has varied.

Case 1.

In this case where the permissible head loss may reach the limit set by the water quality only, the filtration rate has very little influence on the water cost. However, the water cost decreases with the filtration rate, and the maximum filtration rate has to be limited by the fact that the water quality deteriorates when the filtration rate exceeds a certain value. The optimum dosage of activated silica is dependent on the filter design and particle concentration. Three different filters have been selected to illustrate the relationship between the costs of chemicals and the filter operation and the dosage of activated silica (FIG. 10-10).

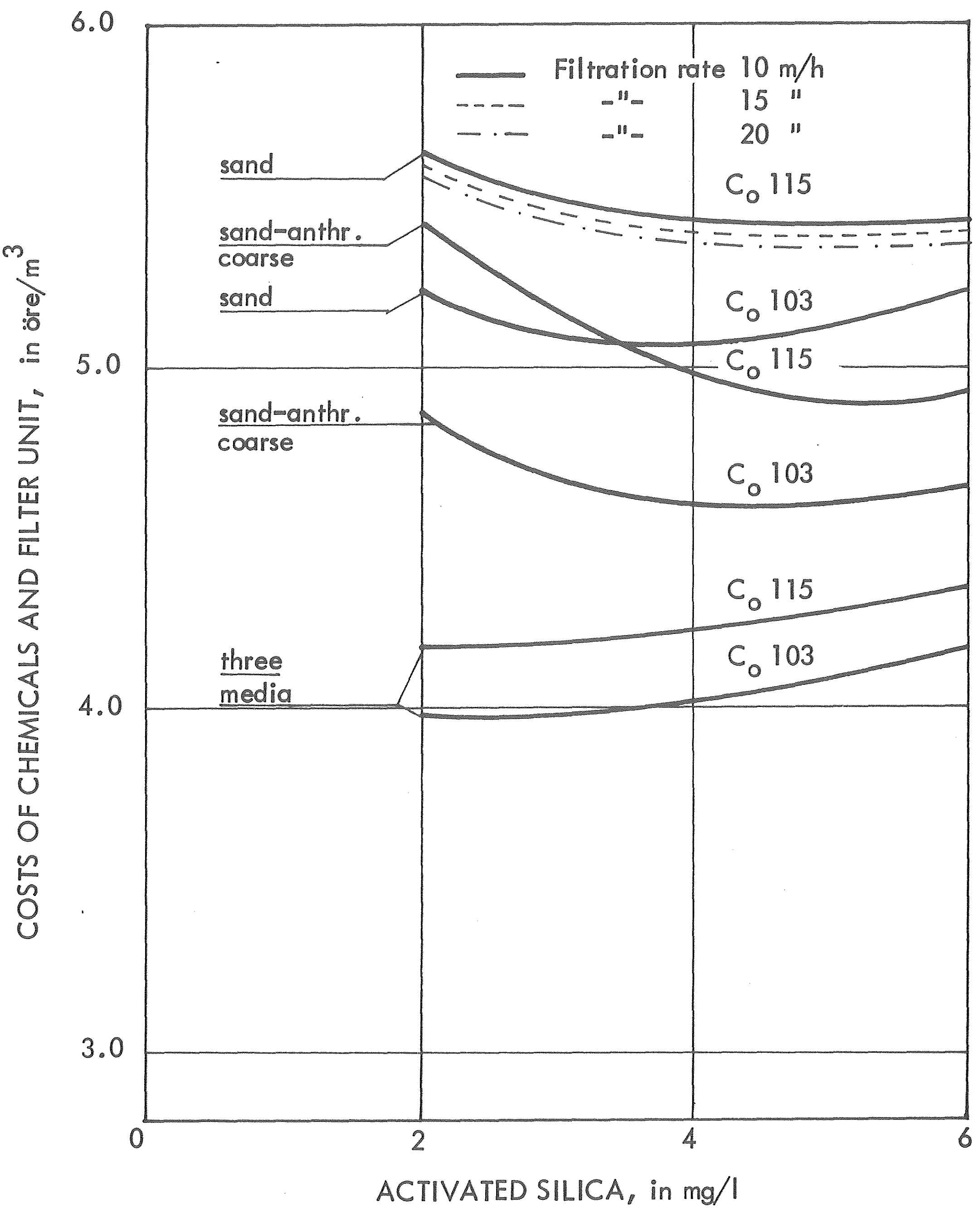


FIG. 10-10. Costs of chemicals and filter operation as a function of activated silica addition. Temperature 5°C.

FIGURE 10-10 shows that the optimum dosage of silica is about 4 mg/l for a conventional sand filter. For a coarse filter such as a sand-anthracite filter, a somewhat higher dosage of silica may be used, especially when the particle concentration is high. The optimum dosage of activated silica is less or about 2 mg/l for a three-media filter, which is characterized by a high ability to resist breakthrough.

For this case it can be concluded that the filtration rate must be limited by quality reasons only. The most economical filter design is a filter characterized by high storage capacity of suspended particles together with an ability to resist a filter breakthrough. Examples of such filters are the three-media filter (filter No. 9) and also a two-media filter with a bed of sand and anthracite of a relatively fine grain size (filter No. 6).

Case 2.

This differs from Case 1 only by a lower price of the backwash water. In FIG. 10-11 the three filters are presented in the same manner as in FIG. 10-10.

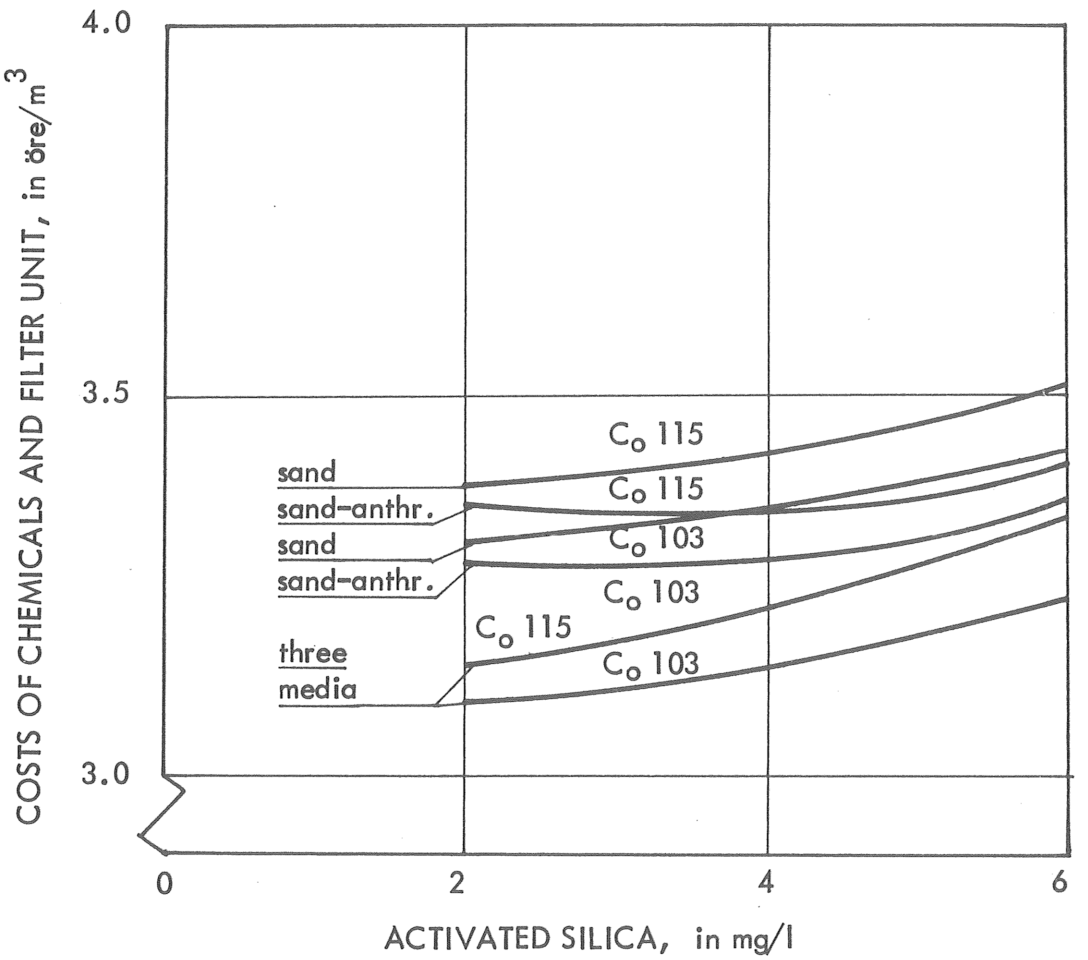


FIG. 10-11. Costs of chemicals and filter operation as a function of activated silica addition. Temperature 5°C.

FIGURE 10-11 shows that the optimum dosage of activated silica is lower in this case than in Case 1. The filtration rate may also in this case be chosen as high as possible but of course considering the water quality. A low dosage of activated silica, however, limits the possibility of using a high filtration rate (see FIG. 7-9).

Case 3.

In this case the maximum available head loss is chosen to be 1.5 m as an example. If a filter reaches its maximum head loss considering the water quality (Eq. 7-11) before the maximum available head is reached, the former head loss determines the filter run time.

The available head should be based on technical and economic considerations. In practice, values between one and two meters are common. As in the previous cases, some filters have been selected to show the costs of chemicals and filter operation as a function of silica addition (FIG. 10-12). In contrast to the previous cases, the filtration rate is of importance in this case; therefore, the influence on the costs caused by the filtration rates, 10 m/h and 15 m/h, has been shown.

FIGURE 10-12 is difficult to interpret but some conclusions may be drawn. Filter designs characterized by an ability to resist a breakthrough have a lower optimum dosage of silica than filters which do not have this ability. For filters with fine grain size or filters with two or three media which mix at the interfaces, the water cost increases rapidly with filtration rate due to a high initial head loss. In FIG. 10-12 the turbidity of the influent to the filter is constant (about 100 % Std. 2). In practice, the normal situation is that the turbidity of the influent increases with decreasing addition of silica. It is therefore necessary to consider the "cost of turbidity" and not only the cost of the filter operation when trying to find the minimum cost of the entire water treatment process. Such an analysis will be shown later on. In the following figure, FIG. 10-13, the costs of chemicals and filter operation have been plotted against the filtration rate in order to show the differences between the filters at the optimum dosages of activated silica.

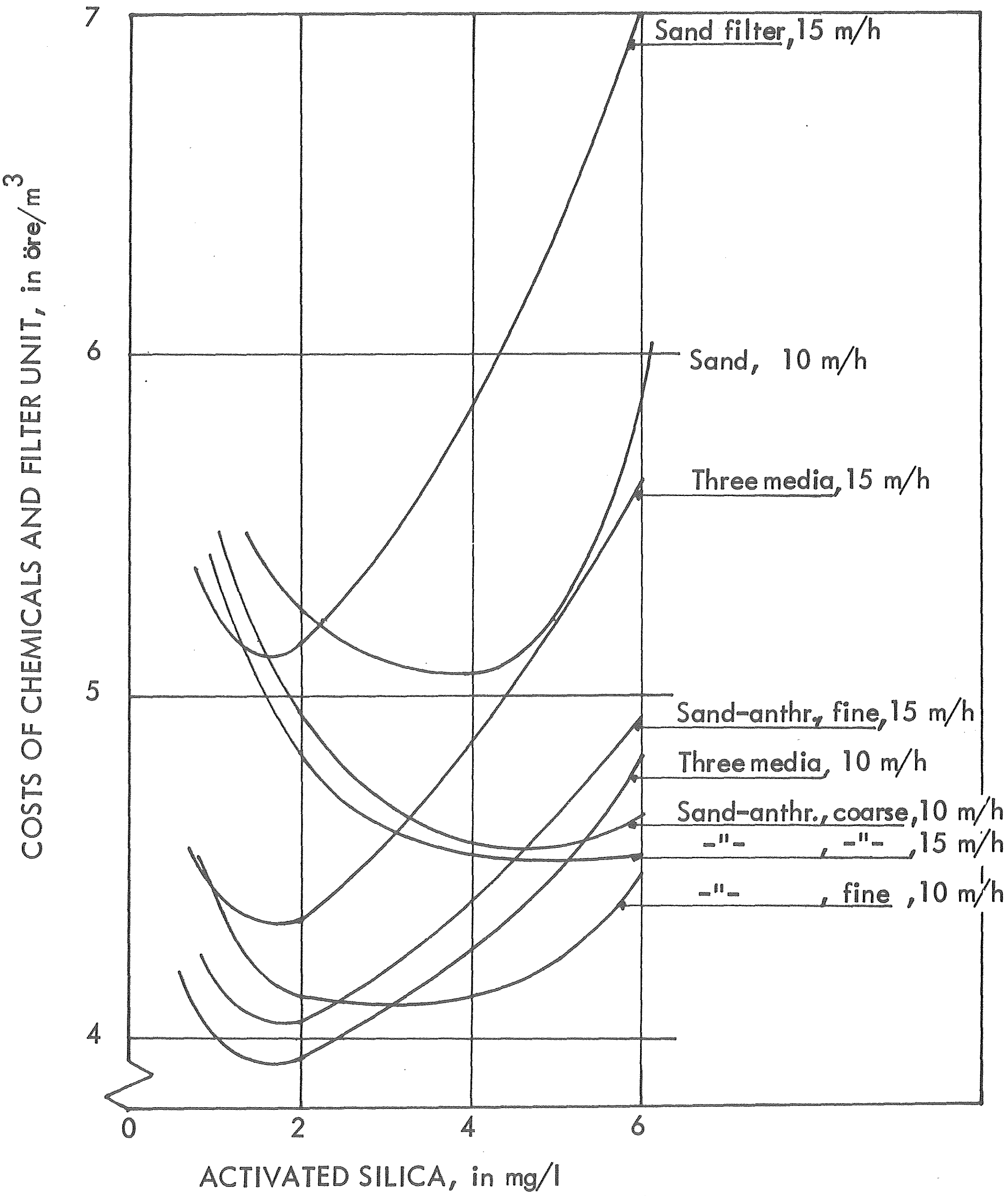


FIG. 10-12. Costs of chemicals and filter operation as a function of activated silica addition. Temperature 5°C.

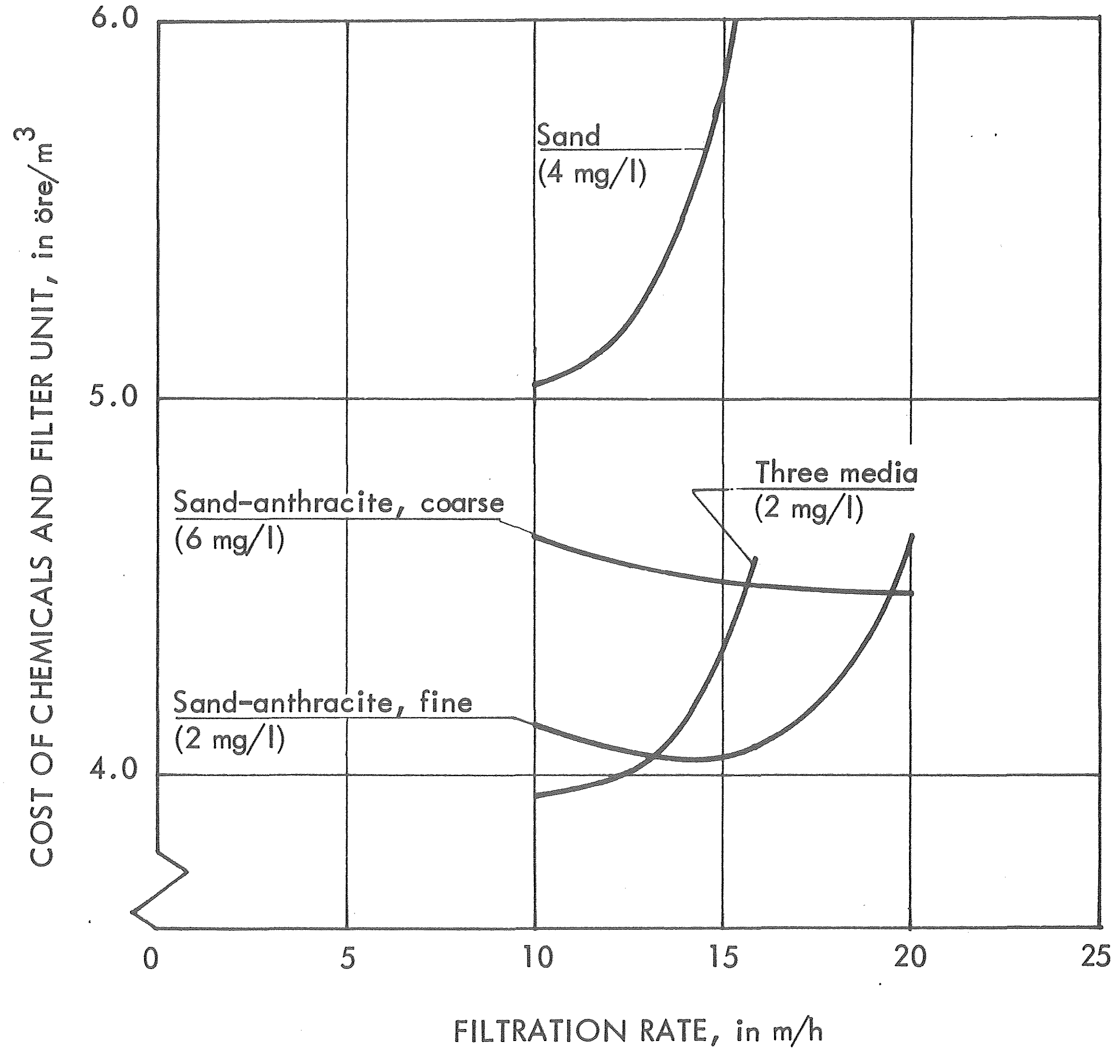


FIG. 10-13. Costs of chemicals and filter operation as a function of filtration rate.

If there is no other head loss limit than that due to water quality, it is always advantageous to have a high filtration rate. But as it has been found that there is an upper limit of the filtration rate which can be accepted with respect to water quality, the maximum filtration rate may be between 15 m/h and 20 m/h. This filtration rate is dependent on the dosage of silica and probably also on temperature and filter design. As shown in FIG. 10-13 the costs increase very rapidly with filtration rate, except for the coarse sand – anthracite filter, due to the influence of the initial head loss of the filters, which shortens the filter run time when the available head is limited to 1.5 m.

From this case one can conclude that it is important to have the right silica addition for each filter and that the available head is also an important factor to consider when designing a filter unit.

Case 4.

This case differs from the previous one in the price of backwash water. Concerning the dosage of activated silica, the relationship between Case 3 and 4 is the same as the one between Case 1 and 2, which means that the optimum dosages of silica are relatively low.

As shown in the different cases, several factors affect the filter operation; it is therefore necessary to analyze each case separately under the prevailing conditions.

The activated carbon filter has not been discussed, but the main reason for using such a filter is to reduce taste and odor in the water. This filter is more expensive than the other ones, e.g. twice as expensive as a conventional sand filter. However, the investment cost is not the most important cost, especially not when the price of the backwash water is high. In the investigation it has also been observed that the activated carbon filter has a high ability to resist a breakthrough of suspended particles even when the filter material is coarser than the sand in a conventional filter. Thus, one can conclude that it is possible to use a relatively coarse activated carbon and this means that the filter run time can be longer than that of a sand filter. This in turn means low running costs due to the low consumption of backwash water. In fact, in some circumstances the activated carbon filter proves to be more economical than a sand filter, and in addition the water quality is improved.

10.5 Combinations of flocculation, sedimentation, and filtration units.

In the previous sections information about the single unit operations has been given. It is, however, essential to investigate their interrelationships during different circumstances. By use of the water treatment plant model, one can calculate and predict the result of any combination of unit operations.

With the guidance of the results obtained some of the most interesting combinations of the unit operations have been selected in order to analyze the entire water treatment process.

Flow:	500 m^3/h
Temperature:	5,10,15^{0}C
Chemicals:	Aluminium sulfate: 40 mg/l Activated silica: 0,2,4,6 mg/l Lime to optimum pH-value
Flocculation system:	Conventional system No. 12 New system No. 10 Residence time in the flocculation units: 30, 45, 60 minutes
Sedimentation units:	Lamella unit Lamella length 3.0 m Lamella distance 0.08 m Lamella width is automatically calculated (B = 1.00 to 1.25 m) Inclination angle 55^0 Overflow rate: 1.25, 1.50, 1.75, 2.0 m/h Lovö basin Length and width are automatically calculated Overflow rate: 1.25, 1.50, 1.75, 2.0 m/h
Filter units:	Sand filter (filter 1) Sand-anthracite filter, (filter 5) Sand-anthracite filter, (filter 6) Three-media filter (filter 9) Filtration rate: 5, 10, 15, 20 m/h

Head loss limit: 1.75 m or a turbidity value of max. 15 % Std. 2

Price of the backwash water: 1.00 Sw.cr./m^3

The given data represent about 500 water treatment plants, analyzed at three different temperatures and four different dosages of activated silica.

The analysis shows that there are, for each flocculation, sedimentation, and filter system, many design possibilities at which the total cost is almost equal. In other words, the minimum cost curve, including capital and maintenance costs is very flat, which means that the cost of the flocculation operation balances the cost of the filtration operation, and the relative level of the costs is given by the cost of the sedimentation operation. This is, of course, a simplification of the situation, but in general one can choose a relatively short detention time in the flocculation unit, which means low cost and a high turbidity in the influent to the filter and a high running cost of the filter operation as a consequence. On the other hand, one can choose a long detention time in the flocculation unit with a low running cost for the filter operation. The filter operation, under the given conditions, has proved to be the most important unit operation. Thus, it is extremely important to pay attention to this operation in the choice of filters, permissible head loss, filtration rate, and backwashing techniques. Different filters have, as shown, different characteristics and the operational conditions have to be considered in relation to the flocculation operation mainly, but also in relation to the sedimentation operation. In general, a long residence time in the flocculation unit and a low surface loading in the sedimentation unit mean that the filtration rate can be higher than at a short residence time and a high surface loading.

In comparing the conventional flocculation system with the new suggested flocculation system one can state that it is possible to choose the detention times within a broader interval and also use shorter detention times with the new system than with the conventional one. This fact is due to the high velocity gradients used in the first reactors of the new system, causing a rapid reduction in the number of particles. This new system also offers great advantages when one wishes to increase the capacity of an already existing plant or when the price of land or other costs related to the plant size are high.

The shape of the breakup curve has to be considered when the surface loading of the sedimentation unit is chosen. The breakup curve is, as shown, dependent on the design of the unit. In general, the surface loading ought to be chosen as close as possible to the point on the breakup curve where the turbidity starts to rise. From that point the slope of the curve is relatively steep, especially for lamella sedimentation units with a high value of the L/S-ratio. It is more difficult to define or to choose the right surface loading for a Lovö-basin for which unit the breakup curve is less steep.

As a conclusion of the analysis of the unit operations and their interrelationships TABLE 10-6 summerizes the parameter intervals within which the most economical water treatment plants are to be found.

TABLE 10-6 Design data of the unit operation at minimum water cost

Filters	Conventional flocculation and lamella sedimentation[x] 1	5	6	9	New flocculation and lamella sedimentation[x] 1	5	6	9
Activated silica dosage in mg/l at								
5°C	4	4-5	4	2-4	4	4-5	4	2-4
10°C	2	2	2	<2	2	2	2	<2
15°C	2	2	2	<2	2	2	2	<2
Flocculation Residence time min.	45-60	45-60	40-60	40-50	30-60	40-60	30-50	30-60
Sedimentation Overflow rate m/h	1.6-1.8	1.5-1.6	1.6-1.8	1.6-1.8	1.6-1.8	1.5-1.6	1.6-1.8	1.6-1.8
Surface loading of a lamella unit m/h (L=3.0, S=0.06)	26	24	26	26	26	24	26	26
Filtration Filtration rate m/h	8-10	20	10-15	10-15	8-10	20	10-15	10-15
Filter run time h	20-30	16	30-50	25-50	20-30		30-50	25-80
Comparative cost of treatment in öre/m3 of water								
Investment floccul.	0.37	0.37	0.35	0.35	0.28	0.33	0.28	0.28
sed.	0.40	0.44	0.40	0.40	0.40	0.44	0.40	0.40
filtr.	0.23	0.17	0.24	0.25	0.23	0.17	0.24	0.24
Running floccul.	0.15	0.15	0.15	0.15	0.15	0.15	0.15	0.15
sed.	0.07	0.07	0.07	0.07	0.07	0.07	0.07	0.07
filtr.	1.85	1.26	0.85	0.90	2.05	1.28	1.05	1.03
Chemicals	2.57	2.57	2.57	2.57	2.57	2.57	2.57	2.57
Total cost at 5°C	5.7	5.0	4.7	4.7	5.7	5.0	4.7	4.7

[x]When the conventional sedimentation basin (Lovö basin) is used, it is advantageous to have a long detention time in the flocculation unit; otherwise, the same design values as for the lamella sedimentation unit can be used.

The total cost of treatment in terms of öre per cubic meter of water is about 10 per cent higher when the Lovö basin is used instead of a lamella unit.

11 CONCLUSIONS OF THE TECHNICAL AND THE ECONOMICAL ANALYSIS

The previously described mathematical models of the unit operations, shown in TABLE 8-1 and in FIG. 10-1 have been used to analyze the water treatment process as a whole.

Analysis of the single unit operations only is not sufficient. It is also of fundamental importance to consider the effect of the combination of unit operations required.The filtration operation has for instance a great influence on the optimum design of the flocculation unit. The number of reactors, the detention time in the flocculation unit, the dosage of activated silica, and the flocculation system as a whole must be considered for each filter design.

In the case of a very expensive filter operation, long residence times in the flocculation unit and a large number of reactors in series are economical. When the filter operation is less expensive, very short detention times (about 30 min.) in the flocculation unit could be used and the number of reactors could be limited to five or six. A comparison between a conventional flocculation system and the suggested new system, with shorter detention times in the first reactors than in the last ones in the series, shows that the new system is advantageous. The high G-values in the first reactors cause the turbidity to decrease very rapidly with time. The new system is also less expensive than the conventional one for an equal process volume.

There is also another possible advantage of the new system. The high velocity in the first reactors might affect the size and the density of the flocs. This could be of great importance, especially for the filter operation. In the present investigation it has been observed that the filtration properties of the flocs have been changed in an advantageous way with increasing G-values. It is, however, necessary to investigate this observation further, and it is then desirable to determine the particle size. This in turn, demands new instruments, e.g. a particle analyzer adapted for this particular field.

According to the flocculation model, the performance of the flocculation process is increased at a small amount of dispersion in the individual reactors. In other words, the performance is higher in a plugflow system than in a completely mixed-flow system.

It is, however, not quite certain that the floc formation is favored by a plugflow system. It is a task for future research to investigate the most favorable degree of dispersion in a reactor.

In the investigation it has also been shown that the flocculation process is dependent on the temperature. This can be compensated for by adding activated silica, but as this addition simultaneously may unfavorably affect the filter performance, it might be worth focusing the research on finding alternative methods to compensate for the temperature influence.

The lamella sedimentation unit used and analyzed in this investigation has been designed to meet the demands of practical application and low cost together with high performance. Several designs of the lamella unit have been tested. The highest surface load of about 25 m/h was obtained with a lamella unit with a length of 3.0 m and a horizontal distance between the lamellae of 0.04 m. The highest performance is, however, not achieved with the least expensive unit. The most economical lamella unit among those tested was found to be a unit with a lamella length of 3.0 m and a lamella distance of 0.05 to 0.06 m. The shape of the turbidity breakup curves for the different units varies with the lamella distance. A narrow distance means a steep breakup curve, whereas a great distance leads to a less steep curve, which from a practial point of view is to be preferred.

In the type of lamella unit studied, the inlet and outlet water system is relatively simple, due to practical reasons. However, this means that the entire lamella plate is not utilized. The economic analysis has shown that an improvement of the inlet and outlet system can be allowed to be fairly expensive, provided that the performance can be increased by about 20 per cent, which from a technical point of view should be feasible.The design of the inlet and outlet system is thus a suitable object for future research.

The cost of the lamella unit per unit volume is very high. It is higher than both flocculation and filter units, and for this reason the surface loading has to be as high as possible. Because of the shape of the turbidity breakup curve, it is necessary to carefully balance the surface loading and the removal efficiency against each other.

In comparing the investment cost of the Lovö basin with that of the lamella unit, one finds that the lamella unit is about 30 per cent less expensive. The maintenance cost for the lamella unit is also less than that of a Lovö basin, which further increases the advantage of the lamella unit. Roughly, one can state that a Lovö basin is at least 50 per cent more expensive than a lamella sedimentation unit.

The filter operation is the most expensive operation of the water treatment process. The investment cost is not as high as the running costs, even in cases where the price of the backwash water is low. The investment cost decreases almost linearly with the filtration rate, whereas the running costs can increase very rapidly with filtration rate when the available head is low and the price of the backwash water is high.

The price of the backwash water may vary from water works to water works, depending on different circumstances. It is, however, important either to reduce the cost of the treatment of backwash water or to minimize the amount of backwash water used at the backwashing of the filter. The investment cost of the equipment needed to achieve a reduction of the cost of the backwashing of a filter can be allowed to be relatively high. Studies of the filter backwashing technique are therefore recommended for future research.

Another important factor in reducing the cost of the filter operation is the choice of a sufficiently high permissible head loss for the filter in question. The permissible head loss ought to be chosen on the basis of a water quality limit. In this case the filter can be controlled, for instance by a turbidity meter connected to the effluent. This technique is, however, uncommon. Instead the head loss through the filter is limited to a certain value, normally between 1.0 and 2.0 m of water. The permissible head loss has been shown to vary with temperature, particle concentration, and dosage of activated silica, and the variation is specific for the various filter designs. A low value of the available head favors filters with a relatively coarse grain size, which in general means filters with a low permissible head loss and a low initial head loss. A high value of the available head favors filters with a high ability to resist a breakthrough of particles, which in turn means filters with a small grain size and a high initial head loss. In the present investigation an equation for calculation of the permissible head loss has been suggested and used in the calculations. However, the equation is empirical and valid only within certain limits, and therefore more detailed studies in order to develop a more general model are needed.

The cost of the filter media is not an important factor as long as the filter run time can be prolonged with the use of the media in question. In this study the filter with the most expensive bed material turned out to produce water at a low cost. In the choice, for example, between a two-media filter with a relatively coarse grain size and a two-media filter with a fine grain size, it is more economical in some cases to choose the latter filter design because of the high ability of that filter to resist a breakthrough of particles.

An increase of the filtration rate normally decreases the cost of water, but the filtration rate has to be limited to some value considering the water quality. In this investigation it has been shown that one can allow filtration rates up to 15 to 20 m/h for a conventional sand filter. The maximum permissible filtration rate may vary with filter design. This is also a task for future research.

Furthermore, the technical and economic analyses have shown that there are several possibilities in the choice of the design parameters for the unit operations in order to achieve a low cost of the water. Therefore, on the basis on this investigation it has only been possible to recommended intervals for the various parameters within which the most economical water treatment processes are to be found (TABLE 10-6). However, the principle of the water treatment plant model suggested can be applied at any water treatment plant where the unit operations consist of flocculation, sedimentation and filtration.

Several other conclusions could be drawn from this investigation but it is the author´s hope that the points of view on the treatment process given in this investigation may be of value in the design of water treatment plants and as a guide for future research.

12 SUMMARY

The main objective of this study has been to complete and combine existing theories for the different unit processes and unit operations in order to get an overall technical and economic model of the treatment process. Several investigators have studied the flocculation, sedimentation, and filtration operations but these studies have primarily dealt with mostly one operation at a time under idealized conditions; thus the operations have generally not been studied together which is mandatory for optimization studies.

A summary of the theory of flocculation kinetics is presented, and based upon this theory a new flocculation principle has been suggested. In this system the residence time in the primary reactors is shorter than in the terminal reactors. The mathematical model describing flocculation performance in reactors in series has been based on the work of others (Argaman and Kaufman, Parker, Kaufman and Jenkins). This model is based upon the assumption that the breakup of flocs is caused by floc surface erosion. It has been improved to some extent in order to be able to use the velocity gradient in a single reactor instead of using the mean velocity gradient of the whole flocculation system. An empirical model has been derived, describing the settling properties of the flocs based on the flocculation performance.

The sedimentation models derived are based on the Hazen/Camp ideal sedimentation theory for discrete particles. Two different models, one for a conventional sedimentation basin and the other for a lamella sedimentation unit, have been suggested.

The four models — the flocculation model, the settling characteristics model, and the two sedimentation models — make it possible to predict the concentration of the suspended solids in the effluent from a conventional as well as from a lamella sedimentation unit at different water temperatures, dosages of coagulant aid, flocculation parameters, and sedimentation designs.

Several different models for the filtration operation have been described in the literature. The different models are developed for ideal conditions and are supposed to predict the removal efficiency as well as the head loss in a filter for any given period of time. The different models derived are more or less special cases of a model suggested by Ives (1969), who also has proposed an approximate filtration model describing the change in head loss with time. The model used in this investigation is based on that approximate model, but it has been necessary to modify it empirically due to wide variations in the filtration variables. An empiric model has been developed to describe the water quality in the filtrate at a given head loss. Another empiric model has been made for the maximum permissible head loss through a filter considering the water quality. The different variables studied and the various models are summarized in TABLE 8-1.

The research has been carried out in cooperation with the Water and Sewage Works in Göteborg. The research activities have mainly been located to the Water Treatment Plants at Alelyckan and Lackarebäck. Also, investigations have been performed for two years in connection with the South-Water Project. The different raw waters studied came from the Göta Älv river, lake Delsjön, and lake Bolmen, and can be regarded as very soft waters with low buffer capacities.

The investigations have been performed on different scales ranging from laboratory experiments to full scale operation. In the investigation aluminium sulfate, lime, and activated silica have been used throughout the experiments. The dosage of aluminium and lime has been almost constant, while the activated silica addition has been varied.

From the studies of the flocculation operation the following conclusions may be drawn. The flocculation process is affected by temperature, especially when the residence time is short and when no activated silica is added.

An increase of the floc volume fraction increases the flocculation performance. The optimum pH-value is influenced by temperature. The optimum pH-value at a temperature of 15 to 16^{o}C is 6.2-6.3 and 6.4-6.6 at a temperature of 4^{o}C.

Several studies were carried out to examine the hydraulic conditions in a reactor. The geometrical design was found to be unimportant at least in the relatively normal units used in these investigations.

Tracer studies performed in order to determine the residence time distribution showed that the energy input (velocity gradient) affected the time distribution. At higher velocity gradients the reactor was almost completely mixed and at lower velocity gradients the reactor was characterized by a lower degree of dispersion.

The residence time distribution in a reactor was also influenced by the rate of flow through the reactor. The amount of dispersion in the reactor decreased with an increase in the rate of flow. The installation of a baffle in a reactor had, of course, a marked effect on the residence time distribution.

The direction of the flow through the reactor was not important for the flocculation performance.

Different paddle designs were briefly studied, and it was observed that the paddle design was important. A paddle design factor was developed to characterize the energy distribution in the reactor. A paddle with a relatively large cross sectional area with several paddle blades was found to be very effective. The floc size distribution was uniform, and the overflow rate in the sedimentation unit could be increased.

A new type of compartmentalization of the total flocculation volume was tested. Instead of a series of reactors of equal size, the primary reactors in the series were smaller than the terminal ones. The results of these tests showed that it was possible to increase the flocculation performance by using this new design.

The results obtained in the different plants agreed well with the proposed flocculation performance model when floc breakup was assumed to be caused by floc surface erosion. In the flocculation model for a tapered multi-compartment flocculator, the following variables have been considered:

Number of reactors
Residence time distribution within a single reactor
Residence time distribution between the reactors
Velocity gradient
Coagulant aid
Temperature

The correlation coefficient for the correlation between the experimentally determined and the theoretically predicted flocculation performance was > 0.95.

The studies of the settling characteristics of the flocs were carried out in a special settling column in which the flocs were allowed to settle while the turbidity was measured at a certain depth as a function of time. The settling characteristics of the flocs were based on the work of Rosén (1967) who had found that the settling velocity of flocs had an approximately Gaussian distribution. The settling velocity equation was based on the hypothesis that the mean settling velocity of the flocs and the standard deviation of the settling velocity were functions of the residual particle concentration obtained in terms of turbidity from the flocculation performance model. Earlier studies carried out by Rosén were also used in order to prove the hypothesis. It was shown that the depth of sedimentation in the settling column was important, both for the mean settling velocity and the standard deviation of the settling velocity of the flocs.

The correlation coefficient for the relationship between the experimentally determined and the theoretically predicted settling velocities was in the range of 0.75 - 0.95, depending on the number of reactors and the mean flocculation residence time used in the study.

During the past decade the separation techniques have been developed extensively, and many new methods for the separation of suspended particles have been applied. The object of this investigation was to study conventional sedimentation basins as well as so-called tilted-plate separators or lamella sedimentation units. At the Division of Water Supply and Sewage, studies of lamella sedimentation have been performed in different pilot plants. In this report the results obtained in these pilot plants and in full scale tests are discussed. The lamella sedimentation technique has been compared with conventional sedimentation techniques.

The validity of the lamella sedimentation theory was tested. Several different system variables such as length of the lamella, distance between lamellæ, and the suspension characteristics have been analyzed. The experimental result agrees with the theory, showing that an increase in settling velocity of the flocs leads to a corresponding increase in the overflow rate. According to the theory the surface loading Q/A is increased in proportion to the relationship between the length of the lamella and the horizontal distance between the lamellæ (L/S-ratio). This has been confirmed by the investigation. Further observations show that the removal efficiency slightly increases with increasing L/S-ratio. The overflow rate Q/BL is independent of the horizontal distance between the lamellæ, but there exists a critical distance depending on the thickness of the sludge sliding down on the lamella. The critical distance is dependent upon the volume concentration of the sludge. The critical distance for the particular sludge studied in this investigation was estimated at 0.04 m. The corresponding critical rate of flow was calculated to be 0.005 m/s. More favorable and constant hydraulic conditions are developed in the lamella sedimentation unit in comparison with the conventional basin. Reynold's number is less than 100 which means that the flow is laminar. It has also been observed that the settling properties of the flocs are changed in the lamella sedimentation unit. The settling velocity distribution of the flocs is uniform.

The surface loading is dependent upon the design of the inlet and outlet of water in the lamella unit. A surface loading as high as 40 m/h has been reached in a unit with an ideal inlet and outlet of water. In the full scale unit at the Water Treatment Plant at Lackarebäck a surface loading of 17 to 18 m/h has been reached. This lamella sedimentation unit has a lamella length of 3.0 m and a horizontal distance between the lamellæ of 0.08 m. The corresponding surface loading of the conventional Lovö-basin is about 3 - 4 m/h.

The sludge handling has not been studied to any large extent, but it may be noted that the sludge can be automatically withdrawn and the water loss will be less than 0.5 % which is less than that obtained for a conventional basin.

If the removal efficiency is assumed to be described by the equation for an ideal vertical sedimentation basin and if the settling velocity is assumed to have an approximately Gaussian distribution, the results obtained in the different pilot and full scale plants agree well with the proposed model.

Filtration through granular media is an important unit operation for solid-liquid separation in water treatment that has been in practical use for a long time. It was not until the last decade, however, that the practical and theoretical development of the operation has been intensified and new filter designs have come into use. In this investigation some of the new filter designs have been studied in order to increase the knowledge of the different factors affecting the filter operation. Comparative tests of several filter designs have been performed in a relatively small pilot plant, but tests on a larger scale have also been carried out under realistic conditions, and finally, some results obtained in the pilot plants have been compared with filter operations in full scale at the Water Treatment Plant at Lackarebäck.

The following filter designs have been investigated under various conditions:

One medium filter:	*Sand* (d_{10} = 0.87 mm)
One medium filter:	*Sand* (d_{10} = 0.65 mm)
Dual-media filter:	*Sand* (d_{10} = 0.87 mm)/*Anthracite* (d_{10}=1.7 mm)
Dual-media filter:	*Sand* (d_{10} = 0.65 mm)/*Anthracite* (d_{10}=0.95 mm)
Three-media filter:	*Garnet*(d = 0.25-1.25 mm)/*Sand* (d_{10}=0.67 mm)/ /*Anthracite* (d_{10} = 1.6 mm)
Upflow filter:	*Sand* (d_{10} = 0.87 mm)
One medium filter:	*Act.carbon* (d_{10} = 1.0 mm)
Dual-media filter:	*Sand* (d_{10} = 0.65 mm)/*Act.carbon* (d_{10}=0.67 mm)

All the downflow filters may be considered representative, and the results obtained may be used as a basis for full scale design. The upflow filter is, however, due to the principle, dependent on the scale, and the actual design of the system, which prevents the bed from lifting. Thus, even if several investigations have been carried out with upflow filters with promising results,it is doubtful if they can be used to predict the result in a full scale plant. Several conclusions may be drawn from the investigation and some of the most important ones are summarized in the following paragraphs.

In terms of turbidity at the initial stage of the filter period, the filtrate quality measured is almost equal for the different filters. The filtrate quality pattern during the filter runtime is, however, different for the various filters, and the quality pattern is strongly dependent on the addition of activated silica and the temperature.

With decreasing grain size, increasing dosage of activated silica, and increasing temperature, a more stable filtrate quality is obtained up to a certain head loss.

The filtrate quality has proved to be almost independent of the filtration rate, at least up to a rate of 15 to 20 m/h. This is true for a conventional sand filter.

It has also been shown that an increase in turbidity in influent causes an increase in removal efficiency.

The maximum permissible head loss through the bed, considering the filtrate quality, has empirically been expressed in a mathematical form. The maximum permissible head loss generally increases for all the filters with an increase in temperature and dosage of activated silica. The maximum permissible head loss is higher for filters with finer media than for filters with coarser media. Especially high head loss can be obtained by the fine dual-media filter or three-media filter. In the dual- and three-media filters, the different layers mix at the interfaces, which in general leads to a high removal efficiency and resistance against a breakthrough due to a high head loss. In the mixing zone the pore width is small, which causes a rapid increase in head loss, and it is thus necessary to be able to control the degree of mixing by choosing the right media.

The increase in head loss with time is dependent upon temperature, dosage of activated silica, particle concentration and filter design. The conventional filters with only one medium are in general more dependent on temperature and the amount of activated silica added than the two- or three-media filters, which are more dependent on the particle concentration. In general, the rate of head loss increase for a coarse dual-medial filter is about half of that of a conventional sand filter. At extremely low temperatures a very rapid increase in head loss has been observed, especially when activaled silica is added to the conventional filters. This is probably due to a change in particle characteristics, originating from a change in the perikinetic flocculation, caused by a decrease in the molecular movement. This emphasizes the necessity of placing the raw water intake under the stratification layer when a lake is used as a water source. The coarse dual-media filters were not affected to the same degree, and thus it may be assumed that the filtration mechanism was changed and that straining probably became the dominant mechanism.

The filter run time is dependent upon the permissible head loss. If the available head is fixed to a certain value, for instance 1.5 m of water, the filtration time for a conventional sand filter can be estimated at 20 hours and 10 hours at the filtration rates of 10 m/h and 15 m/h, respectively. Corresponding values for the coarse sand-anthracite filter were about 45 hours and 25 hours, respectively.

The wash water consumption for the different filters has been assumed to be almost independent of the media size. The consumption of washwater for the filter is dependent on several factors, but in general the washwater sonsumption for a conventional sand filter and a sand-anthracite filter may be estimated at 2 - 3 % and 1 - 2 %, respectively.

The results concerning the conventional sand filters and the activated carbon filters were all in agreement with the results obtained in full scale operation at the Water Treatment Plant at Lackarebäck.

Based on the technical investigation alone, it is impossible to optimize the water treatment process. This investigation has, however, shown that there are several possible design alternatives in addition to the ones used in water treatment plants at the present time.

With the technical investigation as a basis, an economic analysis has been performed. In this analysis the costs, which affect the unit operation volume, have been considered. A picture of the principle of the complete water treatment plant model used is shown in FIG. 10-1.

Analysis of the single unit operations only is not sufficient. It is also of fundamental importance to consider the effect of the combination of unit operations required. The filtration operation has for instance a great influence on the optimum design of the flocculation unit. The number of reactors, the detention time in the flocculation unit, the dosage of activated silica, and the flocculation system as a whole must be considered for each filter design.

In the case of a very expensive filter operation, long residence times in the flocculation unit and a large number of reactors in series are economical. When the filter operation is less expensive, very short detention times (about 30 minutes) in the flocculation unit could be used and the number of reactors could be limited to five or six. A comparison between a conventional flocculation system and the suggested new system with shorter detention times in the first reactors than in the last ones in the series shows that the new system is advantageous. The high G-values in the first reactors cause the turbidity to decrease very rapidly with time. The new system is also less expensive than the conventional one for an equal process volume.

There is also another possible advantage of the new system. The high velocity gradients in the first reactors might affect the size and the density of the flocs. This could be of great importance, especially for the filter operation. In the present investigation it has been observed that the filtration properties of the flocs have been changed in an advantageous way with increasing G-values.

The lamella sedimentation unit used and analyzed in this investigation has been designed to meet the demands of practical application and low cost together with high performance. Several designs of the lamella unit have been tested. The highest surface loading of about 25 m/h was obtained with a lamella unit with a length of the lamella of 3.0 m and a horizontal distance between the lamellæ of 0.04 m. The highest performance is, however, not achieved with the least expensive unit. The most economical lamella unit among those tested was found to be a unit with lamella length of 3.0 m and a lamella distance of 0.05 to 0.06 m. The shape of the turbidity breakup curves for the different units varies with the lamella distance. A narrow distance means a steep breakup curve, whereas a great distance leads to a less steep curve, which from a practical point of view is to be preferred.

In the type of lamella unit studied, the inlet and outlet water system is relatively simple, which means that the entire lamella plate is not utilized. The economic analysis has shown that the improvement of the inlet and outlet system can be allowed to be fairly expensive, provided that the performance can be increased by about 20 per cent, which from a technical point of view should be feasible.

In comparing the investment cost of the Lovö basin with that of the lamella unit, one can state that the lamella unit is about 30 per cent less expensive. The maintenance cost for the lamella unit is also less than that of a Lovö basin, which further increases the advantage of the lamella unit. Roughly, one can state that a Lovö basin is at least 50 per cent more expensive than a lamella sedimentation unit.

The filter operation is the most expensive operation of the water treatment process. The investment cost is not as high as the running costs even in cases where the price of the backwash water is low. The investment cost decreases almost linearly with the filtration rate, whereas the running costs can increase very rapidly with filtration rate when the available head is low and the price of the backwash water is high.

The price of the backwash water may vary from water works to water works, depending on different circumstances. It is, however, important to either reduce the cost of the treatment of the backwash water or minimize the amount of backwash water used at the backwashing of the filters. The investment cost of the equipment needed to achieve a reduction of the cost of the backwashing of a filter can be allowed to be relatively high.

Another important factor in reducing the cost of the filter operation is the choice of a sufficiently high permissible head loss for the filter in question. The permissible head loss ought to be chosen on the basis of a water quality limit. In this case the filter can be controlled, for instance by a turbidity meter connected to the effluent. This technique is, however, uncommon. Instead the head loss through the filter is limited to a certain value, normally between 1.0 and 2.0 m of water. The permissible head loss has been shown to vary with temperature, particle concentration, and dosage of activated silica, and the variation is specific for the various filter designs. A low value of the available head favors filters with a relatively coarse grain size, which in general means filters with a low permissible head loss and a low initial head loss. A high value of the available head favors filters with a high ability to resist a breakthrough of particles, which in turn means filters with a fine grain size and a high initial head loss. In the present investigation an equation for calculation of the permissible head loss has been suggested and also used in the calculations. However, the equation is empirical and valid only within certain limits, and therefore more detailed studies are needed in order to develop a more general model.

The cost of the filter media is not an important factor as long as the filter run time can be prolonged with the use of the media in question. In this study the filter with the most expensive bed

material turned out to produce water at a low cost.
In the choice, for example, between a two-media filter with a relatively coarse grain size and a two-media filter with a fine grain size, it is more economical in some cases to choose the latter filter design because of the high ability of that filter to resist a breakthrough of particles. An increase of the filtration rate normally decreases the cost of water, but the filtration rate has to be limited to some value considering the water quality. In this investigation it has been shown that one can allow filtration rates up to 15 to 20 m/h for a conventional sand filter. The maximum permissible filtration rate may vary with filter design.

Furthermore, the technical and economic analyses have shown that there are several possibilities in the choice of the design parameters for the unit operations in order to achieve a low cost of the water. Therefore, on the basis of this investigation, it has only been possible to recommend intervals for the various parameters within which the most economical water treatment process is to be found (TABLE 10-6). However, the water treatment plant model suggested can be applied in principle at any water treatment plant where the unit operations consist of flocculation, sedimentation and filtration.

13 REFERENCES

Agrawal, G, D, 1966, Electrokinetic Phenomena in Water Filtration, Ph,D, Thesis, Univ. of California, Berkely.

Argaman, Y, & Kaufman,W, 1970, Turbulence and Flocculation. J. Sanit. Engng Div., ASCE, Vol. 96, SA2, p.223

Asemann, K, und Wirth, H, 1973, Der Wasserverbrauch in ausgewählten Bereichen von Wirtschaft under Vervaltung in Frankfurt am Main. Herausgeben von den Stadtwerken und dem Statischen Amt und Wahlmat der Stadt Frankfurt a.M.

Baylis, J, R, 1937, Experience in Filtration, JAWWA, Vol.29, p. 1010.

Bean, L, 1953, Study of Physical Factors Affecting Flocculation, Water Works Engng, Jan., p. 33.

Camp, T, R, 1955, Flocculation and Flocculation Basins, Transactions, ASCE, Vol. 120, p.1.

Camp, T, R, 1946, Sedimentation and Design of Settling Tanks. Tr.A. Soc. Civil Engrs. p. 111.

Camp, T, R, & Stein, P, C, 1943, Velocity Gradients and Internal Work in Fluid Mot, Bost. Soc. Civ. Engng, October, p.219.

Cleasby, J, L, 1969, Approaches to a Filterability Index for Granular Filters,JAWWA, Vol. 61, p. 372.

Cleasby, J, L, & Bauman, R, 1962, Selection of Sand Filtration Rates, JAWWA, Vol. 54, p. 579.

Cleasby, J,L, & Sejkora, G,D, 1975, Effect of Media Intermixing on Dual Media Filtration, J. Environmental Engng Div. ASCE, Vol.101, (EE4), p. 503.

Conley, W, 1961, Experience with Anthracite-Sand Filters, JAWWA, Vol. 53, p.1473.

Craft, T, F, 1966, Review of Rapid Sand Filtration Theory, JAWWA, Vol. 58, p. 428.

Daily, J, W, & Harleman, D, R, F, 1965, Fluid Dynamics, Addison-Wesley (Canada) Ltd, Don Mills, Ontario.

Danckwerts, P, 1953,Continous flow systems. Distribution of residence times. Chem Eng Sci, vol 2, No 1.

Davis, E, & Borchardt, J, A, 1966, Sand Filtration of Particulate Matter, J. Sanit. Engng Div., ASCE, Vol. 92 (SA5), p. 47.

Deb, A, K, 1969, Theory of Sand Filtration, J. Sanit. Engng Div., ASCE, Vol. 95 (SA3), p. 399.

Diaper, E, & Ives, K, J, 1965, Filtration through Size-Graded Media. J. Sanit. Engng Div., Am. Soc. Civil Engrs, Vol. 91 (SA4), p. 89.

Drobny, N, L, 1963, Effect of Paddle Design on Flocculation, J. Sanit. Engng Div., ASCE, Vol. 89 (SA2), p. 17.

Eliasson, R, 1941, Clogging of Rapid Sand Filters. JAWWA, Vol. 33, p. 126.

Elliasen, R, 1935, An Experimental and Theoretical Investigation of the Clogging of a Rapid Sand Filter. Sc.D. Thesis, Massachusetts Inst. Tech.

Fair, G, M, Geyer, J, C, & Okun, D, A, 1967, Water and Waste Water Engineering, John Wiley & Sons Inc. New York.

Gomella, C, 1974, Clarification avant filtration ses progrés récents. International Water Supply Ass., Tenth Congr. 19-22 Aug. Brighton.

Gregory, J, 1964, Molecular Forces and Electrokinetic Effects in Filtration, Ph.D. Thesis, Univ. of London.

Ham, R, K, & Christman, R, F, 1969, Agglomerate Size Changes in Coagulation. J. Sanit. Engng Div., ASCE, Vol.95(SA3), p.481.

Hamann, C,L, & McKinney,R,E, 1968, Upflow Filtration Process, JAWWA, Vol. 60, p. 1023.

Hahn, H, H, & Stumm, W, 1968, Kinetics of Coagulation with Hydrolyzed Al (III). The Rate-Determining Step. J.Colloid Interface Sc. Vol. 28:1 p. 134.

Hannah, S, A, Cohen,J,M,& Robeck,G,G1967, Measurement of Flock Strength by Particle Counting. JAWWA, Vol. 59, p.843.

Harris, H,Kaufman,W,J,& Krone,R,B, 1966, Orthokinetic Flocculation in Water Purification. J. Sanit Engng Div., ASCE, Vol.92 (SA6), p.95.

Hazen, A, 1904, On Sedimentation. Trans, ASCE, Vol. 53:45.

Hedberg, T, 1969, Optimering av den kemiska reningsprocessen samt hårdhetshöjning av mjuka ytvatten. Publikationsserie B nr 69:1, Inst f VA-teknik, CTH, Göteborg.

Hedberg, T, 1974, Studium av flockningsprocessen. Publikationsserie C nr 74:1, Inst f VA-teknik, CTH, Göteborg.

Hedberg, T, 1974, Studium av flockningsprocessen. Del 1: Teoretisk bakgrund. Publikationsserie B nr 74:2, Inst f VA-teknik, CTH, Göteborg.

Hedberg, T, 1974, Kemisk rening av vatten från Bolmen. Rapport från försöksverksamhet 1971-1973. Sydvatten.

Hedberg, T, 1975, Studium av flockningsprocessen. Del 2: Undersökningar. Publikationsserie B nr 75:2, Inst f VA-teknik, Göteborg.

Heertjes, P, M, & Lerk, C, E, 1962, Some Aspects of the Removal of Iron from Groundwater. Proc. Congr. Interaction between Fluids and Particles, London, p.269.

Heertjes, P, M, & Lerk, C, E, 1967, The Functioning of Deep-Bed Filters; Part I: The Filtration of Colloidal Solutions; Part II: The Filtration of Flocculated Suspensions. Trans. Inst. Chem. Engs. Vol. 45, p. 129.

Herzig, J, P, LeClerc, D, M, & LeGoff, 1970, Flow of Suspensions through Porous Media-Application to Deep Filtration. I.and E. C., Vol. 62 (5), p. 8.

Hilmer, A, & Andersson, B, 1973, Renvattenproblematiken. Bull. Serie VA nr 1, LTH, Lund.

Hudson, H, E, Jr, 1957, Flocculation and Flocculation Aids. JAWWA, Vol. 49, p. 242.

Hudson, H, E, Jr, 1965, Physical Aspects of Flocculation, JAWWA, Vol. 57, p. 885.

Hudson,H,E,& Wolfner,J,P,1967, Design of Mixing and Flocculation Basins. JAWWA, Vol. 59, p. 1257.

Hunter, R, J, & Alexander, A, E, 1963, Surface Properties and Flow Behaviour of Kaolinite; Part III: Flow of Kaolinite Sols through A Silica Column. J. Colloid Sci., Vol. 18, p. 846.

Hyde,C,G,& Ludwig,H,F,1944, Some new Features in the Design of Vertical Flocculation Units. JAWWA, Vol. 36, p. 151.

Ingersoll, A, C, McKee, J, E, & Brooks, N, H, 1965, Fundamental Concepts of Rectangular Settling Tanks. Trans. Am. Soc. Civil Engrs., Vol. 121, p. 1179.

Ison, C, R, & Ives, K, J, 1969, Removal Mechanisms in Deep Bed Filtration, Chem. Eng. Sci. Vol. 24, p. 717.

Ives, K, J, 1960, Rational Design of Filters. Proc. Inst. Civ. Engrs. Vol. 16, p. 189.

Ives, K, J, 1960, Simultation of Filtration on an Electronic Digital Computer. JAWWA Vol. 52, p. 933.

Ives, K, J, New Concepts in Filtration. Water and Water Eng. Vol. 65, No. 307, p. 341.

Ives, K, J, 1961, Filtration Using Radioactive Algæ. J. Sanit. Engng Div. ASCE, Vol. 87, (SA3), p.23.

Ives, K, J, 1962, A Theory of the Function of Deep Filters. Proc. Third Congr.Europ. Fed. Chem. Engs; Interaction Between Fluids and Particles, June, p. 260.

Ives, K, J, 1963, Simplified Rational Analysis of Filter Behaviour. Proc. Inst. Civil Engs, Vol. 25, p. 345.

Ives, K, J, 1964, Progress in Filtration, JAWWA, Vol.56, p. 1225.

Ives, K, J, & Pienvichetr, V, 1965, Kinetics of Filtration of Dilute Suspensions. Chem. Eng. Sci. Vol. 20, p. 965.

Ives, K, J, & Sholji, I, 1965, Research on Variables Affecting Filtration, J. Sanit. Engng Div. ASCE, Vol. 91 (SA4), p. 1.

Ives, K, J, & Gregory, J, 1966,Surface Forces in Filtration. Proc. Water Treatment and Exam. Vol. 15 (pt2), p. 93.

Ives, K, J, 1967, Deep Filters. Filtration and Separation Vol. 4 (3/4), p. 125.

Ives, K, J, 1969, Theory of Filtration. Spec. Subj. No. 7. IWSA Congr., Vienna.

Ives, K, J, 1973, Theory of Flocculation for Continuous Flow System. J.Envir. Engng Div., Febr., p. 417.

Iwasaki, T, 1937, Some Notes on Sand Filtration. JAWWA, Vol. 29 p. 1591.

Lagvankar, A, L, & Gemmel, R, S, 1968, A Size-Density Relationship for Flocs. JAWWA, Vol. 60, p. 1040.

Levenspiel, O, 1962, Chemical Reaction Engineering, Chap.9, I. Wiley & Sons, Inc.

Lindquist, E, G, W, 1945, Avsättningsbassänger av Lovö-typ. Kommunalteknisk Tidskrift, nr 2, p. 29.

Ling, J, T, 1955, A Study of Filtration Through Uniform Sand Filters. Proc. ASCE Sanit.Engng Div. Vol. 81, p. 751.

Mackrle, V, & Mackrle, S, 1961, Adhesion in Filters. J. Sanit. Engng. Div. ASCE, Vol. 87, (SA5), p. 17.

Maroudas, A, & Eisenklam, P, 1965, Clarification of Suspensions: A Study of Particle Deposition in Granular Media; Part I: Some Observations on Particle Deposition. Chem. Engng. Sci. Vol. 20, p. 867.

Maroudas, A, & Eisenklam P, 1965, Clarification of Suspensions: A Study of Particle Deposition in Granular Media; Part II: A Theory of Clarification. Chem. Engng. Sci. Vol. 20, p. 875.

O'Melia, C, R, 1972, Coagulation and Flocculation. Weber (Red): Physicochemical Processes for Water Quality Control. Wiley-Interscience, p. 61.

Mohtadi, M, F, & Rao, P, N, 1973, Effect of Temperature of Flocculation of Aqueous Dispersions. Water Research, Pergamon Press, Vol. 7, p. 747.

Miller, D, G, 1967, Rapid Filtration Following Coagulation Including the Use of Multilayer Beds. Proc. Sec. Wat. Treat.Exam. Vol. 16, p. 192.

Mints, D, M, 1951, Kinetics of Filtration of Low-Concentration Water Suspensions in Water Purification Filters (In Russian), Dokl. Akad., Nank SSSR, Vol 78 (2), p. 315.

Mints, D, M, 1966, Modern Theory of Filtration. Int.Water Supply Assoc.Congr., Barcelona.

Parker, D, S, Kaufman, W, I, & Jenkins, D, 1972, Floc Breakup in Turbulent Flocculation Processes. J. Sanit. Engng Div. ASCE Vol.98 (SA1), p. 79.

Rechard, P, A, 1971, Water Budget for the City of Laramie Wyoming, Environmental Protection Agency, Project ≠ 17050 DVO.

Reed, S, C, & Murphy, S, R, 1969, Low Temperature Activated Sludge Settling. J. Sanit. Engng Div. ASCE, Vol. 95 (SA4), p. 747.

Riddick, I, 1961, Zeta Potential: New Tool for Water Treatment. I & II, Chem. Engng, June July, p. 121.

Ritchie, A, R, 1956, Theoretical Aspects of Flucculation and Coagulation. Proc. Soc. Wat. Treat. Exam., p. 81.

Robinson, M, 1964, Flocculation - A Literature Survey of Theory and Practice. Water and Water Engng, March, p. 96.

Rosén, B, 1967, Effektillförselns inverkan på flockuppbyggnaden. Inst f VA-teknik, Chalmers tekniska högskola, Publikation B67:1.

Saitenmacher, L, 1965, Zur Ermittlung des hydraulischen Wirkungsgrades bei vertical durchflossenen Absetzbecken. Wissch. Zeitschrift der Tech. Univ. Dresden 14, Heft 4.

Shekhtman, Yu, M, 1961, Filtration of Suspensions of low Concentration, (In Russian), Published USSR Acad. of Sci. (Inst of Mech.), Moscow.

Stanley, P, R, 1955, Sand Filtration Studied with Radiotracers. Soc. Civil Engrs, Vol. 81, p. 592.

Stein, P, C, 1940, A Study of the Theory of Rapid Filtration of Water through Sand. D. Sci. Thesis, Massachusetts Inst of Tech.

Tesarik, I, 1967, Flow in Sludge-Blanket Clarifiers. J. Sanit. Engng Div., ASCE Vol. 93 (SA6), Proc. Paper 5698, p. 105.

Tolman, S, L, 1949, Use of Models in Solving Flocculation Problems. JAWWA Vol. 41, p. 641.

Tolman, S, L, 1942, Water Conditioning by Flocculation. JAWWA, Vol. 34, p. 404.

Tomas, D, G, 1964, Turbulent Disruption of Flocs in Small Particle Size Suspension. J. Am. Inst. Chem. Eng., Vol. 10, p. 517.

Tomas, H, A, & Archibald, R, S, 1952, Longitudinal Mixing Measured by Radioactive Tracers. Trans. Am. Soc. Civil Engng, Vol. 117, No. 2518, p. 839.

Villemonte, J, R, Rohlich, G, A, & Wallace, A, T, 1967, Hydraulic and removal efficiencies in sedimentation basins. Adv. Wat. Poll. Res., Vol.2, p. 382, Pergamon Press.

Weijman-Hane, G, 1972, Kompendium i VA-teknik. Del V: Rening av försörjningsvatten. Inst f VA-teknik, Chalmers tekniska högskola.

Weijman-Hane, G, 1963, Koagulering, sedimentering och filtrering. Försök med lamellsedimentering. Publikation B63:3, Inst f VA-teknik, Chalmers tekniska högskola.

Weijman-Hane, G, & Rosén, B, 1966, Koagulering, sedimentering och filtrering. Försök med lamellsedimentering II. Publikation B66:4, Inst f VA-teknik, Chalmers tekniska högskola.

Wolf, D, & Resnick, W, 1963, Residence Time Distribution in Real Systems. Ind. and Eng. Chem. Fundamentals, Vol. 2, p. 287.

Wright, A,M, Kavanaugh,M,C, & Pearson, E,A, 1970, Filtration Kinetics in Water and Waste Water, First Annual Progress Report, Sanitary Engng Res. Lab, Univ. of California, Berkely.

Wunsch, W, Tuppeck,F, und Plett, H, 1959, Untersuchung den Einfluss der Aussentemperatur auf den Wasserverbrauch, GWF, Bd. 100, (1959) H. 14, p. 333.

Yao, K, M, 1970, Theoretical Study of Highrate Sedimentation. J. WPCF, Vol. 42, No. 2, Part 1, p. 218.

Ødegaard, H, 1975, Flocculation of Phosphate Precipitates in Waste Water Treatment. Ph. D. Thesis. Div. of Hydraulic Engng Univ., Trondheim.

APPENDICES

APPENDIX 1

PLANT DESCRIPTIONS

PLANT	1	Pilot plant,	Alelyckan
"	2	" -	Alelyckan
"	3	" -	Lackarebäck
"	4	" -	Bolmen
"	5	" -	Lackarebäck
"	6	Full scale,	Alelyckan
"	7	" -	Lackarebäck
"	8	Pilot plant full scale,	Lackarebäck

Appendix 1

PLANT 1

Pilot plant, Alelyckan

In order to carry out research, especially concerning lamella sedimentation, a pilot plant was designed and placed at the Water Treatment Plant at Alelyckan. This pilot plant is described by Weijman-Hane (1963). The pilot plant was completed by filter units and the modified pilot plant is shown in FIG. 1.

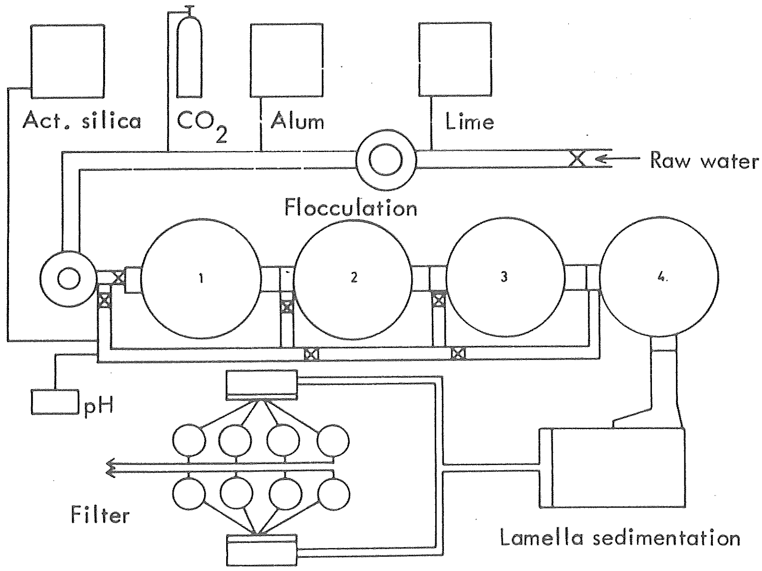


FIG. 1 Sketch of the pilot plant at Alelyckan Water Treatment Plant

The flocculation unit consisted of four circular tanks. The tanks were connected in such a way that the direction of the flow was upward in the tank. The reactor design and the paddle design are shown in FIG. 2.

The stirring equipment consisted of two grids fixed on a common axis. The stirring engine could be varied continuously between 0 and 20 rpm. In order to prevent undesirable rotation of the water each tank was equipped with vertical angle-bars and with horizontally arranged star-formed pipes.

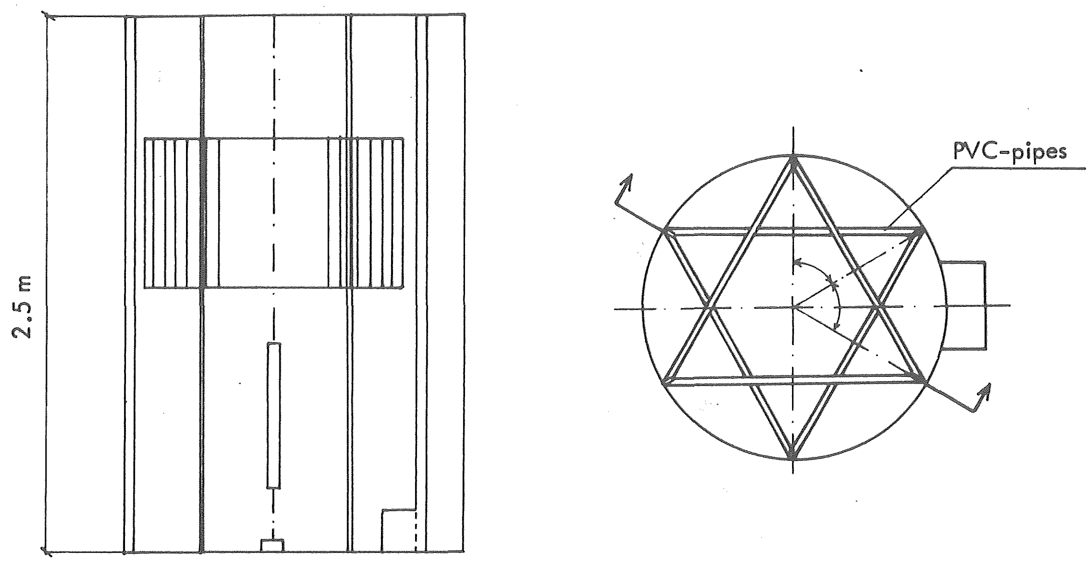


FIG. 2 Reactor with stirring equipment

In this particular pilot plant measurements of the input energy have been carried out (Rosén, 1967). The relationship between the energy input and the revolutions per minute was determined to:

$$w = 10^{-3} \cdot n^3$$

in which w = effect in w/m^3
n = stirring speed rpm

The velocity gradient G can be calculated by the following expression:

$$\bar{G} = \sqrt{\frac{w}{\mu}} = \sqrt{\frac{w^{-3} \cdot n^3}{\mu}}$$

in which μ is the kinematic viscosity of water. The velocity gradient $\bar{G}$ as a function of the stirring speed at different temperatures is shown in FIG. 3.

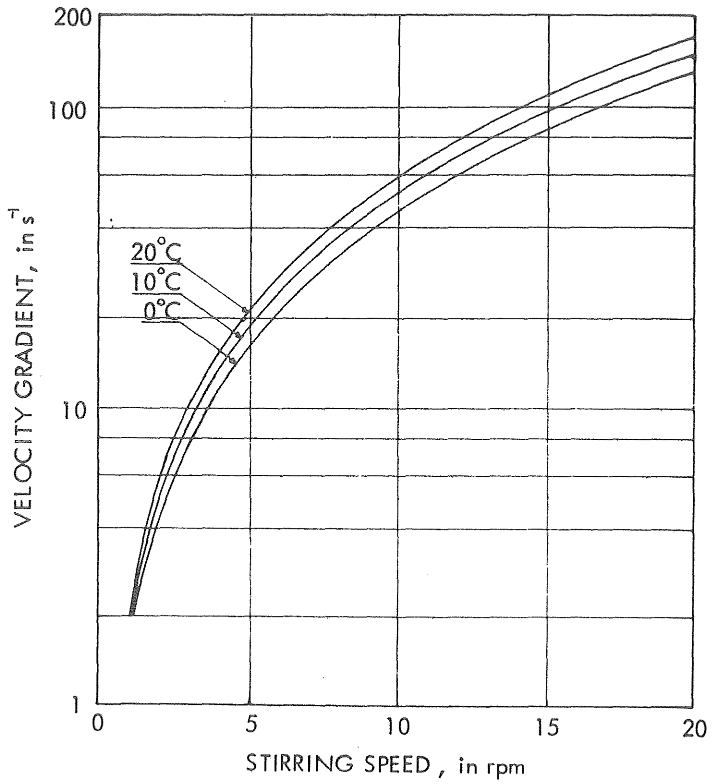


FIG. 3 *G*-value as a function of stirring speed

The lamella sedimentation unit was designed with a lamella width of 1 m, a lamella length of 1.84 m and a horizontal distance between the lamellae of 0.1 m. The unit was equipped with an inlet system of water on both sides at the lower part of the lamellae, FIG. 4

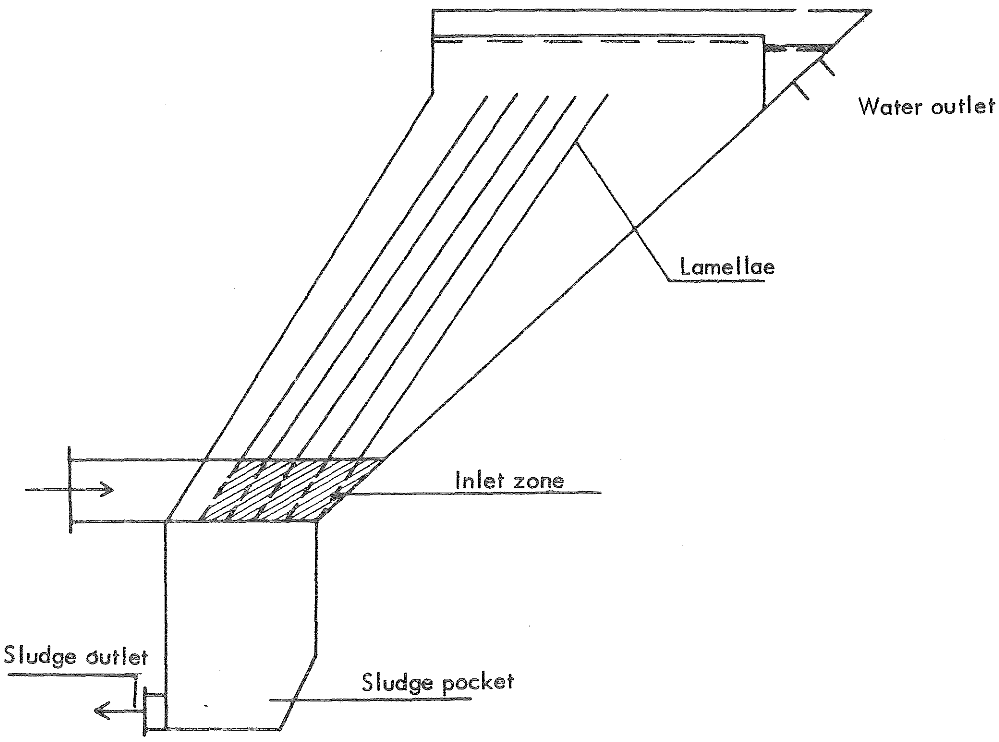


FIG. 4 Section of the lamella sedimentation unit

The filtration unit, FIG. 5, consisted of eight plexiglass tubes with a diameter of 0.125 m and a height of about 3 m. Above the bottomplate of the tubes there was a metal net on which the filter bed was resting. Four of the filters were adapted for conventional, down-flow, filtration and four for up-flow filtration. Along the filter tube there were pipes placed for pressure measurements within the filter bed. The filtration plant was also equipped with filtration rate controls and arrangements for keeping the water pressure constant. All the filters could be used simultaneously at varying filtration rates ranging from 0 to 30 m/h.

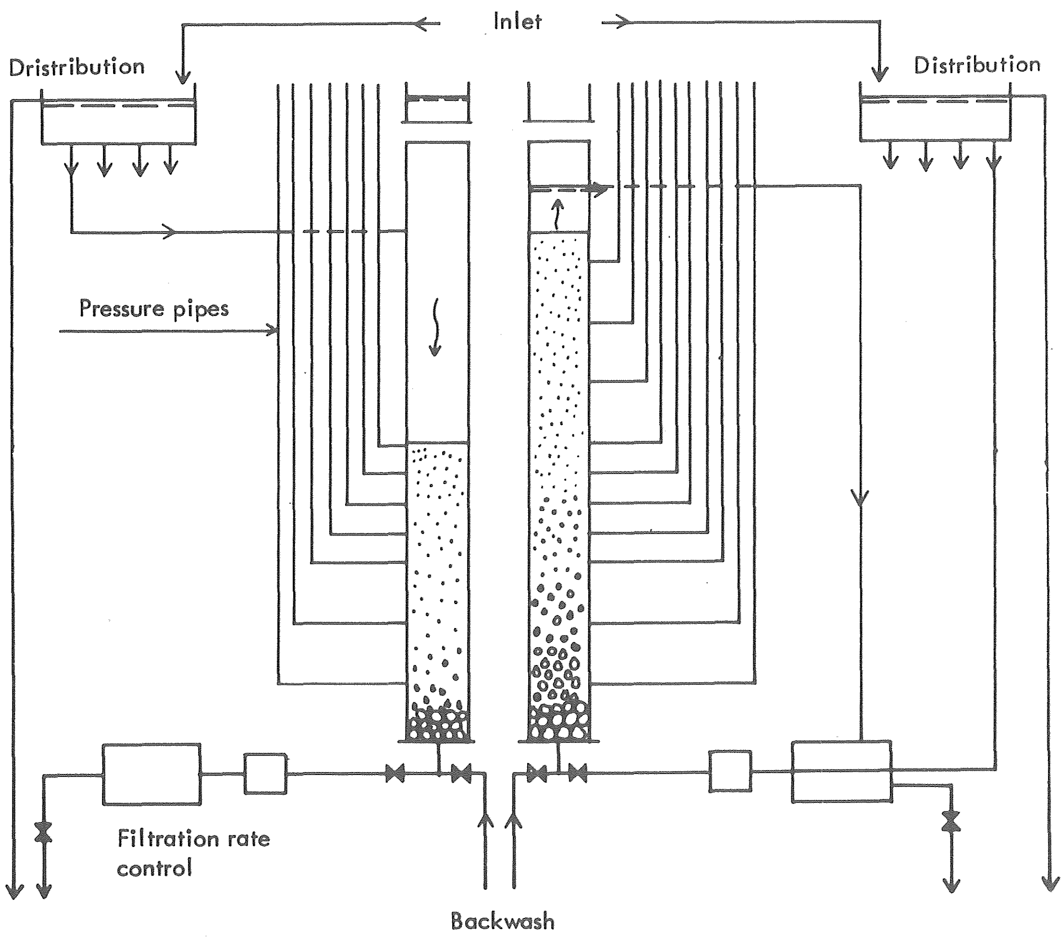


FIG. 5 Sketch of the filtration units

The pH-value was registered and adjusted automatically by lime addition controlled by the pH-meter. The turbidity of raw water, effluent from sedimentation, and filtrate was measured by a registering Sigrist turbiditymeter.

Data:

Chemicals:	Aluminium sulphate, activated silica, lime	
Flocculation:	Volume: 4 x 3.6 m^3	
Sedimentation:	Length of lamella	1.84 m
	Width of lamella	1.0 m
	Horizontal distance between lamellæ	0.1 m
	Inclination angle:	55°
	Number of lamellæ	5
Filtration:	Diameter	0.125 m
	Height	3.0 m
Equipment:	pH-control and registration	
	Turbidity registration	
	Pressure registration	

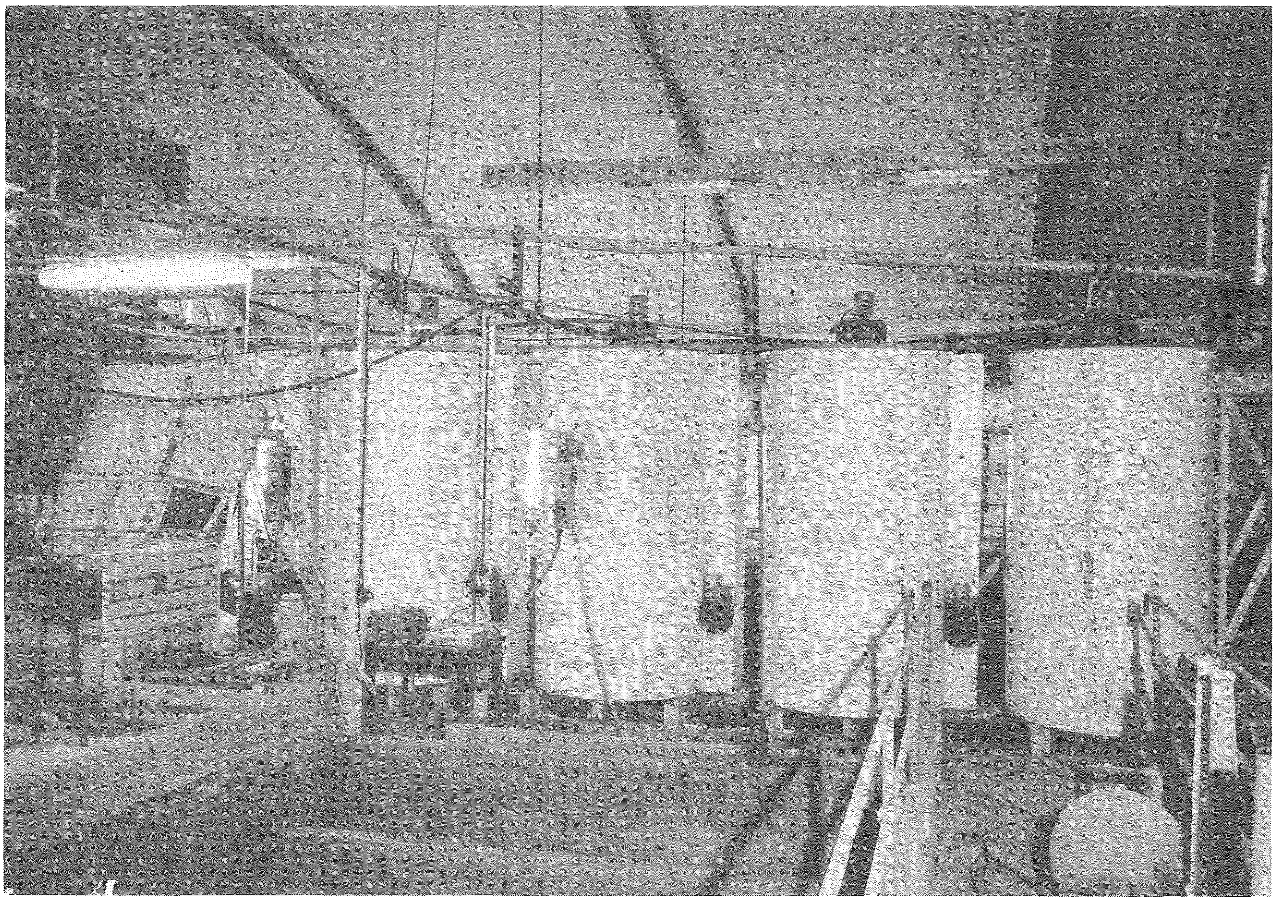


FIG. 6 Photograph of the pilot plant

Appendix 1

PLANT 2

Pilot plant, Alelyckan

A relatively large pilot plant was designed to study the lamella sedimentation. The pilot plant was in operation in 1967 at the Water Treatment Plant at Alelyckan. The flocculation unit and the sedimentation unit were placed in an older available plant which had been considered when designing the pilot plant. The plant capacity was 72 m^3/h.
The chemical dosage equipment consisted of equipment for dosages of aluminiumsulphate (50 ppm) and activated silica (8ppm) and automatic pH-control by use of lime.

The flocculation unit was designed for a constant flow with a residence time of 45 minutes, FIG. 1.

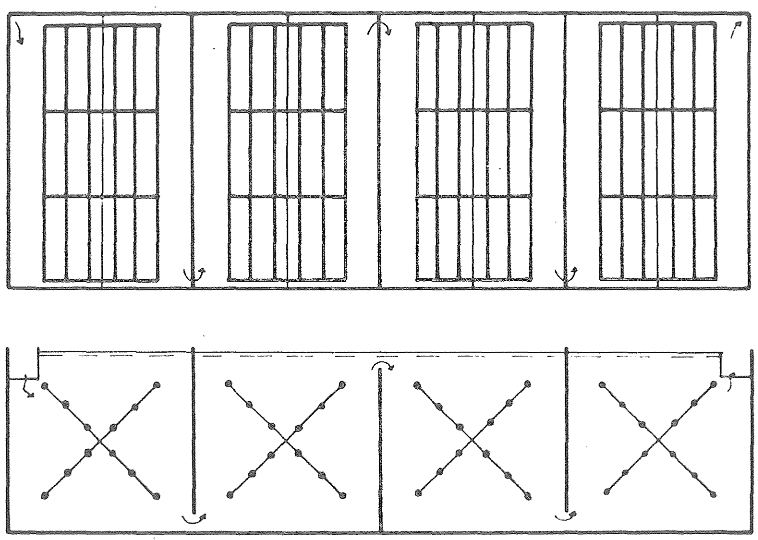


FIG. 1 The flocculation unit

The lamella sedimentation unit FIG. 2 consisted of two packages of lamellæ. One unit had lamella plates of corrugated plates and the other unit had plane plates. The distribution of water from the flocculation unit to the individual lamellæ was done from a centrally placed channel. Water was withdrawn at the surface through perforated pipes which simultaneously caused an even distribution of water to the individual lamellæ. After the sedimentation unit, the water was led to a filtration unit. Sludge was emptied intermittantly every 5 minutes, in 15 to 17 seconds by eight time-regulated valves. The water loss was estimated at about 2,5 %.

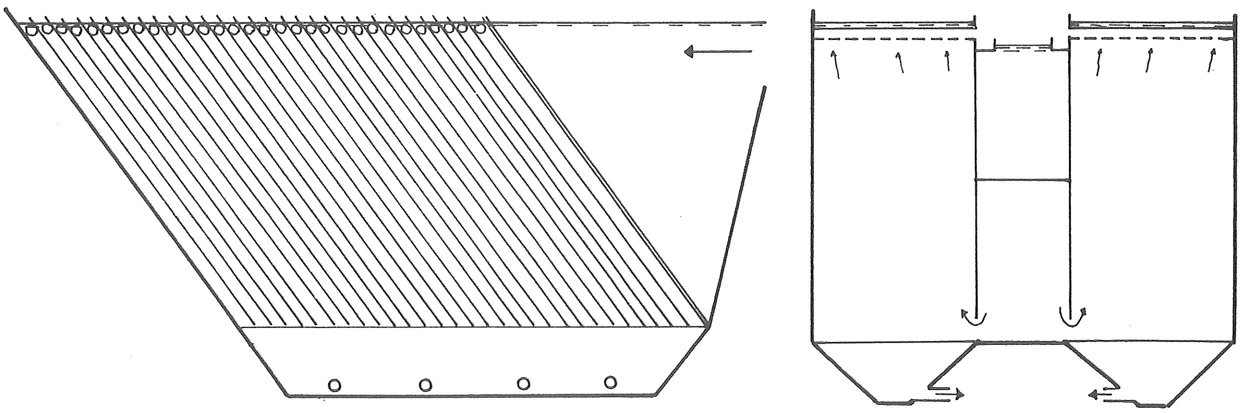


FIG. 2 Lamella sedimentation unit

Data:

Chemicals: Aluminium sulphate, activated silica and lime.

Flocculation: Volume 4 x 20 m^3
Length 3,8 m
Width 2,3 m
Height 2,3 m

Sedimentation: Length of lamella
Width of lamella
Horizontal distance between lamellæ 0,1 m
Inclination angle 55^0
Number of lamellæ 2 x 29

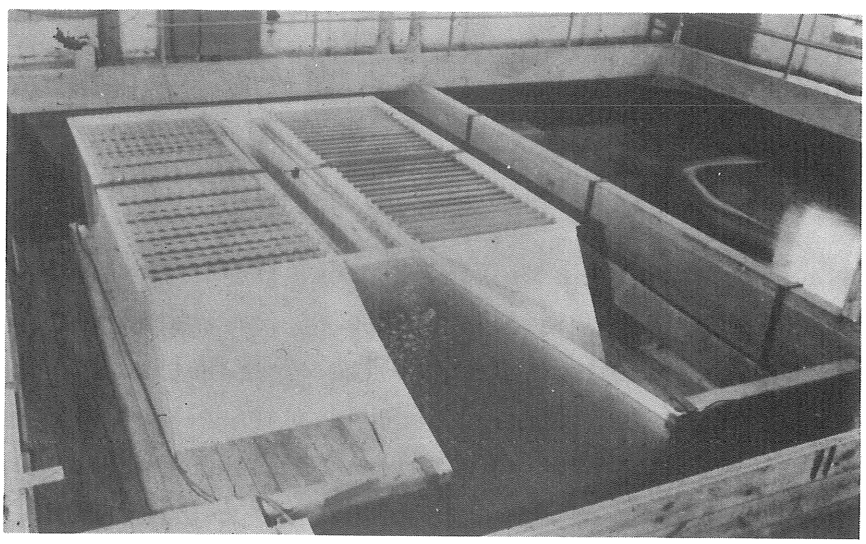


FIG. 3 Photograph of the pilot plant

Appendix 1

PLANT 3

Pilot plant, Lackarebäck

The pilot plant which was used at Alelyckan (Plant 1) was moved in 1969 when the Division of Water Supply and Sewage got the opportunity to use a new research station at the Water Treatment Plant at Lackarebäck. This pilot plant is thus identical to pilot plant 1 except for the design of the lamella sedimentation unit. The inlet system was changed from an inlet on only one side to an inlet on both sides, and the sludge pocket was increased in depth, FIG. 1

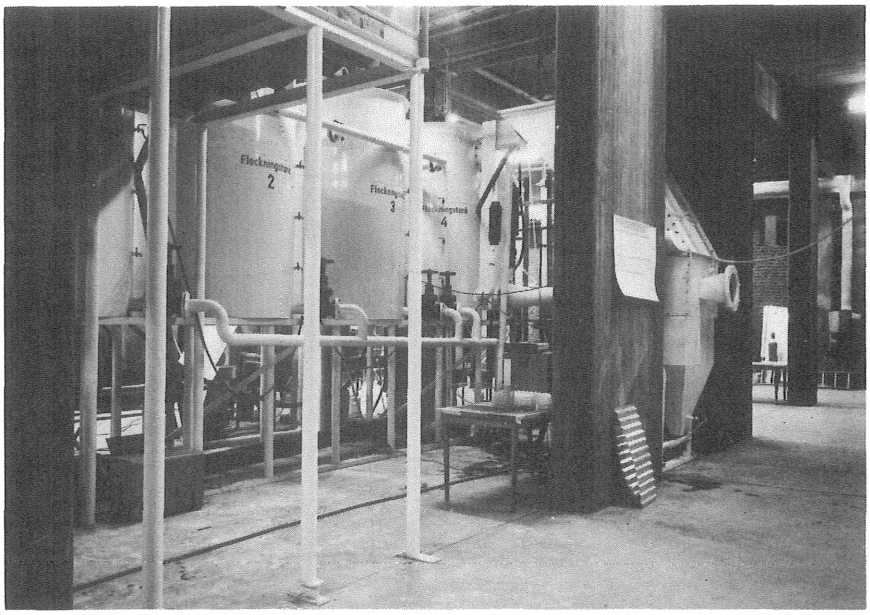


FIG. 1 Photograph of the pilot plant

Appendix 1

PLANT 4

Pilot plant, Bolmen

In connection with the South Water Project which entails delivery of water from lake Bolmen to the south of Sweden, the Division was offered a commission to study their water treatment process. The pilot plant was ready for operation in the beginning of 1971.

The pilot plant was designed in principle as the number 1 plant. The filtration unit was, however, larger. The pilot plant is shown in FIG. 1.

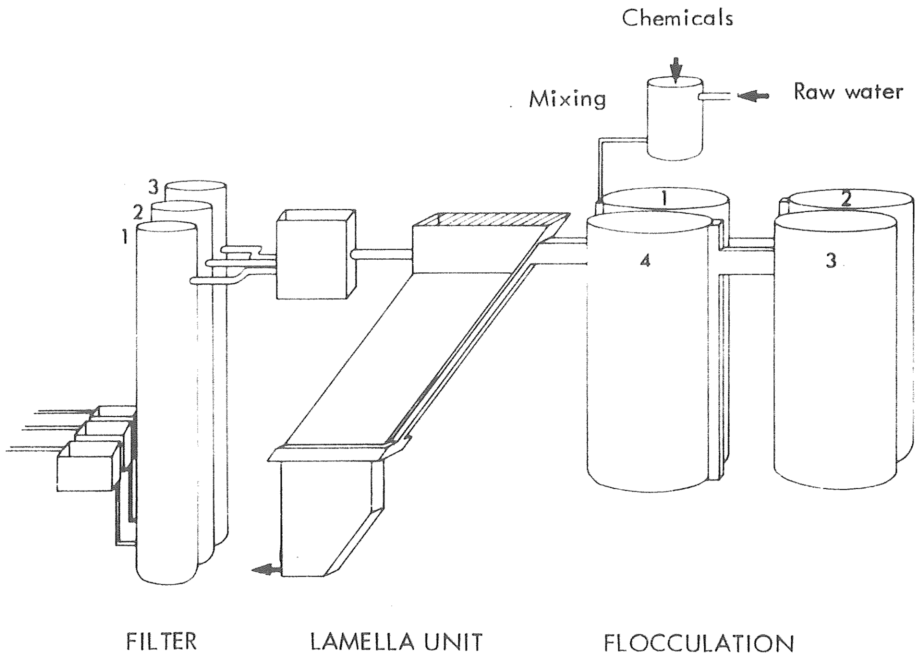


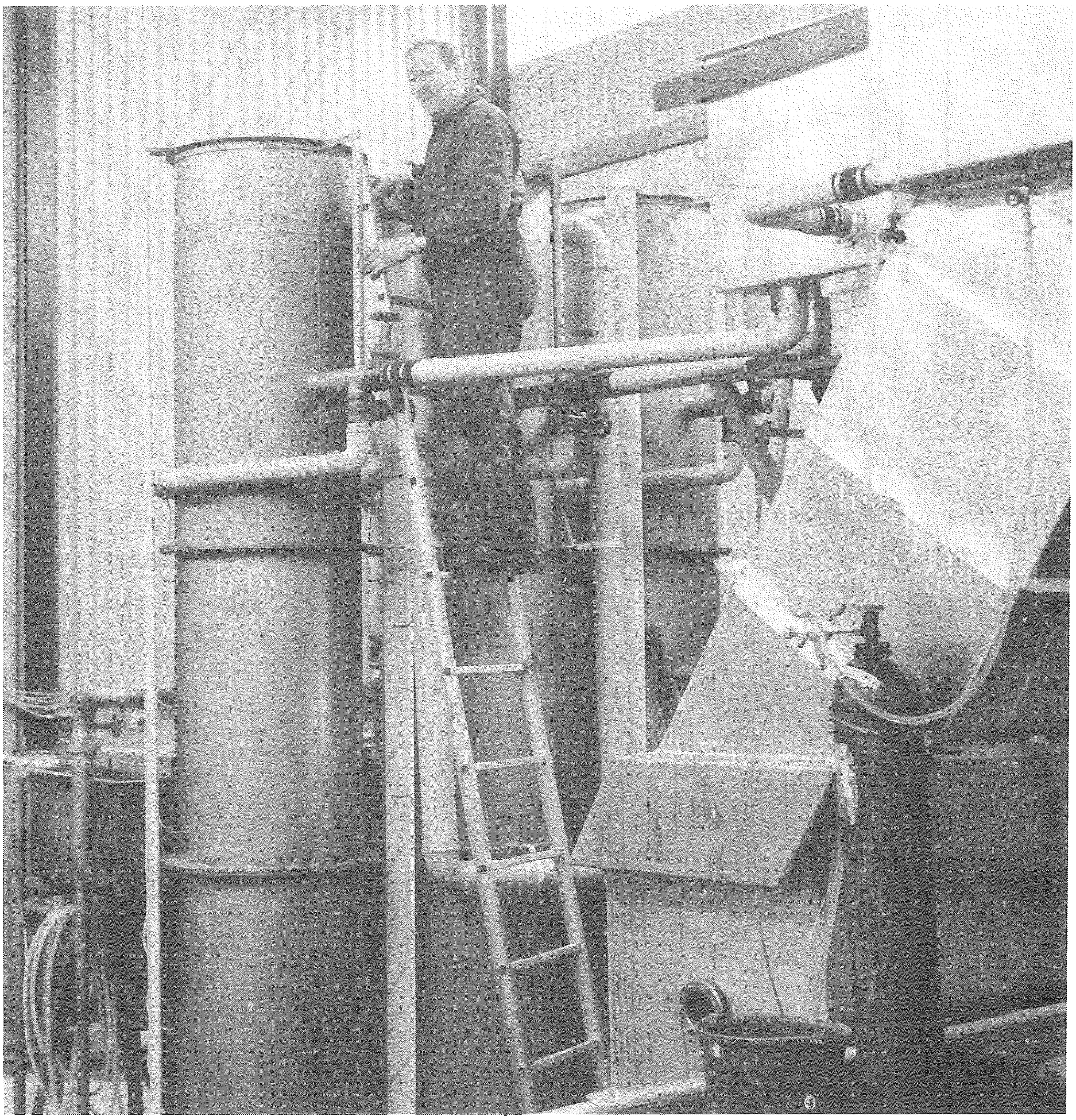
FIG. 1 Sketch of the pilot plant

The flocculation unit was similar to that in pilot plant No. 1. The lamella sedimentation unit was in principle similar, having an inlet system on both sides and the water withdrawn over an edge at the top.

Three filtration units were available for simultaneous tests with different filter designs. The flow of water could be varied and the maximum capacity was 28.8 m^3/h which corresponds to a residence time in flocculation of 27 minutes. The lamella sedimentation unit was designed for a surface rate of about 11 m/h. The filter units were designed for a filtration rate of maximum 25 m/h each.

Data:

Chemicals:	Aluminium sulphate, activated silica, superfloc, lime	
Flocculation:	Volume 4 x 3.25 m^3	
Sedimentation:	Length of lamella	2.5 m
	Width of lamella	1.0 m
	Horizontal distance between lamellæ	0.025 to 0.2 m
	Inclination angle	55^o
	Number of lamellæ	5 - 10
Filtration:	Filter diameter	0.6 m
	Height	3.5 m
Equipment:	pH-control and registration	
	Turbidity registration, Sigrist	



Appendix 1

PLANT 5

Pilot plant, Lackarebäck

In connection with flocculation studies, a smaller more flexible pilot plant was designed at the research station at Lackarebäck.

The pilot plant consisted of flocculation and sedimentation units, FIG. 1.

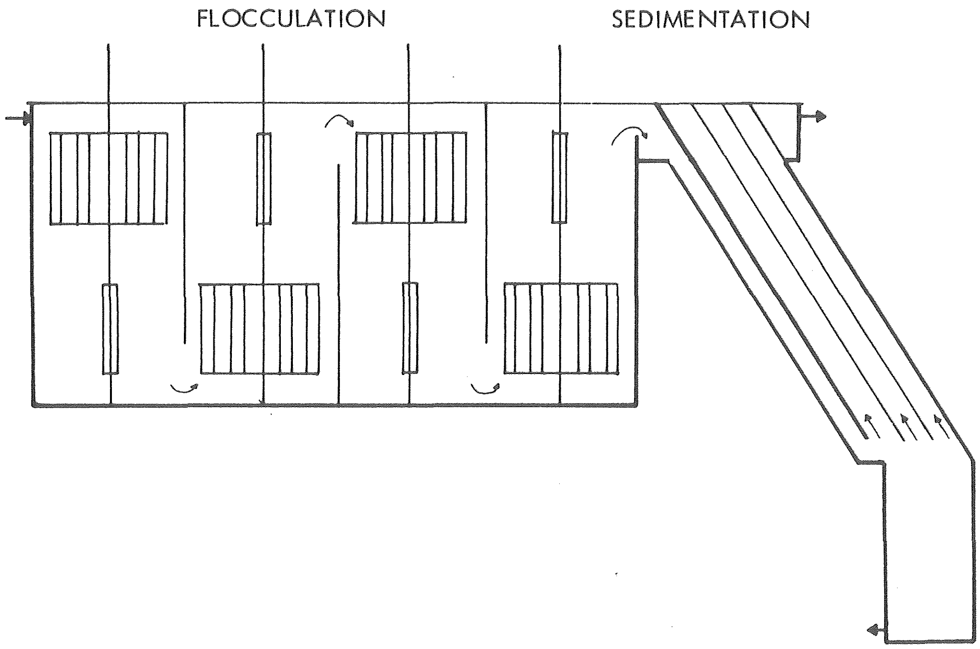


FIG. 1 Sketch of the pilot plant

The pilot plant was designed so that any number of reactors in series could be used and even the paddle design could be changed. In FIG. 2 the velocity gradient *G* calculated by the Camp formula is shown as a function of the stirring speed in revolutions per minute.

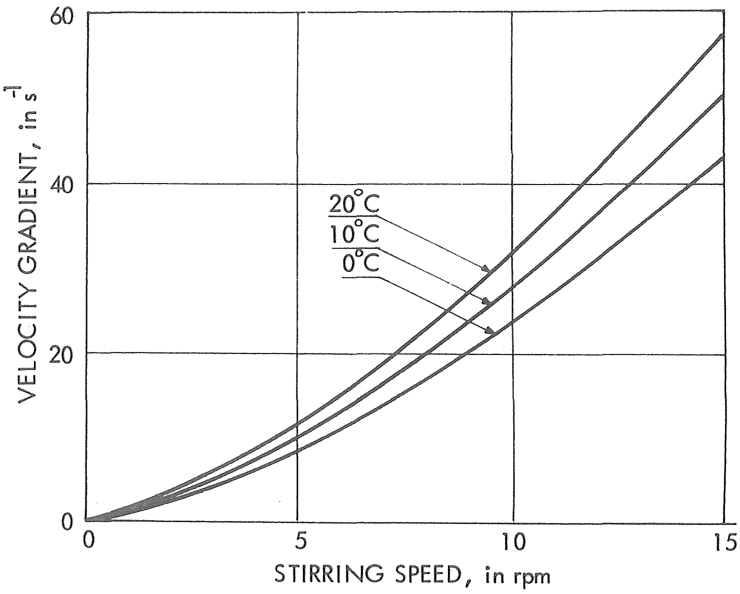
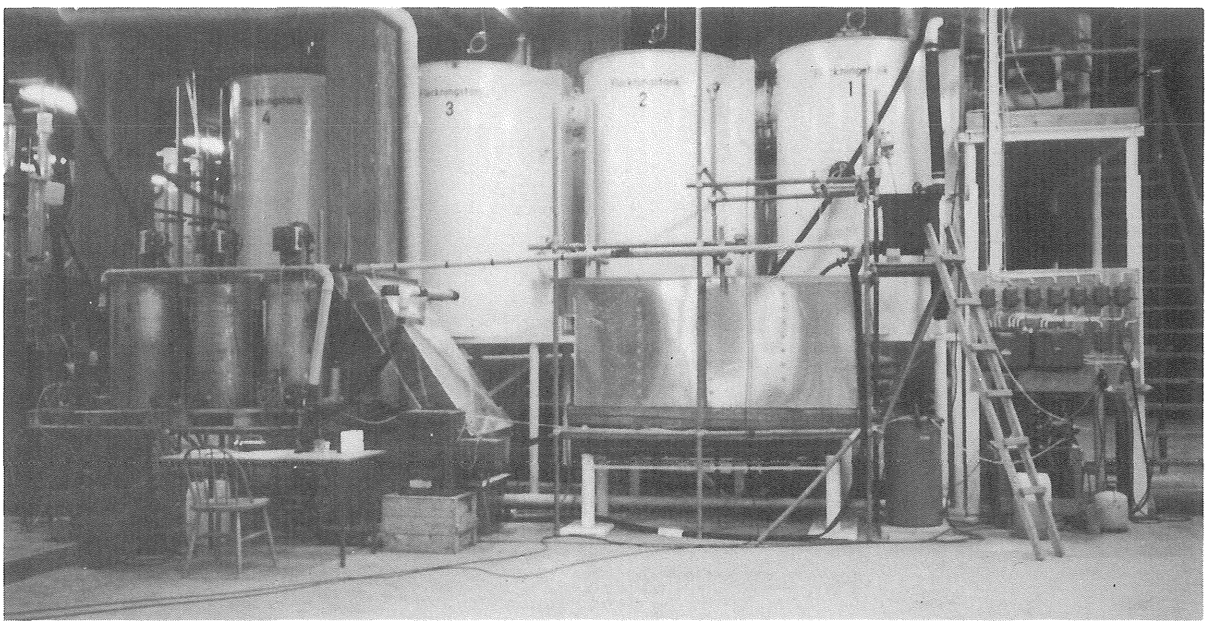


FIG. 2 The velocity gradient as a function of stirring speed

The lamella sedimentation unit was designed similar to previously mentioned units.

Data:

Chemicals:	Aluminium sulphate, activated silica, sodium-hydroxide, lime	
Flocculation:	Volume 1 m^3 divided in 1 to 4 tanks (Length 2.0 m, Width 0.5 m, Height 1.0 m)	
Sedimentation:	Length of lamella	1.5 m
	Width of lamella	0.8 m
	Horizontal distance between lamellæ	0.05 to 0.3 m
	Inclination angle	55^0
	Number of lamellæ	1 to 6



Appendix 1

PLANT 6

Full scale, Alelyckan Water Treatment Plant

During 1960-1968 the research activities were carried out at the Water Treatment Plant at Alelyckan, and some investigations were carried out concerning the sedimentation basins. These investigations were planned as a base for testing the new lamella sedimentation unit.

The Water Treatment Plant is a conventional plant with a maximum capacity of 160,000 m^3 a day.

The flocculation unit consists of 6 reactors in series with a total volume of 750 m^3. The residence time is about 45 minutes.

The sedimentation basins are of two different types, 4 Lovö-basins, volume 1,650 m^3, and Fischerström-basins, length 34 m, divided in 16 parallel horizontal channels.

The filters were 14 conventional sand filters. These filters were changed to active carbon filters in 1974.

Appendix 1

PLANT 7

Full scale, Lackarebäck Water Treatment Plant

The Water Treatment Plant at Lackarebäck, FIG. 1, gets the water from the river Göta älv. The water is then pumped up to lake Delsjön which functions as a water reservoir.

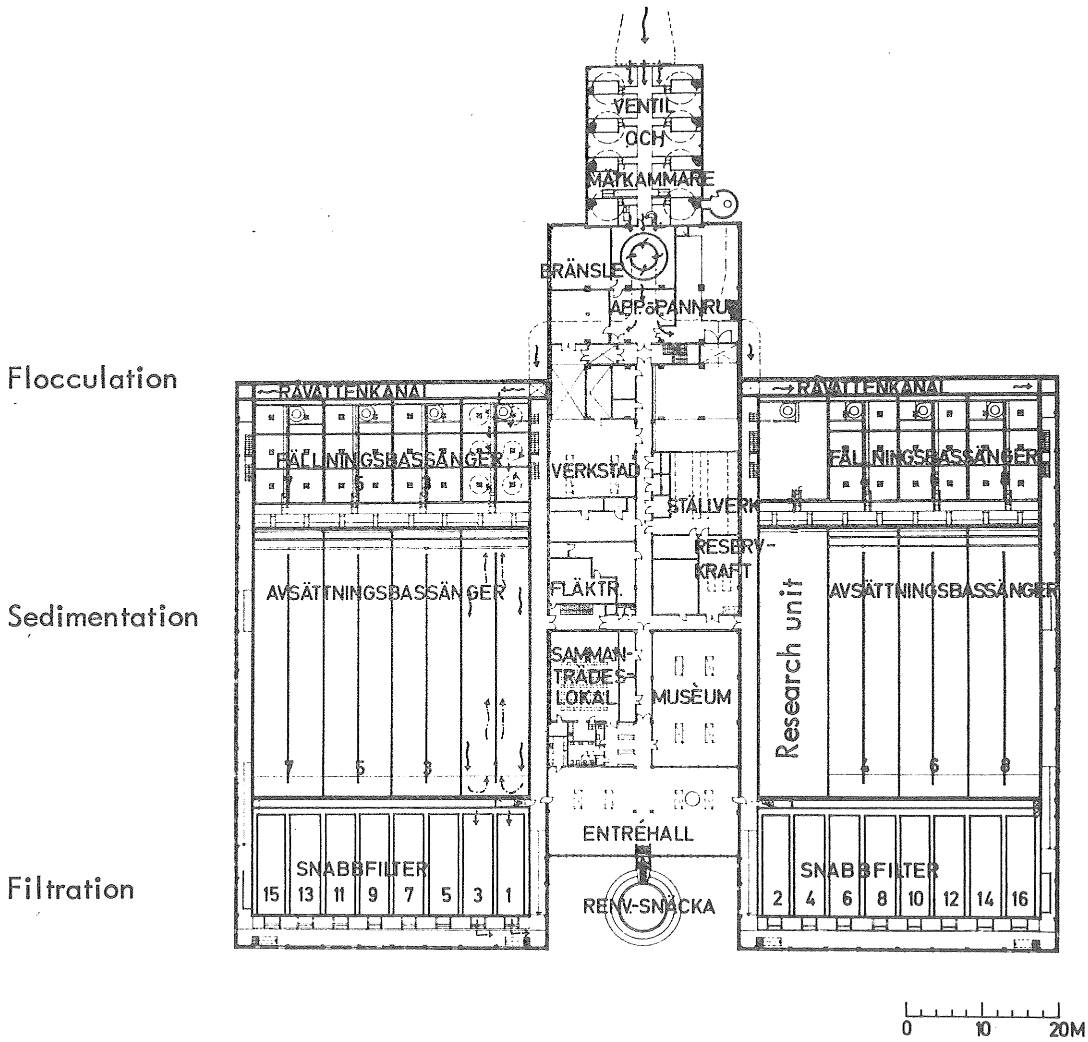


FIG. 1 Plan of the Water Treatment Plant at Lackarebäck. One of the treatment systems, system 2, is used for tests in full scale.

Each treatment system consists of a mixing device, six flocculation reactors in series, a sedimentation basin (Lovö-basin), and two filters. The flocculation unit has a total volume of 910 m^3

and this corresponds to a residence time of 1,25 h at a flow of 5,800 m^3 per hour. Each flocculation reactor, FIG. 2, measures 4,5 x 4,5 x 7,5 and is equipped with a stirring system of bars. In each reactor there are two baffles to eliminate short circuits.

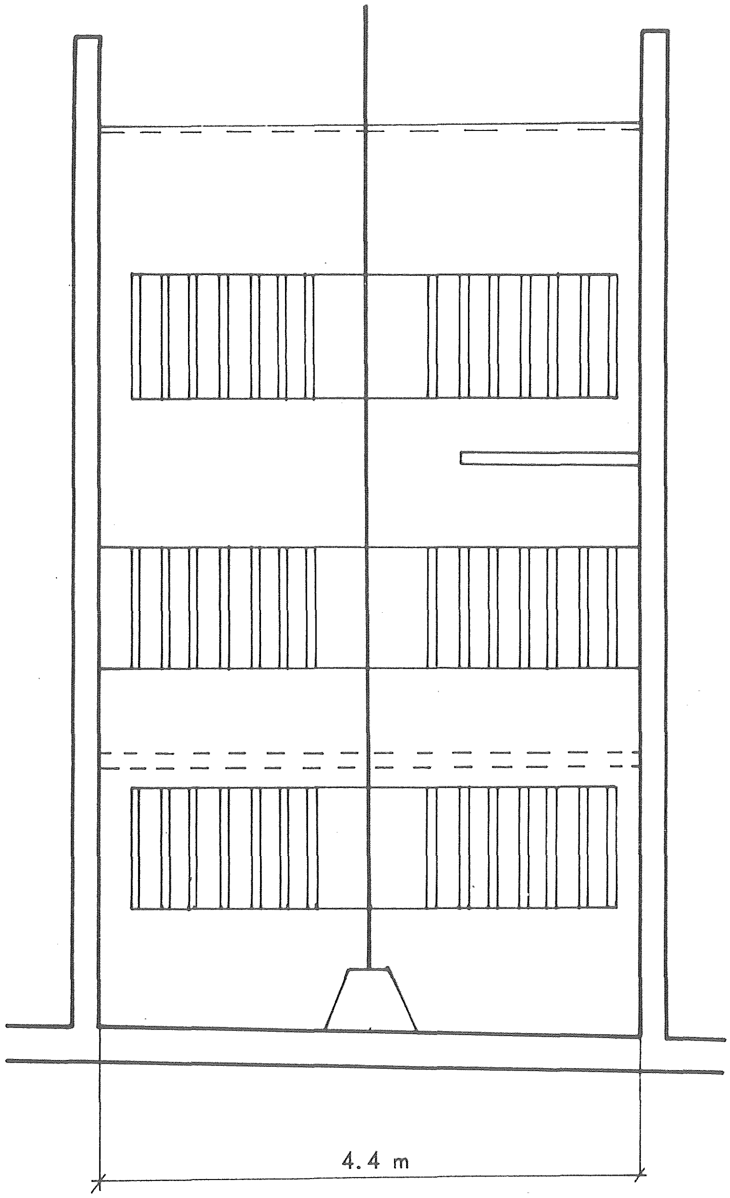


FIG. 2 Design of the flocculation unit

An approximate calculation of the energy input as a function of the stirring speed can be expressed as:

$$G = 0.275 \sqrt{\frac{n^3}{\mu}}$$

in which n is revolutions per minute
μ is the kinematic water viscosity

It has been assumed that the water velocity relative to the paddle (v_r) is half of the paddle velocity (v_p). The influence of fixed bars on the velocity gradient is thus equal to the rotating bars.

The sedimentation units, Lovö-basin, have a length of 35 m, a width of 9 m, and an average depth of 6 m. The basins are emptied of sludge every third or fourth week.

The filtration operation is performed with 16 filters, each 50 square meters. The filter media was originally one meter of sand but this has been changed to active carbon. The filters used to be washed with water and air but nowadays are only washed with water.

Appendix 1

PLANT 8

Pilot plant, full scale, Lackarebäck Water Treatment Plant

In the Water Treatment Plant at Lackarebäck one of the 8 treatment systems was reserved for full scale tests concerning flocculation and sedimentation. During 1968 to 1974 the pilot plant design, especially the flocculation unit, has been changed and the lamella sedimentation has also been modified to some extent.

Originally the flocculation unit was only one big tank (910 m^3) equipped with a centrally placed Vortifloc type stirrer. Later on, 1972, the flocculation unit was rebuilt so that it was equal to the other flocculation units at the plant but without any horizontal baffles, FIG. 1.

The lamella sedimentation unit consisted of four packages of lamellæ with 56 lamellæ in each. The length of the lamella was 3,0 m and the width 1,25 m. The inclination angle was 55^{0} and the horizontal distance between the lamellæ was 0,08 m. The depth of the sludge pocket under the lamella package was 2,8 m and the sludge was emptied in 15 seconds at 30 minute intervals.

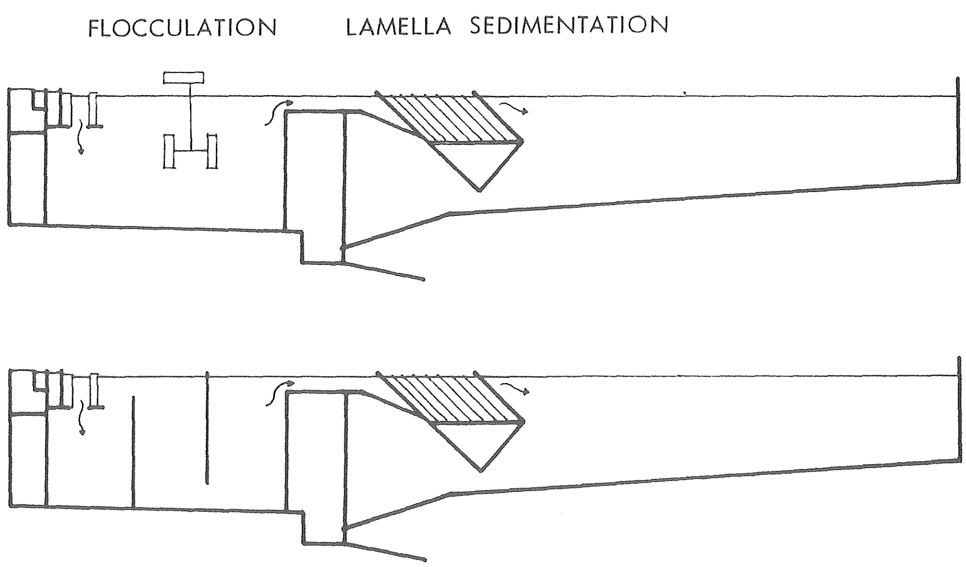


FIG. 1 Section of flocculation and sedimentation units

In 1972 when the flocculation unit was changed even the inlet channels in the lamella sedimentation were changed and each package of lamellæ got its own channel. The bottom of the channel was inclined as shown in FIG. 2 in order to prevent sludge deposits.

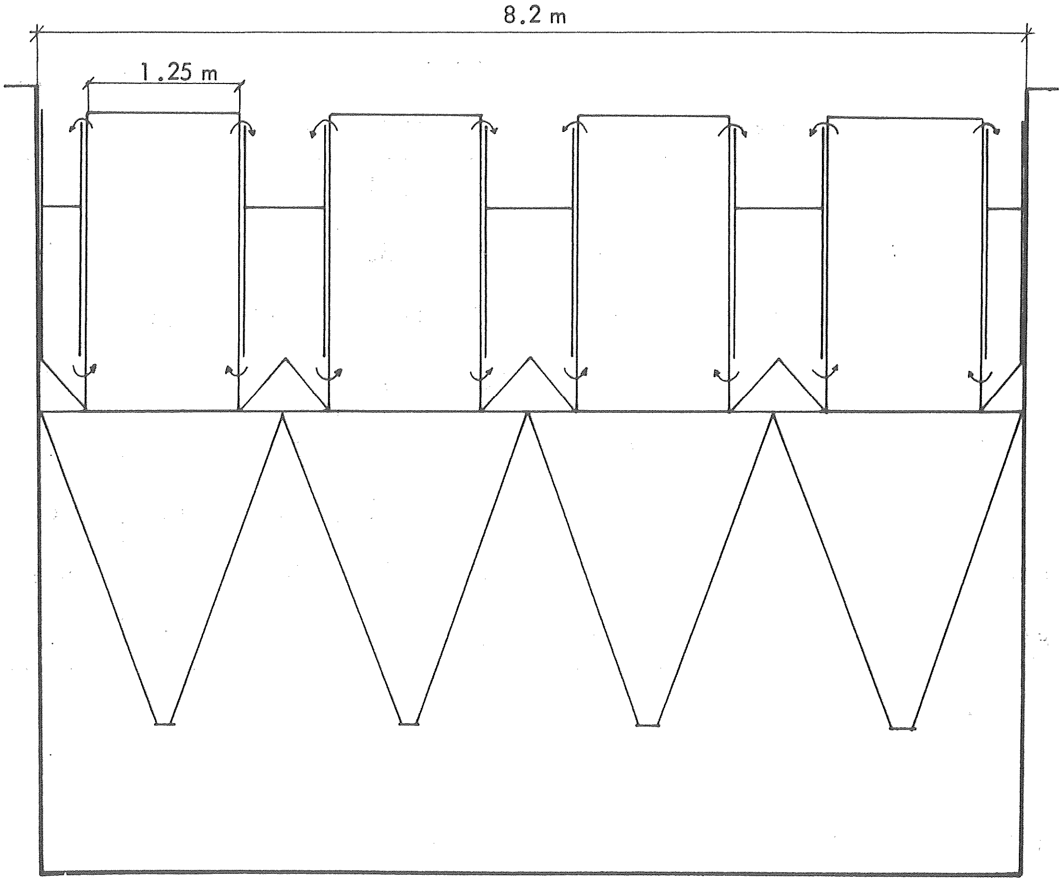


FIG. 2 Lamella sedimentation

APPENDIX 2
RAW WATER CHARACTERISTICS

In the investigation raw waters from the Göta älv river, lake Delsjön and lake Bolmen have been studied.

Some characteristic analyses

Mean values during the year	GÖTA ÄLV [1] 1965	1970	1974	DELSJÖN [1] 1970	1974	BOLMEN[2] 1972
Temperature ^{o}C	71	8.2	7.1	9.1	7.8	10
Colour mg/l Pt	38	41	36	28	30	24
Turbidity ZP-units	1395	1109	1800	192	220	100
COD ppm $KMnO_4$	52	58	46	41	33	24
pH-value	6.9	6.6	6,9	6.7	7.0	6.7
Conductivity µS/cm	85	87	96	102	107	60
Hardness mg/l Ca	9.0	10.1	11	9.9	11.0	8.9
Alkalinity mg/l HCO_3	12	9	10	11	12	5,6

References.

1) Göteborgs VA-verk. Berättelse för år 1960...
2) Bolmen-Bolmån-Lagan, Hamrin,S, Limnologiska institutionen, Lund. (1973).

APPENDIX 3
ORGINAL DATA
Flocculation

In studying the flocculation process several different pilot plants have been used. Some of the research carried out has been described previously. In these earlier reports a large amount of original data has been given and reports from which data has been taken are referred to where necessary.

Hedberg,T.: Optimering av den kemiska reningsprocessen samt hårdhetshöjning av mjuka vatten.

Optimization of the water treatment process and increase of the water hardness.
Publikationsserie B69:1.

Hedberg,T.: Studium av flockningsprocessen.

The flocculation process.
Publikationsserie C74:1.

Hedberg,T.: Studium av flockningsprocessen. Del 2: Undersökningar.

The flocculation process. Part 2: Experimental investigations.
Publikationsserie B75:2.

Hedberg,T.: Kemisk rening av vatten från Bolmen, Rapport från försöksverksamhet 1971-1973.

Chemical treatment at water from the lake Bolmen, Report from investigations 1971-1973.
Sydvatten, 1974.

Apart from these investigations, some studies for the Master of Science Engineering examination have been carried out under supervision of the author mainly at the research station at Lackarebäck (Plant 3).

Examensarbete 1971:9

Slamåterföring vid rening av vatten. – Sludge recycling in water treatment.

Examensarbete 1972:1

Undersökning av flockning, sedimentering och filtrering i full skala vid vattenverket i Lackarebäck. – Examination of flocculation, sedimentation and filtration in full scale at Lackarebäck Water Treatment Plant.

Examensarbete 1972:2

Optimeringsaspekter på den kemiska reningsprocessen för försörjningsvatten. – Optimization aspects on the chemical treatment process.

Examensarbete 1973:1

Studium av flockningsprocessen. – The flocculation process.

APPENDIX 4
ORGINAL DATA
Sedimentation analyses

PLANT 1

The following mean values of the settling velocity of flocs are taken from Rosén (1967): "The influence of input power on flocculation. Sedimentation analysis as a measure of flocculated water". The original data in this report have been used to test the hypothesis suggested.

MEAN SETTLING VELOCITY, v_{fm}, in m/h; C_o in rel. %.

Sedimentation depth 0.3 m

FIGs 39, 40
$v_{fm} = 3.5$ $c_o = 7$ (temperature 2.8 - 9°C)
$v_{fm} = 2.5$ $c_o = 14$

FIGs 41, 42
$v_{fm} = 4.0$ $c_o = 9$ (temperature 14 - 15°C)
$v_{fm} = 2.2$ $c_o = 17$

Sedimentation depth 1.0 m

FIG. 29
$v_{fm} = 3.0$ $c_o = 7$ (temperature 9 - 10°C)

FIGs 32, 33
$v_{fm} = 3.0$ $c_o = 11$ (temperature 8°C)

Chemicals: Aluminium sulphate 50 ppm
Activated silica 8-9 ppm
Lime to pH 6.0

APPENDIX 4

PLANT 7 and PLANT 8

Sedimentation depth 0.3 m

Date	Turbidity ZP units c_{100}	c_o	t_o min.	t_{10} min.	t_m min	System
700306	1174	700	20	15	5	2
	1194	721	35	32	13	2
	1256	300	35	22,5	7	4
	1297	330	32.5	17.5	7.5	4
701104	530	380	20	12	4	2
	670	130	8	7	4	4
701105	560	210	26	16	7,5	2
	530	125	16	8	6	4
701109	550	245	22	16	6	2
	630	150	18	10	3	4
701110	730	275	26	18	3	2
	600	140	18	10	3	4
701111	670	260	26	18	8	2
	730	155	26	18	4	4
701112	620	245	22	18	7	2
	740	210	20	11	5	4
701113	700	215	28	19	10	2
	630	150	20	10	4	4
701116	610	220	26	18	9	2
	670	175	22	14	5	4
701117	550	270	28	17	7	2
	800	180	18	10	3	4
701118	550	265	28	19	6	2
	760	135	16	9	3	4
701124	575	125	26	12	5	4
701125	875	145	24	15	6	4
701126	780	270	31	22	8	2
	660	130	22	13	6	4

APPENDIX 4

PLANT 7

Sedimentation depth 0.3 m

No.[1]	Date	Q m^3/h	Temp. ^oC	Turbidity ZP units c_{100}	c_o	t_o min.	t_5 min.	t_{10} min.	t_m min.
201	8.6	540	11,8	540	220	38,0	30,5	27,5	12,5
202	9.6	468	12,0	580	415	30,0	24,0	21,5	9,5
203	14.6	538	12,5	620	-	33,5	27,0	21,0	10,5
204	15.6	537	12,8	600	320	30,5	28,0	25,5	13,0
205	16.6	538	14,2	600	-	20,0	16,5	15,5	10,0
206	17.6	538	14,5	620	190	43,0	32,0	28,0	13,5
207	17.6	538	14,4	-	140	35,0	26,0	21,5	9,0
208	18.6	538	12,8	640	195	44,5	36,0	30,0	14,5
209	18.6	538	12,9	600	200	39,0	33,0	30,5	17,0
210	22.6	780	13,2	460	195	44,0	33,0	28,0	15,0
211	23.6	501	13,2	560	200	32,5	28,0	23,5	14,0
212	29.6	540	14,0	520	385	37,5	32,5	29,5	21,5
213	1.7	541	14,5	620	415	28,5	23,5	21,5	11,5
301	14.6	468	12,5	655	425	29,0	24,0	21,5	12,0
302	17.6	538	14,5	-	115	35,0	27,5	23,5	13,0
303	18.6	539	12,9	690	185	43,0	30,5	25,0	12,5
304	18.6	538	12,8	530	135	35,0	28,5	25,5	14,5
305	20.6	386	13,1	405	160	47,5	34,0	29,0	11,0
306	21,6	472	13,0	590	135	46,5	34,5	30,5	14,5
307	21.6	781	13,1	480	145	-	-	-	-
308	22.6	780	13,2	430	140	45,0	33,0	28,5	14,0
309	24.6	435	13,5	640	195	31,5	24,0	21,5	11,0
401	8.6	538	11,8	500	185	35,0	22,0	17,5	6,0
402	9.6	465	11,8	520	125	32,5	21,0	17,5	6,5
403	9.6	469	12,1	505	130	33,5	25,0	20,5	8,0
404	15.6	537	12,5	585	-	25,0	16,0	13,5	7,0
405	17.6	538	14,3	600	100	35,0	21,0	18,5	7,5
406	18.6	538	12,8	620	100	35,0	24,5	20,0	8,0
407	18.6	537	12,8	670	115	35,5	25,0	20,5	8,0
408	20.6	386	13,0	600	95	44,0	33,5	30,0	8,5
409	21.6	778	13,0	-	110	37,5	26,0	21,5	11,0
410	22.6	778	13,1	535	125	41,5	28,5	24,0	10,0
411	23.6	540	13,4	540	120	33,0	26,5	22,5	9,5
412	24.6	442	13,5	600	135	41,5	32,5	28,5	12,5
413	30.6	540	14,5	620	175	42,5	32,5	27,5	8,5
501	14.6	538	13,0	-	135	37,0	29,0	24,5	11,0
502	16.6	538	14,0	600	140	35,0	26,5	22,5	9,0
503	16.6	468	14,1	480	130	41,0	32,5	27,5	8,0
504	17.6	539	14,5	540	85	30,0	17,0	15,0	7,5
505	17.6	536	14,2	470	89	26,5	19,5	16,5	8,0
506	18.6	538	12,7	540	85	51,0	39,5	33,0	10,0

No.[1]	Date	Q m^3/h	Temp. ^oC	Turbidity ZP units c_{100}	c_o	t_o min.	t_5 min.	t_{10} min.	t_m min.
507	18.6	538	12,9	480	105	24,0	18,0	16,0	7,5
508	20.6	386	13,0	435	83	42,5	32,5	25,0	6,0
509	21.6	778	13,1	405	100	45,0	29,0	24,5	8,0
510	22.6	775	13,2	435	140	46,0	26,0	22,5	10,0
511	24.6	442	13,5	485	150	40,0	32,5	28,0	13,5
512	29.6	631	14,0	540	180	45,0	36,0	29,5	12,5
513	30.6	457	14,5	415	175	35,0	25,0	21,5	8,5
514	1.7	541	14,5	480	210	40,0	25,0	22,5	12,0
601	8.6	467	12,0	540	125	35,0	18,0	13,5	2,5
602	9.6	467	11,8	435	135	30,0	17,5	12,5	4,0
603	9.6	465	12,0	600	89	35,0	23,0	18,0	5,5
604	14.6	537	12,5	680	-	37,5	27,0	23,5	12,5
605	15.6	539	12,7	580	155	42,5	30,5	25,0	9,0
606	16.6	539	14,0	600	110	42,0	24,5	19,5	6,0
607	16.6	465	14,0	670	115	34,0	25,0	20,5	8,5
608	17,6	538	14,4	730	73	25,0	13,5	9,0	3,0
609	17,6	536	14,2	460	52	32,5	16,5	13,0	5,5
610	18.6	539	12,8	640	89	26,0	19,0	16,0	6,0
611	18.6	538	12,8	620	94	35,0	25,5	22,5	7,5
612	20.6	386	13,0	-	115	32,5	22,5	19,0	8,0
613	21,6	778	13,0	500	94	41,5	28,5	24,5	11,0
614	22.6	775	13,2	450	98	42,5	29,5	24,5	8,5
615	23.6	501	13,2	610	100	32,5	22,5	19,5	7,5
616	24.6	442	13,5	-	105	34,0	22,0	19,0	7,0
617	30.6	539	14,5	520	175	42,5	29,0	22,5	7,0

[1] = The figures mean: The first number refers to the flocculation tank. The following figures refer to the numbers of the test.

APPENDIX 4

PLANT 4

Sedimentation depth 0,12 m

Date	Temp. °C	Turbidity Sigrist Std 5 %		Floc settling velocity m/h		Remark
		c_{100}	c_o	v_{fo}	v_{fm}	
710414		240	70	0,8	1,8 [1]	
710414		240	75	1,6	3,1 [1]	0,95 m sed.depth
710415		240	95	0,7	1,2 [1]	
710415		240	75	1,9	3,0 [1]	0,95 m sed.depth
710415		240	70	0,6	0,9 [1]	
710416		240	65	0,6	1,1 [1]	
710804		240	85	0,7	1,3 [2]	
710805		240	95	0,7	1,2 [2]	
720317		240	32	1,0	2,4 [3]	
720413		240	27	0,8	1,8 [3]	
720728		240	27	0,8	-- [3]	
730417		240	30	0,7	1,1 [1]	

[1] = No coagulant aid added

[2] = 0,1 ppm superfloc

[3] = 2,5-3,5 ppm activated silica

APPENDIX 4

PLANT 5

May 71, 1971 to June 22, 1971

Sedimentation depth 0.15 m

Test No.	Floccula-tion time min.	Numbers of reactors	Turbidity ZP units c_{100}	c_o	t_o min.	t_{10} min.	t_m min.
Sludge recycling 0 %							
1:1	35	4	730	145	28	20	9
2:1	60	4	675	135	25	18	11
3:1	30	2	680	225	23	17	8
4:1	60	2	580	175	19	12	7
5:1	30	1	560	415	99	14	8
6:1	60	1	595	240	16	10	5
Sludge recycling 2 %							
1:2	35	4	2400	145	20	14	8
2:2	60	4	2250	140	19	12	6
3:2	30	2	1850	165	16	10	6
4:2	60	2	1950	135	13	7	4
5:2	30	1	1900	230	18	9	5
6:2	60	1	1000	150	15	7	5
Sludge recycling 4 %							
1:3	35	4	2900	145	20	--	-
2:3	60	4	3800	115	17	10	4
3:3	30	2	5500	190	12	--	-
4:3	60	2	3000	120	17	7	5
5:3	30	1	2450	195	15	9	5
6:3	60	1	2600	155	17	7	4

Chemicals: Aluminium sulphate 40 ppm
Activated silica 0 ppm
Lime to pH 6,1 - 6,2
Sludge recycling in percentage of flow
(Dry solids in the sludge 0,7 %)

APPENDIX 4

PLANT 5

72-02-01 -- 72-08-01

Sedimentation depth 0.15 m — Stationary test, Q = 0

Flocculation time min.	Turbidity ZP units c_{100}	c_0	t_0 min.	t_{10} min.	t_m min.
10	1300	480	30	18	9
20	1350	330	36	18	10
25	1000	115	30	14	5,5
30	1200	165	27	18	8
35	960	71	25	13	4,5
40	1300	110	25	14	6
50	1000	60	25	10	4,5
Flow test - 4 reactors in series					
20	810	115	30	20	10
25	690	115	30	20	9
29	770	105	30	20	8
33	690	100	30	16	9
37	770	125	30	22	9
45	710	79	30	13	8
54	670	62	30	18	9
64	670	52	30	13	7

Chemicals: Aluminium sulphate 40 ppm
Lime to pH 5,9-6,1

APPENDIX 4

PLANT 5

73-12-13 -- 74-01-11

Sedimentation depth 0.30 m

Date	Temp. ^oC	Flocculation time min.	Turbidity FTU c_{100}	c_o	t_o min.	t_{10} min.	t_m min.	Re-mark
731213	3,7	30	9,5	1,1	20	6,5	2,5	1
	3,7	30	9,5	3,5	20	13	5	1
	3,7	30	9,5	2,3	20	15	6	1
731214	3,7	40	9,3	1,0	20	10	2,5	1
731220	3,4	40	10,0	2,5	22	19	12,5	2
740104	3,3	40	8,2	1,1	20	13	5	2
740111	3,4	40	9,7	2,0	20	13	7	2

[1] = Stationary test, Q = 0

[2] = Flow test, 4 reactors in series

Chemicals: Aluminium sulphate 40 ppm
Activated silica 6 ppm
Sodium-hydroxide to pH 6,2

APPENDIX 5
ORGINAL DATA
Conventional sedimentation

PLANT 6

Alelyckan Water Treatment Plant

During the investigation of the lamella sedimentation two conventional sedimentation basins were also tested, one Lovö-basin and one Fischerström-basin. The original data from these investigations are in tabular form in the Publication B63:3. Further treatment of this material has lead to the following FIG. 1 in which three different temperature intervals are separated. The result thus obtained differs quite markedly from that presented in the report. No definite breaking point in capacity of the Lovö-basin can be observed.

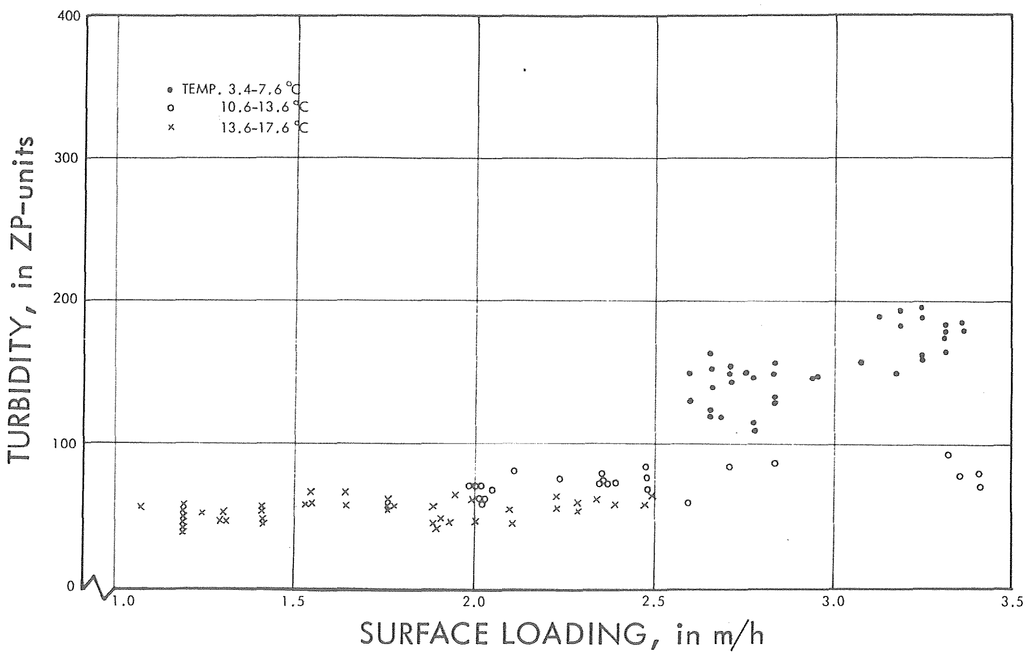


FIG. 1 The residual turbidity as a function of surface loading at different temperatures.

APPENDIX 6
ORGINAL DATA
Lamella sedimentation and
conventional sedimentation

Lamella sedimentation has been studied in different scales in order to test the performance. The reports from which information has been taken are referred to here:

Weijman-Hane,G.: Koagulering, sedimentering och filtrering. Försök med lamellsedimentering.

Coagulation, sedimentation and filtration. Experiments with lamella sedimentation.
Publikationsserie B63:3.

Rosén, B.: Koagulering, sedimentering och filtrering.

Experimentals with lamella sedimentation.
Publikationsserie B66:4.

Hedberg, T.: Optimering av den kemiska reningsprocessen samt hårdhetshöjning av mjuka ytvatten.

Optimization of the water treatment process and increase of the water hardness.
Publikationsserie B69:1

Hedberg, T.: Studium av flockningsprocessen.

The flocculation process.
Publikationsserie C74:1.

Hedberg, T.: Studium av flockningsprocessen. Del 2: Undersökningar.

The flocculation process. Part 2: Experimental investigations.
Publikationsserie B75:2.

Hedberg, T.: Kemisk rening av vatten från Bolmen. Rapport från försöksverksamhet, 1971-73.
Chemical treatment of water from the lake Bolmen.
Report from investigations, 1971-73.
Sydvatten, 1974.

Lamella sedimentation studies at Alelyckan Water Treatment Plant will now be presented.

PLANT 2, ALELYCKAN

The scope and result of the experimental investigation.

While studying lamella sedimentation in this relatively large pilot plant the flow through the flocculation unit was constant at 108 m^3/h. The overflow rate in the sedimentation unit was varied. As a measure of the removal efficiency, turbidity measurements in the effluent water have been performed. The result is summarized in FIG. 1 and the curve shows an average value. The investigation carried out from January to October 1967 shows that the maximum surface loading of the actual lamella sedimentation unit was about 15 m/h to 20 m/h.

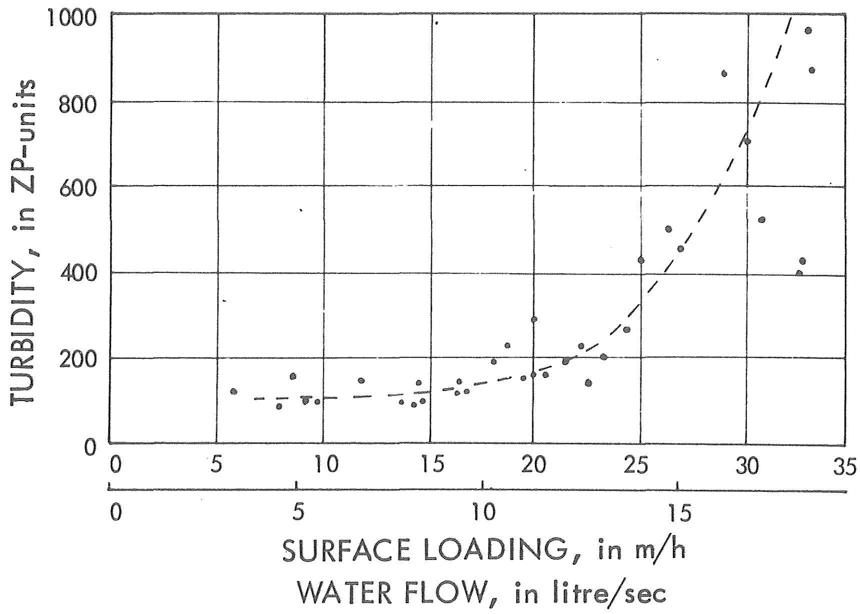


FIG. 1 Residual turbidity as a function of flow surface loading

In order to study closely the distribution of water to the individual lamellæ, measurements of the rate of flow in between each lamella were carried out by use of a current meter. Simultaneously, samples were taken for turbidity measurements. The measurements were performed at different capacities. The result is shown in FIG. 2.

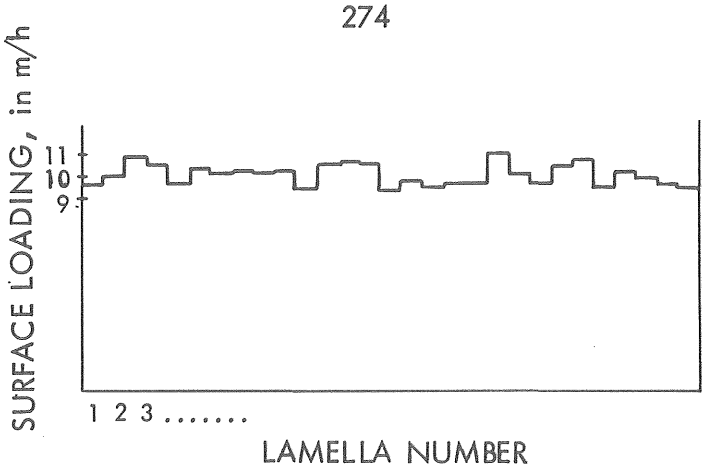


FIG. 2 Surface loading in the individual lamallae. Mean surface loading of the whole lamella unit was 10 m/h.

The results obtained show that the distribution of water to the lamellæ was even. The variations in turbidity between the individual lamellæ were relatively higher than the variations in rate of flow which might be related to difficulties in sampling and measuring.

In order to study the water distribution over the lamella plate, tests were performed in a lamella unit with only one lamella plate but with an equal inlet and outlet system as in the pilot plant. A tracer, dye, was injected at different places and the functioning lamella plate surface could be estimated at about 60 %.

The sludge pocket under the lamella package had to be, due to practical reasons, designed with a very small depth. The pocket was equipped with eight time controlled sludge valves. Sludge was emptied during 16 sec. intervals every 5th minute. The water loss was calculated at 2,5 %

PLANT 7 and 8, LACKAREBÄCK WATER TREATMENT PLANT

At the Lackarebäck Water Treatment Plant a conventional sedimentation basin, – Lovö-basin, and a lamella sedimentation unit were compared in full scale. A brief report of this investigation is given in the following.

In the research treatment system in the Water Treatment Plant a lamella sedimentation unit was built in 1969. The pilot plant was tested and ready for operation in March 1970. The main test program was started in June 1970 and the first real tests were carried out during June 4, 1970 to September 8, 1970. The results are shown in TABLE 1. The flow of water was varied between 285 - 600 m^3/h, a variation predetermined by the Water Treatment Plant production.

TABLE 1 Results Test series 1 June 4,1970 to September 8, 1970

Date	Temp. °C	Capacity m^3/h	LOVÖ m/h	LAMELLA m/h	Turbidity LOVÖ ZP units	Turbidity LAMELLA ZP units
4.6.	9,0	500	1,7	22,2	84	81
5.6.	9,0	500	1,7	22,2	115	110
8.6.	9,0	378	1,3	16,7	105	88
9.6.	9,0	475	1,6	21,0	99	113
10.6.	10,0	490	1,6	21,8	89	95
11.6.	10,5	560	1,9	25,0	82	87
12.6.	10,0	360	1,2	16,0	78	120
15.6.	10,5	413	1,4	18,4	87	121
16.6.	10,5	450	1,5	20,0	91	116
17.6.	11,0	448	1,5	19,9	97	122
30.6.	13.0	438	1,5	19,5	95	166
1.7.	13,0	351	1,2	15,6	84	97
2.7.	13,0	440	1,5	19,5	83	102
3.7.	14,0	380	1,3	16,9	73	86
6.7.	16,0	347	1,2	15,4	52	63
7.7.	16,0	438	1,5	19,6	69	100
8.7.	16,0	435	1,4	19,4	64	74
9.7.	16,0	349	1,2	15,5	71	74
10.7.	17,0	353	1,2	15,7	79	81
13.7.	17.5	315	1,0	14,0	45	236[1]
14.7.	17,0	325	1,1	14,5	50	85
15.7.	17,0	417	1,4	18,5		206[1]
25.8.	16,5	325	1,1	14,5	90	119
26.8.	16,0	330	1,1	14,7	90	109
27.8.	16,0	574	1,9	25,5	85	113
28.8.	16,5	414	1,4	18,4	79	123
31.8.	18,0	285	1,0	12,7	95	104
1.9.	17,0	400	1,3	17,8	70	120
2.9.	17,0	410	1,4	18,2	86	126
3.9.	16,5	470	1,6	20,9	77	118
4.9.	16,5	360	1,2	16,0	80	115
7.9.	15,0	325	1,1	14,4	94	116
8.9.	15,0	500	1,7	22,2	95	191[1]

[1] = high sludge level

Turbidity measurements were carried out continuously. Samples were also taken occasionally from the different sedimentation basins, Lovö-basin, and the lamella unit.

The sludge was withdrawn from the lamella units at intervals by four separate sludge valves. The sludge level in the sludge pocket was controlled daily by a photometer weight. From June 3, 1970 to July 13, 1970 the interval was 15 minutes, the opening time was 3 minutes, and the flow was about 0.5 litre per second, which corresponds to a loss of water of about 0.1 %. The sludge withdrawal was increased 1970-07-13 to 0.2 % with an interval of 20 minutes and an opening time of 1 minute at a waterflow of 1 litre per second from each sludge pocket. The investigation showed that the sludge withdrawal was too low and the sludge level in the pocket reached the lamella package in two weeks. During the period August 25, 1970 to September 9, 1970 the sludge level was carefully studied. A variation in the sludge level due to the flow through the unit could be observed. An increased sludge depth increased the thickness of the sludge changing the viscosity and the discharge characteristics. This lead to a delayed relationship between the variations in flow and the variations in sludge level. In the case of a marked increase in flow at the same time as a high sludge level, the effluent turbidity was affected. It could also be observed that the lowest turbidity was obtained after a washing of the unit. The water temperature was between 10 and 17°C. The floc formation was almost equal in the conventional system (6 reactors in series) as in the pilot plant (1 reactor). Sedimentation tests were not carried out.

During the second test series from November 2, 1970 to January 1, 1971, the flow of water was higher or 430 - 725 m^3/h. The water temperature was between 3.2 and 7.2°C. In order to avoid disturbances from the sludge pockets the sludge level was kept low, less than 1 m, by an increase in sludge withdrawal corresponding to 1 % loss of water. The result is shown in TABLE 2.

TABLE 2 Results Test series 2 Nov. 2, 1970 to Jan. 7, 1971

Date	Temp. °C	Capacity m³/h	Capacity LOVÖ m/h	Capacity LAMELLA m/h	Turbidity LOVÖ ZP units	Turbidity LAMELLA ZP units
2.11.	6,8	630	2,1	28,0	83	250
3.11.	6,8	650	2,2	28,9	105	260
4.11.		675	2,3	30,0	100	270
5.11.	6,1	560	1,9	24,9	115	190
6.11.	7,0	645	2,2	28,7	70	310
9.11.	5,5	525	1,8	23,4	105	240
9.11.	5,5	660	2,2	29,3	-	420
9.11.	5,5	800	2,7	35,6	-	≈600
10.11.	4,0	700	2,3	31,0	105	360
11.11.	5,5	713	2,4	31,7	130	445
12.11.	4,4	680	2,3	30,2	125	345
13.11.	3,8	680	2,3	30,2	110	365
16.11.	3,5	550	1,8	24,5	120	380
17.11.	3,6	680	2,3	30,2	140	300
18.11.	3,4	680	2,3	30,2	105	300
19.11.	3,4	725	2,4	32,2	120	340
20.11.	3,5	700	2,3	31,0	105	330
23.11.	3,2	510	1,7	22,7	110	206
24.11.	3,8	560	1,9	24,9	105	330
25.11.	3,8	560	1,9	24,9	120	295
26.11.	4,0	640	2,1	28,4	115	340
27.11.	5,0	640	2,1	28,4	105	320
28.11.	-	475	1,6	21,1	-	220
29.11.	-	475	1,6	21,1	-	220
30.11.	-	560	1,9	24,9	-	310
1.12.	4,5	560	1,9	24,9	90	320
2.12.	5,0	560	1,9	24,9	120	300
2.12.	5,0	640	2,1	28,4	-	500
3.12.	5,0	560	1,9	24,9	-	350
4.12.	4,5	560	1,9	24,9	135	370
5.12.	-	430	1,4	19,1	-	260
6.12.	4,2	430	1,4	19,1	-	260
7.12.	4,2	500	1,7	22,2	135	340
8.12.	4,0	580	1,9	25,8	135	380
9.12.	4,5	500	1,7	22,2	145	350
10.12.	4,5	500	1,7	22,2	130	380
11.12.	4,5	575	1,9	25,5	155	350
14.12.	4,5	575	1,9	25,5	135	390
15.12.	5,5	570	1,9	25,3	100	340
16.12.	4,5	600	2,0	26,6	175	520
17.12.	4,0	570	1,9	25,3	150	400
21.12.	5,0	600	2,0	26,6	150	445
22.12.	3,5	635	2,1	28,2	150	600
7.1.	-	575	1,9	25,5	90	360

During the tests at this low water temperature great differences in floc formation in the conventional system and in the pilot plant were observed. In order to explain the differences sedimentation analyses were carried out daily during the period November 2, 1970 to November 26, 1970. From this investigation it was obvious that the floc formation was significantly less effective in the pilot plant reactor compared to the conventional system. The settling velocity was almost twice as high in the six reactor system as in the one reactor system.

In order to further improve the sludge withdrawal the outlet valves were changed to bigger valves. To get a better adaption to the maintenance of the Water Treatment Plant a new outlet channel from the lamella unit was built from January to February 1971. This channel was connected to the filtration unit reserved for the pilot plant system.

The plant was tested from March 16, 1971 to April 8, 1971 and the result from this period is shown in TABLE 3.

TABLE 3 Results March 16, 1971 to April 6, 1971

Date	Temperature ^{o}C	Capacity lamellæ m^3/h	Capacity lamellæ m/h	Turbidity ZP units
16.3	1,5	270	12	310
17.3	1,4	270	12	290
19.3	1,6	270	12	275
22.3	1,4	323	15	350
24.3	1,4	388	18	600
26.3	1,4	235	11	255
29.3	1,5	300	14	340
31.3	1,7	310	14	275
1.4	1,7	270	12	290
2.4	1,9	270	12	260
5.4	2,0	300	14	320
6.4	2,0	260	12	290

The plant was further completed with a weir for effluent water during April 8-26, 1971 and the plant was in operation from April 4, 1971 to June 4, 1971. The flow through the plant was decreased by 50 % during this period and was used in normal operations at the Water Treatment Plant.

The floc formation was, especially from March 16, 1971 to April 8, 1971, very poor at low temperatures.

The sludge was emptied in 5 seconds and at a flow of 15 litres per second every 40 minutes which corresponds to a loss of water of about 0.2 %. In spite of larger sludge valves it could be observed that the sludge discharge decreased when the depth of sludge in the sludge pocket increased. The sludge discharge interval was reduced to 20 minutes with equal discharge time. The increased sludge withdrawal was not sufficient because it was necessary to have a longer discharge time than 5 seconds in order to reach a maximum discharge capacity. The discharge time was thus increased to 20 seconds and the sludge was withdrawn every hour.

The dry solids concentration in the sludge and the sludge depth as a function of the sludge discharge is shown in TABLE 4. This

TABLE 4 Sludge studies, April 29, 1971 to June 4, 1971

Date	Dry solids %	Loss of water %	Sludge depth m	Discharge interval min.	Discharge duration sec.
29.4.	1,05	0,2	0,9	40	5
5.5.	0,83		1,80	40	
6.5.	0,70		1,65	40	
13.5.	0,60		2,40	35	5
14.5.	0,63		2,50	35	
17.5.	0,60		2,10	35	
18.5.	0,62		2,50	35	
19.5.	0,74	≈0,4	2,40	20	5
21.5.	0,77		1,90	20	
24.5.	0,38		1,90	60	20
25.5.	0,36		1,70	60	
26.5.	0,32		1,80	60	
27.5.	0,29	0,5	1,70	60	
28.5.	-		1,70	60	
1.6.	-		-		
2.6.	-		-		
4.6.	0,39		1,60	60	

investigation was carried out during a period of moderate variations in the flow. One can see from the table that a short discharge time implied a high dry solids content in the sludge due to a high sludge depth. Due to the high sludge viscosity, an increase in the discharge interval did not affect the sludge withdrawal very much. At a longer discharge time, however, the sludge level could be kept 1 meter below the lamella package. The dry solids content in the sludge decreased from 7 grams per litre to 3.5 grams per litre. The sludge withdrawal chosen corresponds to a water loss of about 0.5 %, a value which may be regarded as the maximum sludge withdrawal necessary.

The residual turbidity was, especially at the low water temperature, higher (200 Zeiss Pulfrisch units) in the pilot plant than in the Lovö-basin (120 Zeiss Pulfrisch units) due to the flocculation system. In order to improve the flocculation operation the flocculation unit was redesigned similar to the conventional system in the Water Treatment Plant. This was done in 1972. The results from the tests during 1972-1973 are shown in TABLE 5.

TABLE 5 Results - Test series October 23, 1972 to May 1973

PLANT			LAMELLA			LOVÖ		
Date	Temp. °C	Flow m³/h	Flow m³/h	Surface rate m/h	Turb. Std 10 %	Flow m³/h	Surface rate m/h	Turb. %
1972								
23/10	6,0	5110	540	24,0	53	870	2,65	-
25/10	6,0	5400	450	20,0	16	715	2,25	-
26/10	6,0	5360	445	19,8	18	710	2,25	-
27/10	6,0	5600	465	20,6	20	745	2,35	17
27/10	6,0	5600	595	26,1	40	-	-	-
28/10	6,0	4060	336	15,0	20	-	-	-
29/10	6,0	4060	336	15,0	19	-	-	-
30/10	5,6	5490	576	25,7	32	-	-	-
31/10	5,6	5500	455	20,2	21	730	2,30	19
1/11	6,3	5510	457	20,3	20	730	2,30	18
1/11	6,3	5510	583	25,8	27	935	3,00	18
2/11	6,7	5550	461	20,5	21	735	2,30	19
3/11	6,3	5540	460	20,5	20	735	2,30	18
3/11	6,3	3560	296	13,0	13	-	-	-
3/11	6,3	3760	313	13,9	19	-	-	-
4/11	6,3	-	-	-	-	-	-	-
5/11	-	-	-	-	-	-	-	-
6/11	-	-	-	-	-	-	-	-
7/11	5,6	5570	478	21,2	30	725	2,30	24
8/11	5,6	5550	477	21,2	30	720	2,25	23
9/11	6,3	5570	607	27,0	36	930	3,00	20
10/11	6,3	5560	478	21,2	29	-	-	-
11/11	6,1	3920	337	15,0	18	-	-	-
12/11	6,0	4010	345	13,3	18	-	-	-
13/11	5,3	5570	610	27,0	42	950	3,00	26
14/11	5,6	5570	480	21,3	32	-	-	-
15/11	5,6	5590	490	21,8	28	735	2,30	24
16/11	5,3	5580	480	21,4	28	735	2,30	24
17/11	5,3	5630	485	21,5	29	742	2,35	20
18/11	5,3	4680	403	17,9	22	618	1,95	18
19/11	4,6	4680	403	17,9	22	618	1,95	18
20/11	3,9	5780	620	28,0	70	960	3,05	24
20/11	3,9	5960	649	29,0	80	785	2,50	-
21/11	3,9	5550	477	21,2	34	730	2,30	16
21/11	3,9	5550	605	26,8	75	945	3,00	17
22/11	4,2	5620	482	21,4	35	730	2,30	14
23/11	4,2	5640	485	21,5	33	735	2,30	20
24/11	4,2	5640	402	17,8	23	735	2,30	20
25/11	4,0	5650	403	17,9	22	735	2,30	21
26/11	4,0	5650	402	17,8	22	735	2,30	21
27/11	4,0	5590	509	22,6	28	860	2,70	23
27/11	4,0	5600	397	17,6	22	745	2,35	22
28/11	4,0	6050	429	19,0	25	805	2,55	16
29/11	3,9	6200	440	19,5	28	825	2,60	16

PLANT			LAMELLA			LOVÖ		
Date	Temp. °C	Flow m³/h	Flow m³/h	Surface rate m/h	Turb. Std 10 %	Flow m³/h	Surface rate m/h	Turb. %
30/11	3,9	5570	395	17,5	24	740	2,35	21
1/12	3,9	4490	318	14,1	23	595	1,90	21
2/12	3,9	4490	318	14,1	23	595	1,90	21
3/12	3,9	4490	318	14,1	23	595	1,90	23
4/12	3,9	5660	400	17,8	25	752	2,40	23
5/12	4,2	5630	395	17,7	25	750	2,35	23
6/12	3,9	5560	402	17,8	23	740	2,35	22
7/12	4,2	5600	400	17,8	24	745	2,35	21
8/12	4,2	4500	320	14,3	19	600	1,90	18
9/12	4,2	4500	320	14,3	19	600	1,90	18
10/12	4,2	4500	320	14,3	19	600	1,90	18
11/12	4,2	4500	320	14,3	19	600	1,90	18
11/12	4,2	5650	505	22,2	45	860	2,70	21
12/12		4500	320	14,3	19	600	1,90	20
13/12		-	-	-	-	-	-	-
14/12		5500	395	17,5	21	730	2,30	17
15/12		5500	395	17,5	22	730	2,30	18
16/12		4500	320	14,3	20	600	1,90	19
1973								
8/3	2,4	5580	375	16,7	20	730	2,30	20
9/3	2,5	5580	375	16,7	17	730	2,30	20
10/3	2,6	4480	300	13,4	13	595	1,90	18
11/3	2,6	4490	300	13,4	13	595	1,90	18
12/3	2,6	4490	300	13,4	14	595	1,90	18
13/3	2,6	5580	485	21,6	21	970	3,10	21
13/3	2,6	5580	375	16,7	16	745	2,35	19
14/3	3,0	5580	375	16,7	17	745	2,35	19
15/3	3,0	5580	375	16,7	16	745	2,35	19
16/3	3,3	5580	375	16,7	16	745	2,35	19
17/3	-	-	-	-	-	-	-	-
18/3	-	-	-	-	-	-	-	-
19/3	3,6	5600	375	16,7	17	745	2,35	20
19/3	3,7	5600	487	21,6	35	985	3,10	25
20/3	3,7	5600	375	16,7	17	745	2,35	21
20/3	3,7	5600	487	21,6	30	985	3,10	25
21/3	4,0	5550	375	16,7	17	745	2,35	20
22/3	4,1	5500	370	16,4	16	740	2,35	19
22/3	4,3	5500	370	16,4	15	740	2,35	18
24/3	4,6	4500	300	13,3	13	600	1,90	17
25/3	4,8	4500	300	13,3	13	600	1,90	17
26/3	5,0	5600	487	21,6	22	975	3,10	19

PLANT			LAMELLA			LOVÖ		
Date	Temp. °C	Flow m³/h	Flow m³/h	Surface rate m/h	Turb. Std 10 %	Flow m³/h	Surface rate m/h	Turb. %
1973								
6/4	5,5	4270	285	12,6	8	567	1,80	13
7/4	5,6	4470	300	13,3	9	600	1,90	12
8/4	5,6	4410	295	13,2	9	550	1,90	14
9/4	5,7	5600	485	21,5	17	965	3,00	15
9/4	5,7	5600	975	16,7	14	745	2,35	14
10/4	5,7	5580	370	16,6	14	740	2,35	14
11/4	5,8	5570	487	21,6	25	965	3,00	17
12/4	5,8	5600	375	16,6	14	745	2,35	14
13/4	5,7	5600	375	16,6	13	745	2,35	13
14/4	5,7	4150	275	12,2	8	550	1,75	13
15/4	5,8	4170	280	12,4	9	555	1,80	13
16/4	5,7	5570	485	21,5	20	960	3,05	17
16/4	5,7	5620	376	16,7	12	745	2,35	13
17/4	6,0	5600	485	21,5	20	960	3,05	17
17/4	6,0	5600	375	16,6	12	745	2,35	13
18/4	6,2	5600	375	16,6	9	745	2,35	16
-								
-								
-								
-								
-								
24/4	6,6	5540	370	16,5	10	740	2,30	13
25/4	6,8	5540	370	16,5	10	740	2,30	13
26/4	6,7	5600	375	16,6	11	745	2,35	14
27/4	7,2	5570	375	16,6	11	745	2,35	11
2740	7,2	3750	250	11,2	8	500	1,60	14
28/4	7,2	3750	250	11,2	10	500	1,60	14
29/4	7,1	3750	250	11,2	11	500	1,60	14
30/4	7,5	3800	260	11,3	9	510	1,60	12
1/5	7,7	3800	260	11,3	9	510	1,60	13
2/5	7,9	5600	375	16,6	10	745	2,35	15
3/5	8,4	5600	375	16,6	10	745	2,35	13
4/5	8,2	5620	376	16,7	11	745	2,35	12
4/5	8,2	4500	300	13,4	9	600	1,90	14
5/5	8,6	4150	277	12,3	12	550	1,75	17
6/5	9,0	4150	277	12,3	12	550	1,75	17
7/5	9,4	4150	277	12,3	11	550	1,75	17
8/5	9,4	5600	375	16,6	11	745	2,35	14
9/5	9,5	5600	485	21,5	22	960	3,05	15

APPENDIX 7
ORGINAL DATA
Filtration

Filtration studies have mainly been performed in pilot plants (Plant 1, 3 and 5). Original data from these investigations are to be found in the following reports.

Hedberg, T.: Optimering av den kemiska reningsprocessen samt hårdhetshöjning av mjuka ytvatten.

Optimization of the water treatment process and increase of the water hardness.
Publikationsserie B69:1

Hedberg, T.: Kemisk rening av vatten från Bolmen. Rapport från försöksverksamhet 1971-1973.

Chemical treatment at water from the lake Bolmen, Report from investigations 1971-1973.
Sydvatten, 1974.

Apart from this filtration has been studied in the following Master of Science Engineering examination reports.

Examensarbete 72:1

Undersökning av flockning, sedimentering och filtrering i full skala vid Lackarebäcksverket. — Examination of flocculation, sedimentation and filtration in full scale at Lackarebäck Water Treatment Plant.

Examensarbete 72:2

Optimeringsaspekter på den kemiska reningsprocessen för försörjningsvatten. — Optimization aspects on the chemical treatment process.

Examensarbete 75:10

Utveckling av filter media. — Development of filter media.

Some tests have also been carried out in pilot plant 3 in order to complete the information on the filtration operation. Some tests have also been performed in full scale at the Lackarebäck Water Treatment Plant.

The results from these investigations are shown in the following:

PLANT 5

To complete the filtration studies reported earlier different filter designs have been studied under various conditions during the period June 18, 1974 to January 30, 1976.

Filter designs, Media characteristics

Filter No. (see FIG.7-1)	D_{10} mm	D_{60} mm	D_{60}/D_{10} mm	Remark
1 Sand	0,85	1,1	1,3	
2 Sand	0,85	1,1	1,3	PLANT 4
3 Sand	0,85	1,1	1,3	PLANT 4
4 Sand	0,65	0,95	1,5	
5 Sand Anthracite	0,85 1,70	1,1 2,20	1,3 1,3	
6 Sand Anthracite	0,65 0,95	0,95 1,15	1,5 1,2	
7 Act. carbon	1,0	1,65	1,7	
8 Sand Act. carbon	0,65 0,67	0,95 1,1	1,3 1,6	
9 Garnet I Garnet II Sand Anthracite	0,7 0,25 0,65 1,60	1,25 0,43 0,95 2,00	1,8 1,7 1,5 1,3	
10 Sand	0,85	1,1	1,3	

Results are shown in the following TABLE 1

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of 0,5 (m water) % Std 2	1,0	1,5	2,0	Head loss at a turbidity of 15 % Std 2 Sigrist m water
1.	1	740618	14,5	6	35	100	6	6	6	6	> 2,5
		0626	15,2	6	65	100	6	5	5	5	> 2,2
		0703	17,5	0	30	13	6	7,5	12	-	1,3
		0723	17,5	0	50	19	10	10	11	-	1,3
		0820	17,4	4	30	58	8	8	8		> 1,8
		750414	3,9	0	90	29	11	20	40	55	0,7
		0421	5,3	0	55	15	12	22	-	-	0,8
		0428	7,3	6	110	77	9	10	11	-	1,8
		0505	8,7	6	30	71	8	8	8	-	> 1,8
1.	2	0611	14,0	0	90	17	-	-	-	-	-
		0612	14,0	4	110	58	-	-	-	-	-
		0613	14,0	6	110	110	-	-	-	-	-
1.	3	0616	15,6	0	115	17	-	-	-	-	0,5
		0617	15,6	0	64	14	-	-	-	-	-
		0618	15,6	0	195	23	-	-	-	-	0,4
1.	4	0625	16,1	0	60	16,1	8	15	-	-	0,9
		0630	17,3	0	120	23,5	15	-	-	-	0,5
		0702	17,8	2	115	33,3	9	9	26	-	1,2
		0707	18,2	2	60	24	9	9	9	9	> 2,0
		0714	19,6	2	35	20,4	9	9	-	-	> 1,3
		0812	20,0	0	55	19,4	9	8	11	-	≃ 1,7
		0818	20,5	0	115	31,5	10	15	-	-	1,0
		0819	20,3	0	110	26,6	9	12	-	-	1,1
		760113	1,7	6	55	79	9	8	8	8	> 1,9
		0114	1,7	6	55	83	9	8	8	7,5	> 2,2
		0115	1,7	6	55	80	10	9	8	7	> 2,4
		0116	1,7	6	34	75	8	7	-	-	-
		0120	1,7	6	45	80	10	10	10	10	> 2,2
		0121	1,7	6	110	180	23	20	-	9	-
		0122	1,7	2	100	55	24	25	11	11	0

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of (m water) % Std 2				Head loss at a turbidity of 15 % Std 2 Sigrist m water
							0,5	1,0	1,5	2,0	
4.	1	740618	14,5	6	35	100	5	5	5	5	> 2,5
		0628	15,5	6	60	125	5	5	5	5	> 2,5
		0703	17,5	0	30	24	5	6	9	10	≃ 2,2
		0723	17,5	0	55	27	10	12	17	16	1,2
		0820	17,4	4	30	66	8	8	8	-	> 1,8
		750414	3,9	0	90	25	10	42	60	-	0,4
		0421	5,3	0	55	16	15	-	-	-	0,5
		0428	7,3	6	115	17	9	17	-	-	0,9
		0505	8,7	6	30	63	8	7	7	-	> 1,6
4.	2	0611	14,0	0	90	40	-	-	-	-	-
		0612	14,0	4	110	64	-	-	-	-	-
		0613	14,0	6	110	96	-	-	-	-	-
4.	3	0616	15,6	0	115	21	-	-	-	-	-
		0617	15,6	0	64	16	-	-	-	-	-
		0618	15,6	0	195	24	-	-	-	-	-
4.	4	0625	16,0	0	60	19,4	8	23	-	-	0,8
		0630	17,3	0	120	26,3	22	-	-	-	0,3
		0702	17,8	2	115	37,5	14	36	-	-	0,6
		0707	18,2	2	60	32,3	9	10	15	21	1,4
		0714	19,6	2	35	25	8	-	-	-	-

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of (m water) % Std 2: 0,5	1,0	1,5	2,0	Head loss at a turbidity of 15 % Std 2 Sigrist m water
5.	1	740618	14,5	6	35	18	5	5	5	-	> 1,5
		0625	15,2	6	65	27	5	5	5	10	1,6
		0703	17,5	0	30	6	7	-	-	-	> 0,6
		0723	17,5	0	50	10	10	15	-	-	0,9
		0820	17,4	4	30	13	8	8	8	-	≃ 1,6
		750414	3,9	0	90	15	15	50	-	-	0,5
		0421	5,3	0	55	6	18	-	-	-	0,4
		0428	7,3	6	115	33	9	9	-	-	0,1
		0505	8,7	6	30	14	8	7	-	-	> 1,0
5.	2	0611	14,0	9	90	8	-	-	-	-	-
		0612	14,0	4	110	19	-	-	-	-	-
		0613	14,0	6	110	23	-	-	-	-	-
5.	3	0616	15,6	0	115	7,5	-	-	-	-	0,3
		0617	15,6	0	64	5,7	-	-	-	-	-
		0618	15,6	0	195	12	-	-	-	-	0,3
5.	4	0625	16,1	0	60	6,8	15	-	-	-	0,5
		0630	17,3	0	120	10,7	40	-	-	-	0,4
		0702	17,8	2	115	13,6	9	-	-	-	0,55
		0707	18,2	2	50	7,1	9	20	-	-	0,8
		0714	19,6	2	35	7,1	8	-	-	-	-
		0812	20,0	0	55	6,8	7	15	-	-	0,7
		0818	20,5	0	115	13,6	14	-	-	-	0,5
		0819	20,3	0	110	12,5	12	-	-	-	0,55
		760113	1,7	6	55	20	9,5	-	8	-	> 1,6
		0114	1,7	6	55	20	9,5	7,5	6,5	-	> 1,6
		0120	1,7	6	45	40	10	13	-	-	-
		0121	1,7	6	60	50	11	10	-	-	> 1,0
		0121	1,7	6	110	50	28	22	9,5	-	0
		0122	1,7	2	100	30	30	9	-	-	-

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of (m water) % Std 2: 0,5	1,0	1,5	2,0	Head loss at a turbidity of 15 % Std 2 Sigrist m water
6.	1	740618	14,5	6	35	37	5	5	6	5	> 2,2
		0625	15,2	6	65	50	5	5	5	5	> 2,6
		0703	17,5	0	35	9	5	5	-	-	> 1,0
		0723	17,5	0	50	12	10	10	-	-	> 1,2
		0820	17,4	4	30	26	8	8	8	8	> 2,1
		750414	3,9	0	90	17	9	25	-	-	0,9
		0421	5,3	0	55	9	9	-	-	-	0,8
		0428	7,3	6	110	46	9	8	8	-	1,7
		0505	8,7	6	30	26	8	7	7	-	> 1,6
6.	2	0611	14,0	0	90	10	-	-	-	-	-
		0612	14,0	4	110	35	-	-	-	-	-
		0613	14,0	6	110	38	-	-	-	-	-
6.	3	0616	15,6	0	115	11,5	-	-	-	-	> 0,5
		0617	15,6	0	64	8,5	-	-	-	-	-
		0618	15,6	0	195	20,5	-	-	-	-	≃ 0,5
6.	4	0625	16,1	0	60	8	8	-	-	-	> 0,7
		0630	17,3	0	120	15,7	8	50	-	-	0,75
		0702	17,8	2	115	18,2	8	10	-	-	> 1,3
		0707	18,2	2	50	10	8	8	-	-	> 1,2
		0714	19,6	2	35	8,8	8	-	-	-	> 0,7
		760113	1,7	6	55	35	8,5	8	-	-	> 1,0
		0114	1,7	6	55	30	9	8	-	6,5	> 2,0
		0120	1,7	6	45	20	10	10	-	-	> 1,0
		0121	1,7	6	100	70	24	-	9	-	-
		0122	1,7	2	110	50	24	9	-	-	> 1,0

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of (m water) % Std 2				Head loss at a turbidity of 15 % Std 2 Sigrist m water
							0,5	1,0	1,5	2,0	
7.	1	740618	14,5	6	35	63	-	5	5	5	> 2,5
		0625	15,2	6	65	67	5	5	5	5	> 2,5
		0703	17,5	0	30	10	5	10	-	-	> 0,9
		0723	17,5	0	50	13	11	17			0,6
		0820	17,4	4	30	37	8	8	8	8	> 1,8
		750414	3,9	0	90	15	12	-	-	-	0,3
		0421	5,3	0	55	8	14	-	-	-	0,5
		0428	7,3	6	115	45	10	20	-	-	0,8
		0505	8,7	6	30	34	8	7	7	-	> 1,8
7.	2	0611	14,0	0	90	10	-	-	-	-	-
		0612	14,0	4	110	37	-	-	-	-	-
		0613	14,0	6	110	52	-	-	-	-	-
7.	3	0616	15,6	0	115	13	-	-	-	-	-
		0617	15,6	0	64	10	-	-	-	-	-
		0618	15,6	0	195	14	-	-	-	-	-
7.	3	0625	16,1	0	60	11,1	10	-	-	-	0,6
		0630	17,3	0	120	15	32	-	-	-	0,4
		0702	17,8	2	115	20,8	10	35	-	-	0,6
		0707	18,2	2	50	14,8	9	15	-	-	1,0
		0714	19,6	2	35	11,4	8	-	-	-	> 0,8
		760113	1,7	6	55	45	8	7,5	-	-	> 1,0
		0114	1,7	6	55	40	8	7,5	-	7	> 1,8
		0121	1,7	6	110	95	18	-	-	9	> 2,0
		0122	1,7	2	110	45	24	10	-	-	-

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of 0,5 (m water) % Std 2	1,0	1,5	2,0	Head loss at a turbidity of 15 % Std 2 Sigrist m water
8.	1	740618	14,5	6	35	105	5	5	5	5	> 2,3
		0626	15,2	6	60	125	5	5	5	5	-
		0703	17,5	0	35	18	5	5	5	-	> 1,6
		0723	17,5	0	50	28	10	10	10	10	> 2,4
		0820	17,4	4	30	63	8	8	8	-	> 1,8
		750414	3,9	0	90	29	9	15	35	-	1,0
		0428	5,3	6	110	71	8	8	8	-	≃ 2,0
		0505	8,7	6	30	67	8	8	7		> 1,7
8.	2	0611	14,0	0	90	18	-	-	-	-	-
		0612	14,0	4	110	51	-	-	-	-	-
		0613	14,0	6	110	90	-	-	-	-	-
		0616	15,6	0	115	22	-	-	-	-	-
		0617	15,6	0	64	16	-	-	-	-	-
		0618	15,6	0	195	26	-	-	-	-	-
8.	4	0625	16,1	0	60	17,6	8	8	-	-	> 1,0
		0630	17,3	0	120	24,0	8	34	-	-	0,8
		0702	17,8	2	115	37,5	8	8	8	-	> 1,8
		0707	18,2	2	50	28,5	8	8	9	8	> 2,0
		0714	19,6	2	35	23,5	9	8	8	8	> 2,0
		760113	1,7	6	55	75	8	-	7	-	> 1,5
		0114	1,7	6	55	77	8	-	7	6,5	> 2,0
		0120	1,7	6	45	60	9	9	9	9	> 2,0
		0121	1,7	6	110	150	20	14	-	8	> 2,0
		0122	1,7	2	110	80	21	-	9	-	> 1,5

TABLE 1 Result Filtration PLANT 3

Test serie 1 Stirring speed in the flocculation tanks 10, 5, 2, 1,5 rpm
" 2 " - 20, 5, 2, 1,5 "
" 3 " - 20, 5, 2, 1,5 "
" 4 " - 10, 5, 2, 1,5 "

Filter type	Test serie No.	Date	Temp °C	Act. silica ppm	Turbidity in sed. % Std 2	$\frac{\Delta H_f}{\Delta T_f}$ mm/h	Turbidity at a head loss of 0,5 (m water) % Std 2	1,0	1,5	2,0	Head loss at a turbidity of 15 % Std 2 Sigrist m water
9.	1	740618	14,5	6	35	77	5	5	5	5	> 2,6
		0625	15,2	6	65	77	5	5	5	5	> 2,2
		0703	17,5	0	30	22	5	5	5	5	> 2,0
		0723	17,5	0	50	42	10	10	10	10	> 2,2
		0820	17,4	4	30	48	8	8	8	-	> 1,8
		0414	3,9	0	90	21	10	10	30	-	1,2
		0421	5,3	0	55	11	10	15	-	-	1,0
		0428	7,3	6	110	56	8	8	8	-	≃ 2,0
		0505	8,7	6	30	21	8	8	8	-	> 1,6
		750611	14,0	0	90	9	-	-	-	-	-
		0612	14,0	4	110	30	-	-	-	-	-
		0613	14,0	6	110	36	-	-	-	-	-
9.	3	750616	15,6	0	115	14	-	-	-	-	-
		0617	15,6	0	64	8,5	-	-	-	-	-
		0618	15,6	0	195		-	-	-	-	-
9.	4	0625	16,1	0	60	10,8	8	-	-	-	> 0,8
		0630	17,3	0	120	17,6	9	10	-	-	1,2
		0702	17,8	2	115	21,0	8	8	7	-	> 1,6
		0707	18,2	2	50	14,2	8	8	8	-	> 1,5
		0714	19,6	2	35	9,1	8	8	-	-	> 1,0
		760113	1,7	6	55	40	-	7,5	-	-	> 1,2
		0114	1,7	6	55	40	-	7,5	6,5	-	> 1,5
		0120	1,7	6	45	30	-	9	10	-	> 1,4
		0121	1,7	6	110	105	-	15	-	-	> 2,0
		0122	1,7	2	110	40	-	23	9	-	> 1,2

PLANT 7

During a period October 12, 1975 to October 22, 1975 an investigation was performed in full scale at the Lackarebäck Water Treatment Plant to test the plant performance. The filtration rate through the active carbon filters was increased from about 5.5 m/h to 8.0 m/h.

TABLE 2 Results from the plant performance test. Mean values from 14 filters

Date	Temp. ^oC	Activated silica ppm	Turbidity c_o % Std 5	Head loss increase mm/h	Filtration rate m/h
751012	10	2	28	17	5,6
751013	10	2	36	29	5,6
751014	10	2	40	29	5,6
751015	10	2	40	27	6,1
751016	10	2	43	32	6,7
751020	10	2	40	34	6,7
751021	10	2	40	38	7,3
751022	10	2	40	42	8,0

APPENDIX 8

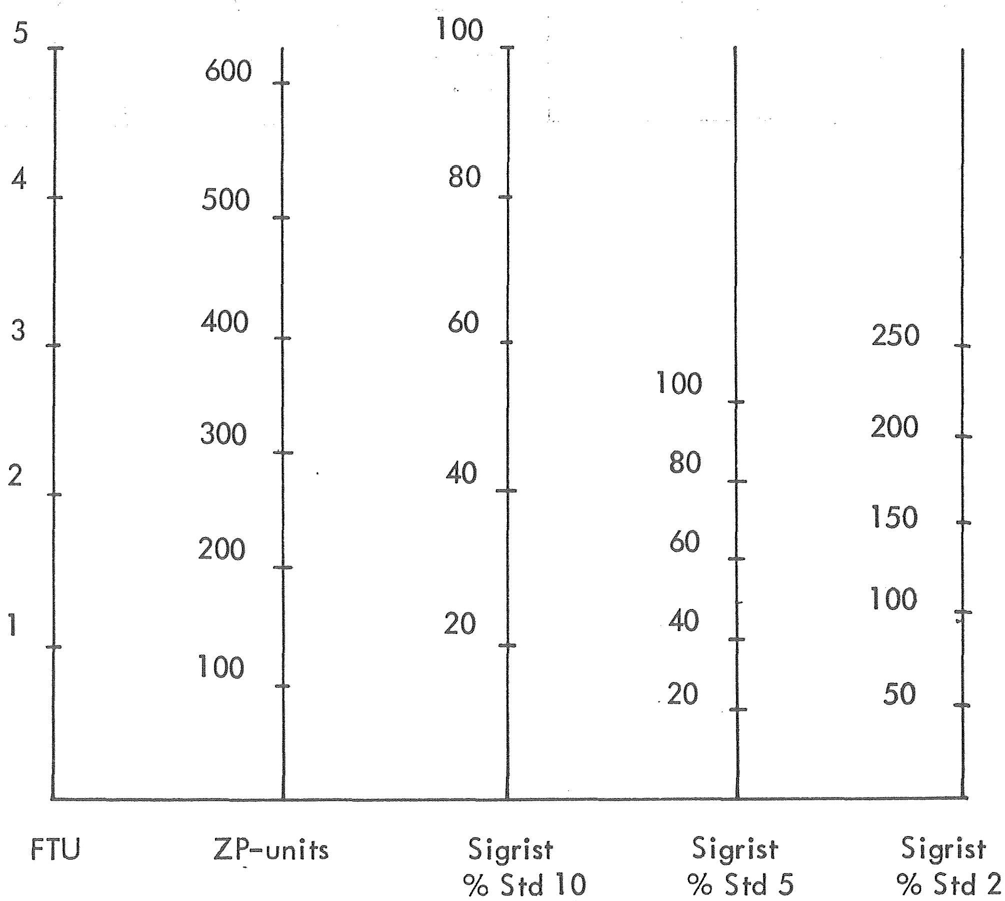
TURBIDITY MEASUREMENTS

In the investigation three different turbidity instruments have been used

Zeiss-Pulfrich instrument (Zeiss Pulfrich-unit, ZP units)
Sigrist photometer ($mg/l\ Si\ O_2$)
Hach-instrument (FTU-units)

The different instruments have been calibrated against each other and a conversion table for the turbidity units has been obtained. In the Sigrist photometer different turbidity standards can be used. Each standard is calibrated in $mg/l\ Si\ O_2$ (for example 100 % standard 1 = 1 $mg/l\ Si\ O_2$)

Conversion table for turbidity units



APPENDIX 9
COEFFICIENTS AND COSTS

COEFFICIENTS

Flocculation

Equation (3-3)

C_o rel %
T sec
G sec^{-1}
K_1 and K_2, see FIG. 3-14
$P = 4/3$
$K_{R\,10} = 7.5$ rel %
$a = 8$, $b = 1$, $d = 1$

Equation (3-5)

C_o rel %
v_f m/h
$a = 20$, $b = 0.6$

Equation (3-9)

C_o rel %
$K_g = 154$, $l_g = -90$

Sedimentation

Equation (5-4)

v_f m/h
$K_T = 0.9$
$K_A = 0.7 - 1.0$
$\gamma_o = 0.005$
$\gamma = 0.0013$
$\cos\alpha = 0.57$

Equation (5-7)

c_o rel %
$f(S) = \frac{8-S}{5}$

Filtration

Equation (7-9)

C % Std 2 Sigrist

C_o % Std 2 Sigrist

$a = 15$, $b = 0.08$, $d = 0.02$

$f_v = 1.0$

Equation (7-11)

$H_{f\ max}$ mm of water

A mg/l

C_o % Std 2 Sigrist

t ^oC

Filter type	1	4	5	6	7	8	9
δ_1	150	150	160	180	275	150	160
δ_2	40	40	25	50	30	100	70
δ_3	180	180	160	180	140	180	180
δ_4	120	120	120	120	120	120	120
δ_5	20	20	40	40	40	40	40
δ_6	200	100	50	200	100	200	200

Equation (7-22)

H mm of water

t ^oC

T_f h

A mg/l

C_o % Std. 2, Sigrist

v m/h

Filter type	1	4	5	6	7	8	9
K_{H_o}	345	345	245	335	230	280	420
a	0.35	0.40	0.13	0.20	0.30	0.37	0.13
b	0.25	0.25	0.23	0.20	0.20	0.66	0.23
d	0.033	0.033	0.033	0.033	0.033	0.033	0.033
f	1	1	1	1	1	1	1
g	1.0	1.0	0	0.45	0.5	0.8	0
h	0.005	0.005	0	0	0.005	0.03	0
p	0	0	0.22	0	0	0	0.22
j	0.15	0.15	0.053	0.10	0.08	0.06	0.05
α	2	2	2	2	2	2	2
ρ	2	2	2	2	2	2	2
γ	1.25	1.25	1.60	1.25	1.30	1.90	1.60
δ	0	0	0.2	0	0	0.2	0.2

COSTS

Capital costs

Interest 6 % 25 years

Chemicals

Coagulant	500 Sw.cr./ton
Activated silica	425 - " -
Lime	400 - " -

Tax and operation costs are included

Flocculation

Building costs

Concrete	800 Sw.cr./m^3
Walls	300 Sw.cr./m^2
Roof	350 - " -
Foundation	150 - " -
Stirring equipment	150 Sw.cr./m^3 reactor volume
Baffles	100 Sw.cr./m^2

Operational costs

Ventilation, heating	10 Sw.cr./m^3	building volume and year
Maintenance	9 Sw.cr./m^3	water per hour and year

Sedimentation

Building costs

See flocculation

Lamella sedimentation unit:

Side walls	1300 Sw.cr./m^2
Plates	100 - " -

Operational costs

Ventilation, heating	10 Sw.cr./m^3	building volume and year
Maintenance		
Lamella sed.	4 Sw.cr./m^3	water per hour and year
Lovö basin	10 - " -	- " -

Filtration

Building costs

Filter media in Sw.cr./m^3

Filter type	1	2	3	4	5	6
	250	550	550	3000	2500	600

Filter bottom	500 Sw.cr/m^2	
Regulator	30 000 Sw.cr./filter	
Pumps	3 Sw.cr./m^3	water per hour and meter
Other	50 Sw.cr./m^2	

Operational costs

Ventilation, heating	10 Sw.cr./m^3	building volume and year
Pump	3 Sw.cr./m^3	water per hour, meter, and year
Maintenance	4 - " -	water per hour and year