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PRELIMINARY DESIGN OF A WATER TREATMENT PLANT

by

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ABSTRACT

A preliminary design of a water treatment plant with a maximum design flow of 6000 gallons per minute is developed. The cost associated with turbidity removal is minimized by finding the optimal balance between the amount of turbidity removed by sedimentation and the amount removed by filtration.

A cost function for sedimentation is derived showing the relationship between the cost of sedimentation and the sedimentation basin effluent turbidity. A cost function for filtration is similarly derived showing the relationship between the cost of filtration and the filter influent turbidity. The total turbidity removal cost is the sum of the sedimentation cost and the filtration cost. The use of spreadsheets to calculate the total turbidity removal cost is described.

A clear optimal sedimentation basin effluent turbidity is found and is used to form the basis for the preliminary water treatment plant design.

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INTRODUCTION

In a conventional water treatment plant the sequential order of treatment processes is coagulation, flocculation, sedimentation, and filtration. In coagulation, colloidal particles are destabilized so they can begin the initial aggregation process. Flocculation is a transport process that causes the collision of destabilized colloids to further aggregate. Sedimentation is a solid-liquid separation process where some of the aggregated solids are allowed to settle out. In filtration the filter media traps most of the particles that were not removed by sedimentation.

This thesis is a preliminary design of a water treatment plant, which will form the basis for the final design of the North Table Mountain Water Treatment Plant (NTMWTP). The coagulation, flocculation, and sedimentation processes will be designed using generally acceptable engineering design parameters. The sedimentation process will also be theoretically designed and compared to the practical design to see if any correlation exists. The equations that govern the settling velocity of particles in a liquid will be used to theoretically size a sedimentation basin. The filtration process will not be designed as an adequate filtration system already exists at the NTMWTP.

The cost associated with turbidity removal will be minimized by finding the optimal balance between the amount of turbidity removed by sedimentation and the amount removed by filtration. The cost of sedimentation is a function of the sedimentation basin effluent turbidity where the larger, or more costly, the basin the lower the effluent turbidity. The cost of filtration is a function of the filter influent turbidity where the higher the influent turbidity the higher the cost of filtration. Therefore, an optimal size sedimentation basin exists such that the total cost of turbidity removal is minimized.

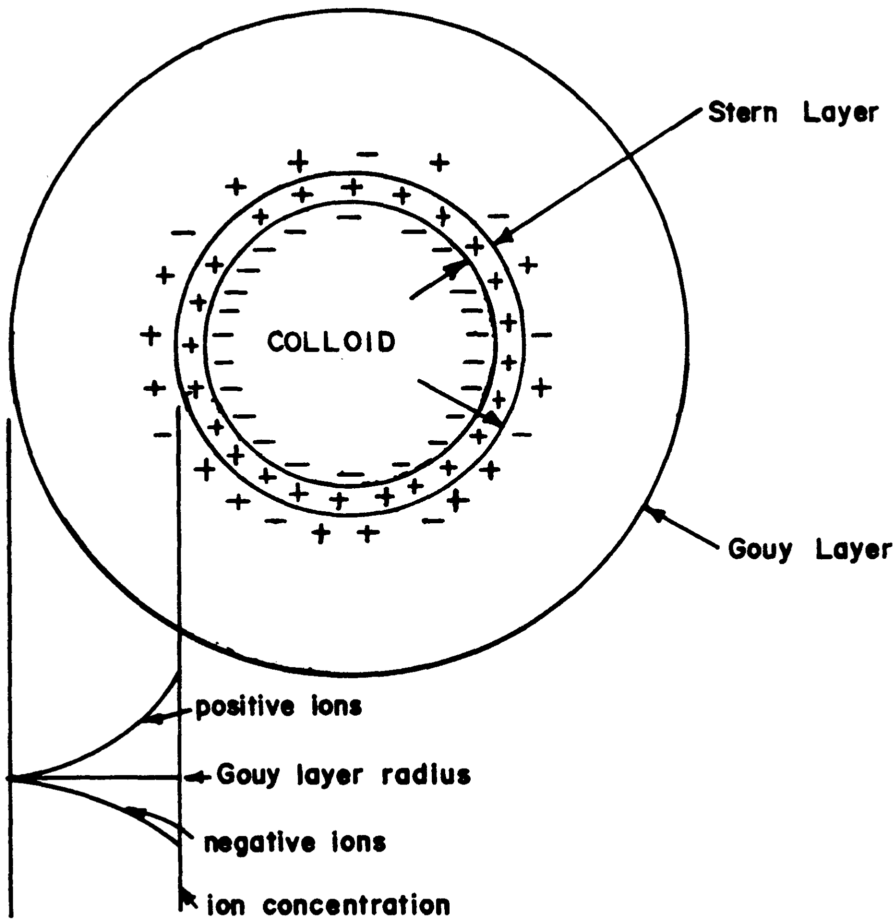
The finished water leaving the NTMWTP is pumped at rates in increments of 1000 gallons per minute (GPM) with the rate leaving the plant dependent on demand. For this reason the plant design will be in english units to be compatible with existing conditions at the NTMWTP.

Part of this thesis deals with the settling velocity of particles in water which is primarily dependent on the diameter of the settling particle. Since colloidal particles or aggregated colloids are measured in microns, metric units will be used whenever settling velocity calculations are performed. To convert microns to feet or inches would produce a measurement of an unsatisfactory scale.

COAGULATION

Coagulation is the process of destabilization or neutralization of the electrical charge on suspended colloidal particles in a raw water supply. The electrical charge of a colloid generates a repulsive force that prevents aggregation of the colloids. Once the electrical charge has been neutralized the colloidal particles can begin to aggregate into a size large enough to settle.

The work of Gouy, Chapman, and Stern explain the stability of colloids through the electric double layer theory[1]. The electric double layer theory states that a colloidal suspension does not have a net electrical charge; therefore the charge on the colloid must be counter balanced by ions of the opposite charge in the solution. A negatively charged particle will accumulate positively charged ions at its surface in a compact layer. This compact layer is called the Stern layer (see Figure 1). The Stern layer is surrounded by a diffuse layer of negative and positive ions in what is known as the Gouy layer (see Figure 1). The thickness of this diffuse layer is governed by the ionic strength of the solution. The higher the ionic



Colloidal Particle

Figure 1

strength the thinner the diffuse layer. The concentration of negative ions within the Gouy layer increases with an increase in distance from the colloid; whereas the concentration of positive ions will decrease (see Figure 1). Beyond the Gouy layer the concentrations of positive and negative ions are equal.

When two colloids of a similar nature come close to each other their diffuse layers overlap and electrostatically interact. This electrostatic interaction results in a repulsive force between them. The repulsive force increases as the colloids come closer together. A thick diffuse layer will lead to a greater electrostatic force.

Coagulation Mechanisms

Coagulation is accomplished by adding a coagulant along with intense mixing to destabilize the colloid. A coagulant is either a metal salt, or a high molecular weight non-ionic, cationic, or anionic polymer. This destabilization step occurs instantaneously after the coagulant is added. Because this reaction is instantaneous the level of mixing energy is critical to efficient coagulation. It is generally accepted that coagulation occurs through any of four different mechanisms[2]:

- Double Layer Compression
- Adsorption and Charge Neutralization
- Enmeshment by Precipitation (sweep floc)
- Adsorption and Interparticle Bridging

Double Layer Compression:

This method compresses the diffuse or Gouy layer that surrounds the colloid. This is accomplished by adding an electrolyte that increases the ionic strength of the solution. The charge of the electrolyte added is always the opposite of the charge of the colloid. This reaction takes place instantaneously. As the ionic strength of a solution increases the charge density in the diffuse layer increases and the thickness of the diffuse layer decreases. When the diffuse layer is thin enough the repulsive forces that would normally develop in the diffuse layer do not develop and rapid particle aggregation occurs.

This mechanism naturally occurs when relatively low ionic strength river water comes in contact with high ionic strength sea water, which causes the colloids in the river water to coagulate. The coagulated colloids then aggregate to a size large enough that they can settle out and form river deltas.

Adsorption and Charge Neutralization:

This method destabilizes colloids by electrostatic attraction. Electrostatic attraction occurs when surfaces are oppositely charged. This is accomplished by adding an electrolytic coagulant to the raw water where the positively charged coagulant ions adsorbs onto the negatively charged colloid thus neutralizing the net charge of the colloid. This reaction is much slower than double layer compression and takes between ten to thirty seconds to complete. Once the charge of the colloid has been neutralized the particles can aggregate to a size large enough to facilitate settling. This coagulation mechanism is predominant in most water treatment plants.

Overdosing a coagulant can result in a charge reversal of the colloid which will restore the electrostatic repulsive force. Practical experience and bench scale testing can determine the proper coagulant dosage.

Enmeshment by a Precipitant (Sweep Floc):

This method is accomplished by adding a metal salt such as alum ($Al_2(SO_4)_3 \cdot 14H_2O$) or ferric chloride ($FeCl_3$) to a high enough dosage to cause a precipitation of a metal hydroxide. Colloid particles become entrapped in the precipitant as it is formed and when they collide with the formed precipitant. This is also known as sweep floc coagulation. The disadvantage of sweep floc coagulation is that large amounts of a metal salt are necessary and it is not very effective for low turbidity waters because the amount of formed precipitant is not large enough to effectively trap the colloids.

Adsorption and Interparticle Bridging:

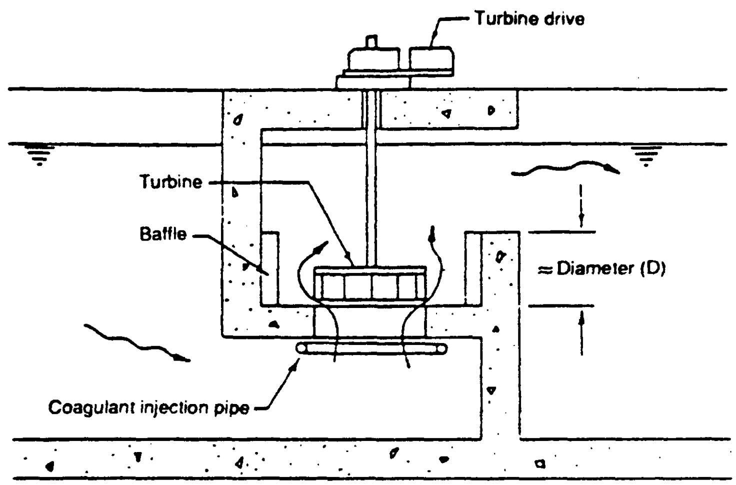
Adsorption and interparticle bridging is accomplished through the addition of a synthetic organic polymer to the raw water. Long high molecular weight cationic, anionic, and nonionic polymers have all proven to be effective. The type of polymer that is most effective is dependent on the characteristics of the raw water. When a polymer comes in contact with a colloid a portion of it will adsorb onto the colloid. Further colloids will attach themselves to other sites on the same polymer molecule thus forming particle-polymer aggregates where the polymer is the bridge. Eventually enough colloids will attach themselves to the polymer that settling will occur.

Experience has shown that double layer compression and charge neutralization are the two most predominant coagulation mechanisms when treating the raw water that supplies the North Table Mountain Water Treatment Plant (NTMWTP). Adsorption and interparticle bridging also play a role in the coagulation of NTMWTP raw water but effective coagulation is not possible by the addition of just a polymer. The use of alum as a primary coagulant in conjunction with a cationic polymer produces the best results.

Coagulation Mixing Units

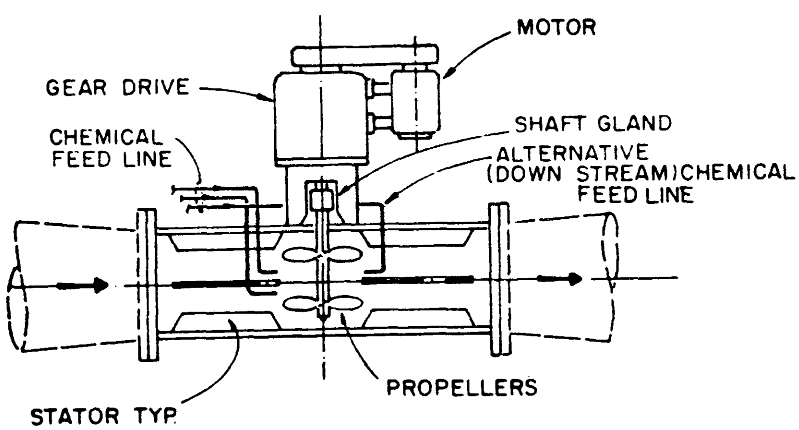
The purpose of a coagulant mixer is to provide the energy necessary for effective coagulation. The parameter used to measure the mixing energy is the velocity gradient ($\bar{G}$). Acceptable values for $\bar{G}$ range from 700 to 1000 seconds^{-1}. As double layer compression coagulation takes place instantaneously a $\bar{G}$ value as high as 10000 second^{-1} for a short duration is recommended[3] There are several types of commonly used mixing devices for coagulation:

- Backmix reactor (see Figure 2)
- In-line Blenders (see Figure 3)
- Hydraulic Jumps
- Motionless Static Mixers (see Figure 4)



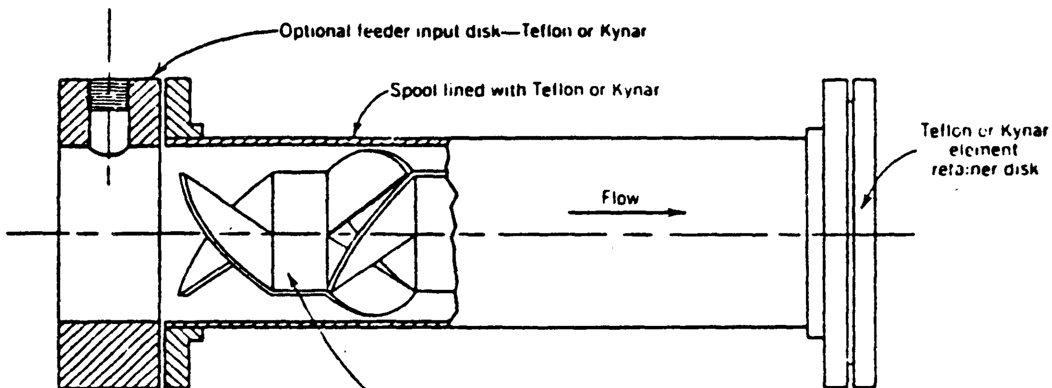
Back Mixer

Figure 2



In-Line Blender

Figure 3



Motionless Static Mixer

Figure 4

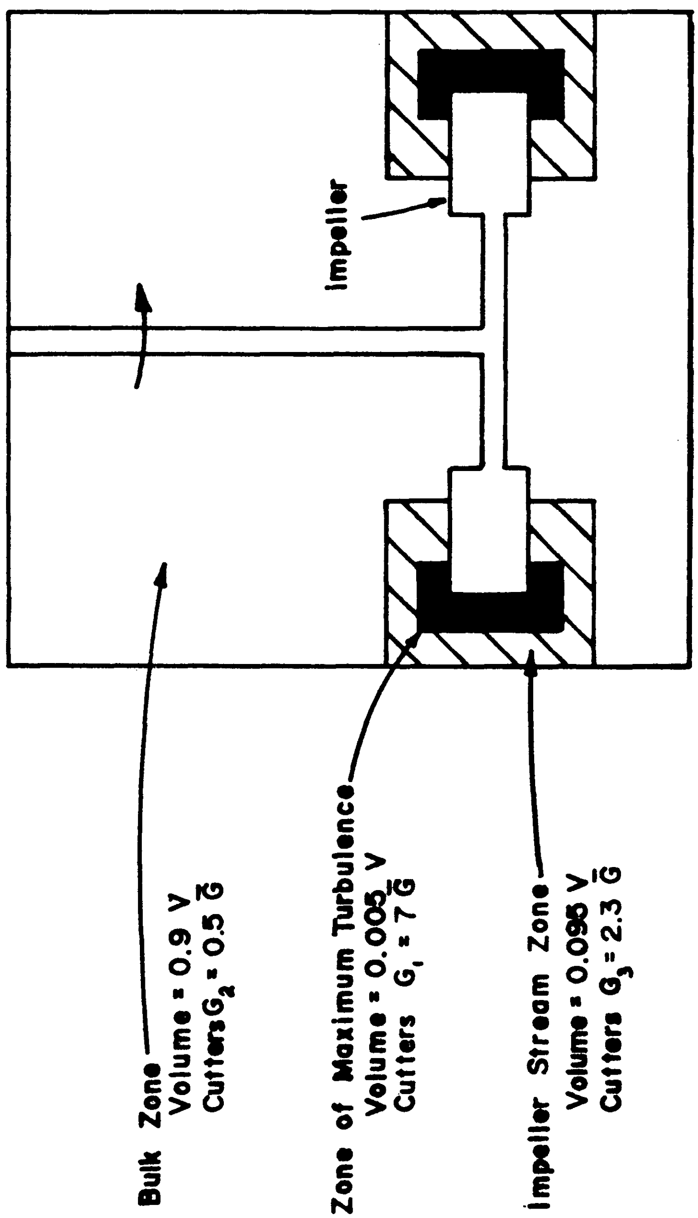
Backmixer:

This is the most common type of mixer in use today. The average velocity gradient or mixing intensity produced by these types of mixers is calculated using the equation developed by Camp and Stein[4].

$$(\overline{G}) = \left(\frac{P}{\mu * V}\right)^{0.5} \tag{1}$$

Where: $\bar{G}$= velocity gradient (seconds^{-1})
P= power input (ft*lb/sec)
μ= dynamic viscosity (lb*sec/ft^2)
V= volume (ft^3)

This equation is a simple bulk approximation of a back mixer as a whole unit and does not address mixing intensities of different parts of the mixing unit. Cutter[5] found that the mixing intensities or turbulence in a stirred tank are separated into three zones: 1) maximum turbulence intensity near the impeller, 2) impeller stream zone, and 3) bulk zone (see Figure 5). The energy dissipation in the three zones in terms of a factor of the $\bar{G}$ value as calculated by Camp and Stern's equation are shown in Figure 5.



Cutters Stirred Tank

Figure 5

Amirtharajah recommends the following guidelines for designing a mechanical back mixer[6]:

- A square vessel is superior in performance to a cylindric vessel.
- Stator baffles are advantageous.
- A flat bladed impeller performs better than a fan or propeller impeller.
- Chemicals introduced at the agitator blade level enhance coagulation.
- A variable speed motor allows the $\bar{G}$ value to be changed by the plant operator.

Inline Blenders:

These are manufactured devices and function by the same mechanics as a back mixer. The only difference between the two types of mixers is that the inline blender mechanism is enclosed (see Figure 3). Inline blenders have the advantage that little short circuiting occurs. Their disadvantage is that the working part of the mixer is internal and a total plant shut down is necessary for any repairs. Another disadvantage of a blender is that it cannot meet the recommended detention time for the coagulation process which is 10-30 seconds. This time allows for adsorption and charge neutralization to take place as well as ensuring that a thoroughly mixed coagulated water is sent to the flocculation step.

Hydraulic Jumps:

Hydraulic jumps provide a $\bar{G}$ value in the range of 800 seconds^{-1} [7]. They have the advantage of the absence of any moving parts as the jump is produced in a Parshall flume. Their disadvantage is that they provide no flexibility as the mixing intensity is fixed. The detention time provided by a hydraulic jump is also less than the recommended value.

Motionless Static Mixer:

The mixing turbulence for this type of mixer is produced by a series of fixed sloping vanes within the mixer (see Figure 4). A motionless static mixer has the advantage that the energy input is provided by any available head. The mixing energy intensity is related to the flow through the mixer which places a constraint on this type of mixer. A disadvantage of this type of mixer is that the mixing intensity is constant over the length of the mixer so that there is no point of intense mixing.

The type of mixer that is most appropriate for the NTMWTP is the back mixer for the following reasons:

1. Normal plant operations require that the flows through the plant be varied throughout the day. As the mixing intensity of a motionless static mixer is dependent upon the flow this type of mixer would not be appropriate.

2. The motionless static mixer and the hydraulic jump do not provide the recommended detention time.

3. The inline blender offers all of the advantages of the backmixer but any maintenance would require a total shut down of the plant.

4. A back mixer can be designed to provide a high mixing intensity zone with a very short detention time to provide the parameters necessary for double layer compression.

Backmixer Design

For effective sedimentation to occur the coagulation process must be effectively performed. To ensure effective coagulation minimum design constraints will be placed on the mixer design such that cost optimization cannot be performed. Cost optimization for turbidity removal will be done in a latter section of this thesis.

The three primary parameters for the design of a backmixer are the average velocity gradient, the initial intense mixing energy, and the detention time. For effective coagulation to take place the following parameters need to be met:

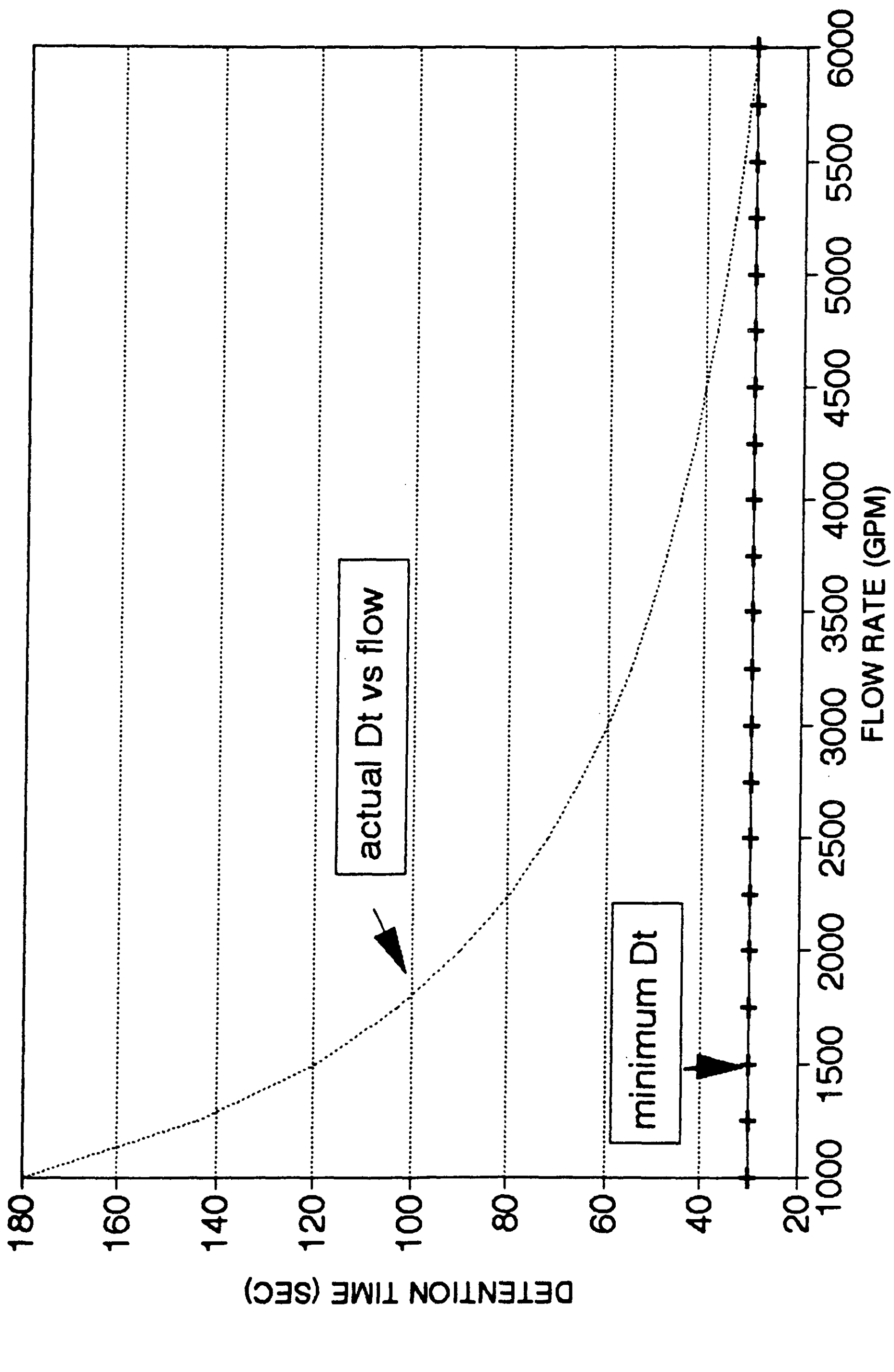
Parameter	Value
Average velocity gradient $\bar{G}$	700-1000 sec^{-1}
Initial intense mixing	up to 10000 sec^{-1}
Total detention time	10-30 sec
Intense mixing detention time	up to 1 sec

To ensure effective double layer compression coagulation the value for $\bar{G}$ will be set at a minimum of 1000 seconds^{-1}. The detention time will be set at the upper value of 30 seconds so that a well mixed and coagulated water is sent to the flocculation step.

The following constraints are placed on the NTMWTP and will be used to ensure that the above parameters are met:

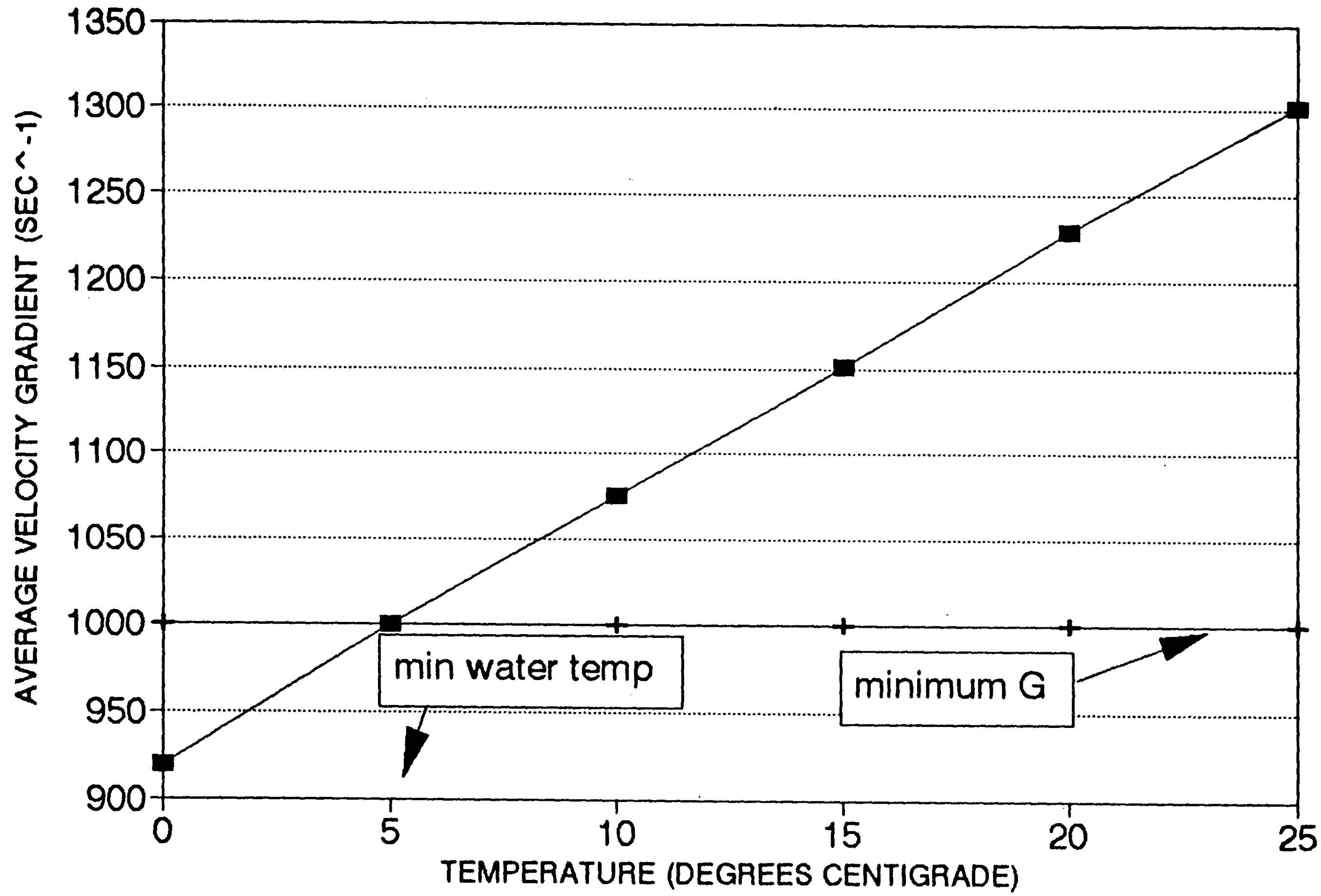
Constraint	Value
Maximum flow	6000 GPM
Minimum flow	2000 GPM
Maximum viscosity	$3.17*10^{-5}$ lb*sec/ft^2
Minimum viscosity	$2.20*10^{-5}$ lb*sec/ft^2

The following calculations use the maximum flow and the maximum raw water viscosity since these two parameters will provide the minimum $\bar{G}$ value. A $\bar{G}$ value above the recommended value will not have a detrimental effect on coagulation but more energy than necessary will be expended. Figure 6 shows how the detention time varies with flow. Figure 7 shows how the $\bar{G}$ value varies with water temperature as the viscosity of the water varies with temperature.



Coagulation Basin-Detention Time VS Flow

Figure 6



Coagulation Basin-Velocity Gradient
VS Water Temperature

Figure 7

VOLUME CALCULATION:

$$Volume = Flow * Detention\ Time\ (Dt)$$

$$Volume = \left(\frac{6000\ gal}{min}\right) * \left(\frac{1\ min}{60\ sec}\right) * \left(\frac{1\ ft^3}{7.48\ gal}\right) * (30\ sec\ Dt) = 400\ ft^3$$

POWER REQUIRED (CAMP & STEIN EQUATION):

$$Power\ required = (\overline{G})^2 * \mu * volume$$

$$Power = (1000\ sec^{-1})^2 * \left(3.17\ x\ 10^{-5}\ \frac{lb*sec}{ft^2}\right) * (400\ ft^3) = 12680\ \frac{ft*lb}{sec}$$

$$HP = \left(12680\ \frac{ft\ *\ lb}{\text{sec}}\right) \div \left(\frac{550\ ft\ *\ lb/\text{sec}}{HP}\right) = 23.1\ HP$$

Assuming an 80% efficient motor:

$$(23.1\ HP)/(0.80) = 28.75\ HP$$

This suggests that a readily available 30 HP motor is appropriate

INITIAL INTENSE MIXING:

The following calculations are based on the work by Cutter[5] (see Figure 5).

$$Cutters\ G = 7 * (\bar{G})$$
$$Cutters\ G = 7 * (1000\ sec^{-1}) = 7000\ sec^{-1}$$

$$Volume = Dt * flow$$
$$Volume = (.005) * 400\ ft^3 = 2\ ft^3$$

$$Dt = 2\ ft^3 * \left(\frac{min}{6000\ gal}\right) * \left(\frac{60\ sec}{min}\right) * \left(\frac{7.48\ gal}{ft^3}\right) = .15\ sec$$

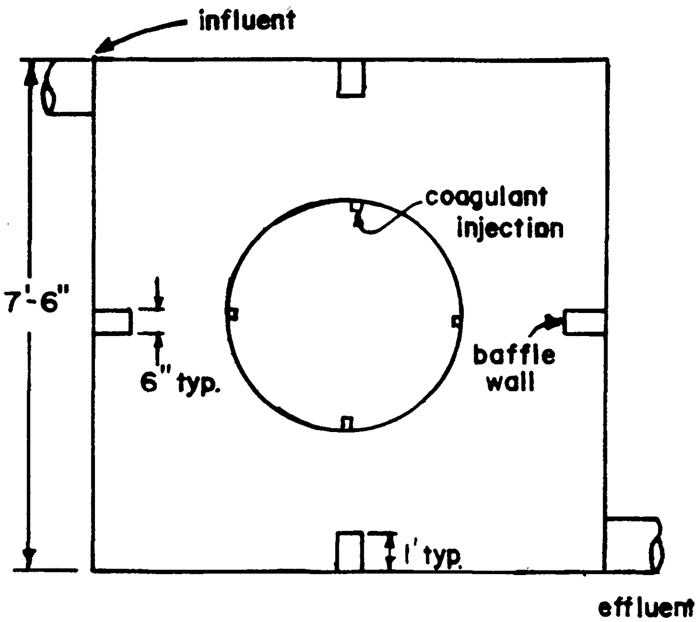
Dt less than 1 second meets suggested design parameter (see design parameters)

RECOMMENDED BACKMIXER DESIGN:

Volume: 400 ft^3, square in shape with baffles

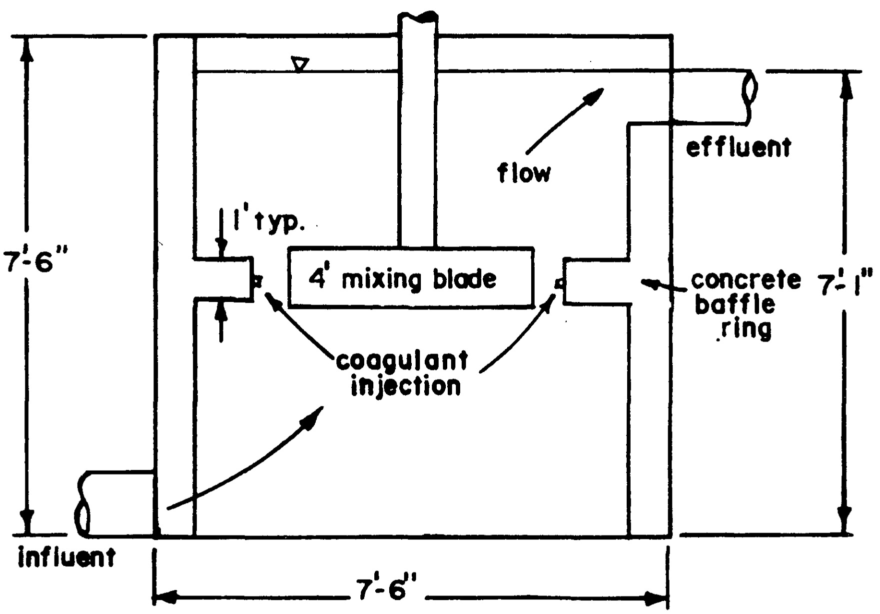
Mixing motor: 30 HP motor with a variable speed drive to allow the $\bar{G}$ to remain at the recommended value of 1000 $seconds^{-1}$ regardless of the incoming flow or water temperature.

The coagulant will be introduced in the zone of maximum turbulence to ensure double layer compression coagulation. (see Figures 8 & 9)



Back Mixer (Plan View)

Figure 8



Back Mixer (Elevation View)

Figure 9

FLOCCULATION

After the raw water has been effectively coagulated it is then flocculated. Flocculation is a transport process that brings about the collisions necessary to aggregate the destabilized colloids. The aggregated particles are typically called floc. The rate at which aggregation occurs is governed by the rate of collisions. As floc particles grow in size, shearing forces begin to cause the floc to break up. In a properly flocculated water the formation and break up of floc particles will reach a steady state and the distribution of floc particle size will remain constant.

Flocculation Transport Processes

There are three transport processes which are:

Perikinetic Transport Process:

Perikinetic flocculation is brought about by random thermal motion or Brownian motion of water molecules causing collisions between the floc particles.

Orthokinetic Transport Process:

Orthokinetic flocculation is accomplished by imparting a velocity gradient to the water to increase the number of floc particle collisions.

Differential Settling Transport Process:

Differential settling flocculation occurs because two different sized particles will settle at a different velocity and one particle will overtake the other causing a collision.

The collision frequency function (k) for each transport process are calculated as follows:[8]

Perikinetic:

$$k_p = \left(\frac{2}{3}\right) * \left(\frac{k_b * T}{\mu}\right) * \frac{(d_i + d_j)^2}{(d_i * d_j)} \tag{2}$$

Orthokinetic:

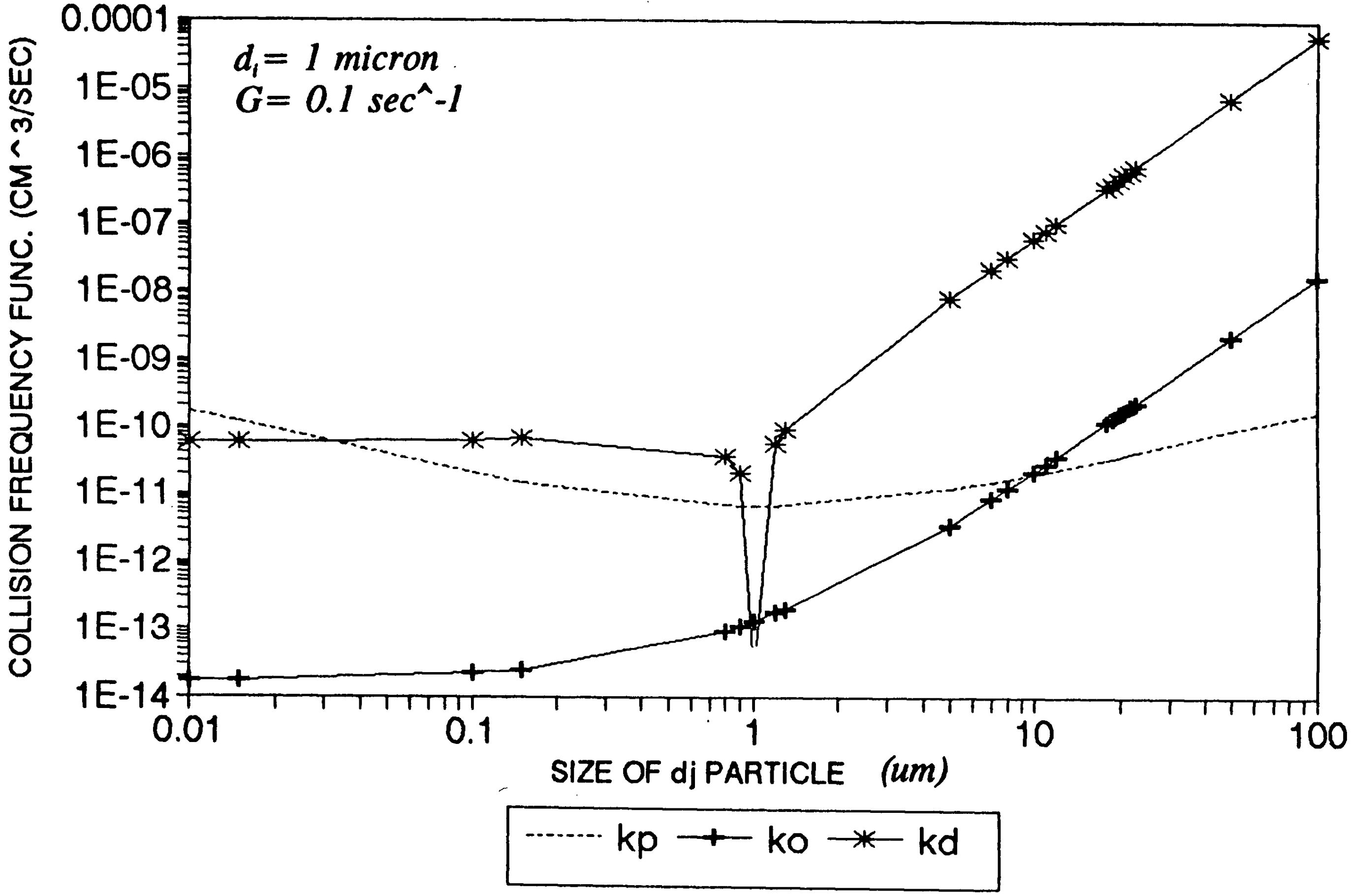
$$k_o = \left(\frac{1}{6}\right) * (d_i + d_j)^3 * \overline{G} \tag{3}$$

Differential Settling:

$$k_d = \left(\frac{\pi * g}{72}\right) * \frac{(SG-1)}{\nu} * (d_i + d_j)^3 * |(d_i - d_j)| \tag{4}$$

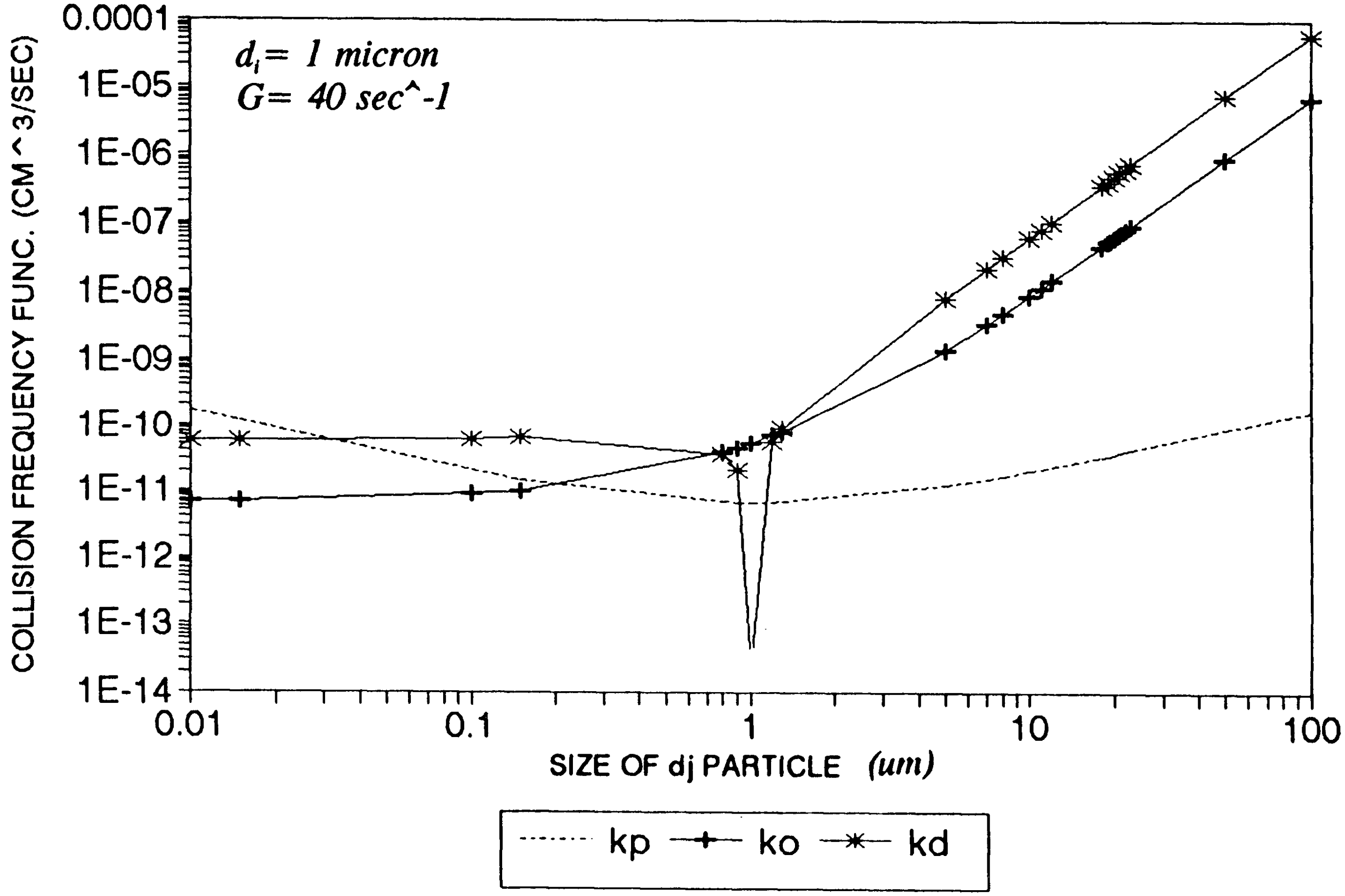
Where:
- k= collision frequency function (cm^3/sec)
- k_b= Boltzman's constant ($Kg*cm^2/sec^2$ °K)
- T= absolute temperature (°K)
- μ= dynamic viscosity ($N*sec/cm^2$)
- d_i= particle diameter of size i (cm)
- d_j= particle diameter of size j (cm)
- $\overline{G}$= velocity gradient (sec^{-1})
- g= gravity constant (cm/sec^2)
- SG= specific gravity of the particle
- v= kinematic viscosity (cm^2/sec)

Equations 2 through 4 were plotted using various imparted $\bar{G}$ values and particle sizes. In Figures 10-15 the particle size d_j is varied along the x-axis. In the upper left hand corner of the graph the chosen value for the particle size d_i and the chosen value of $\bar{G}$ are shown. Following is a discussion on the three transport mechanisms and how Figures 10-15 show the effectiveness of them under different conditions.



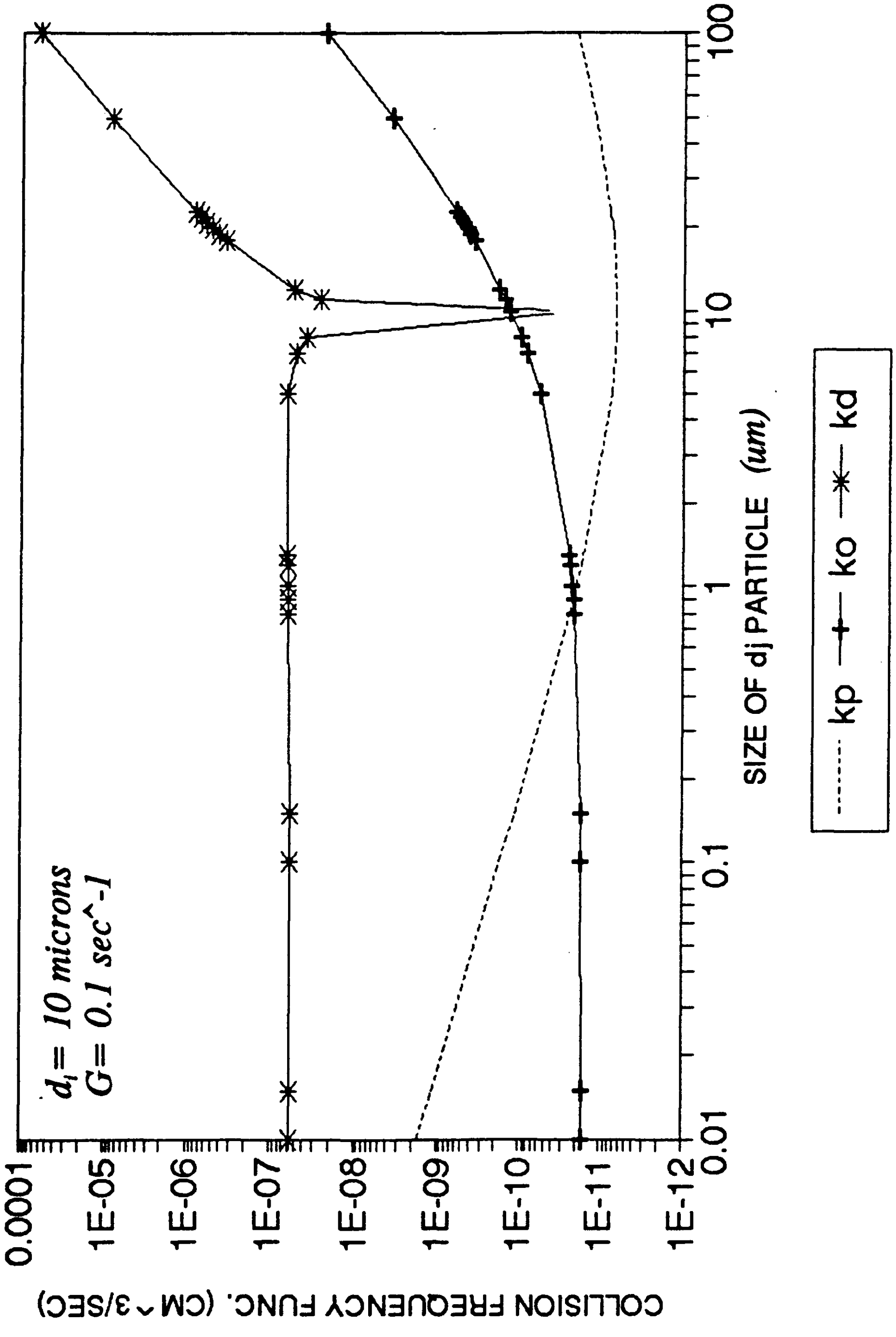
Collision Frequency Function
(d_i=1 micron, $\bar{G}$=0.1 sec^{-1})

Figure 10



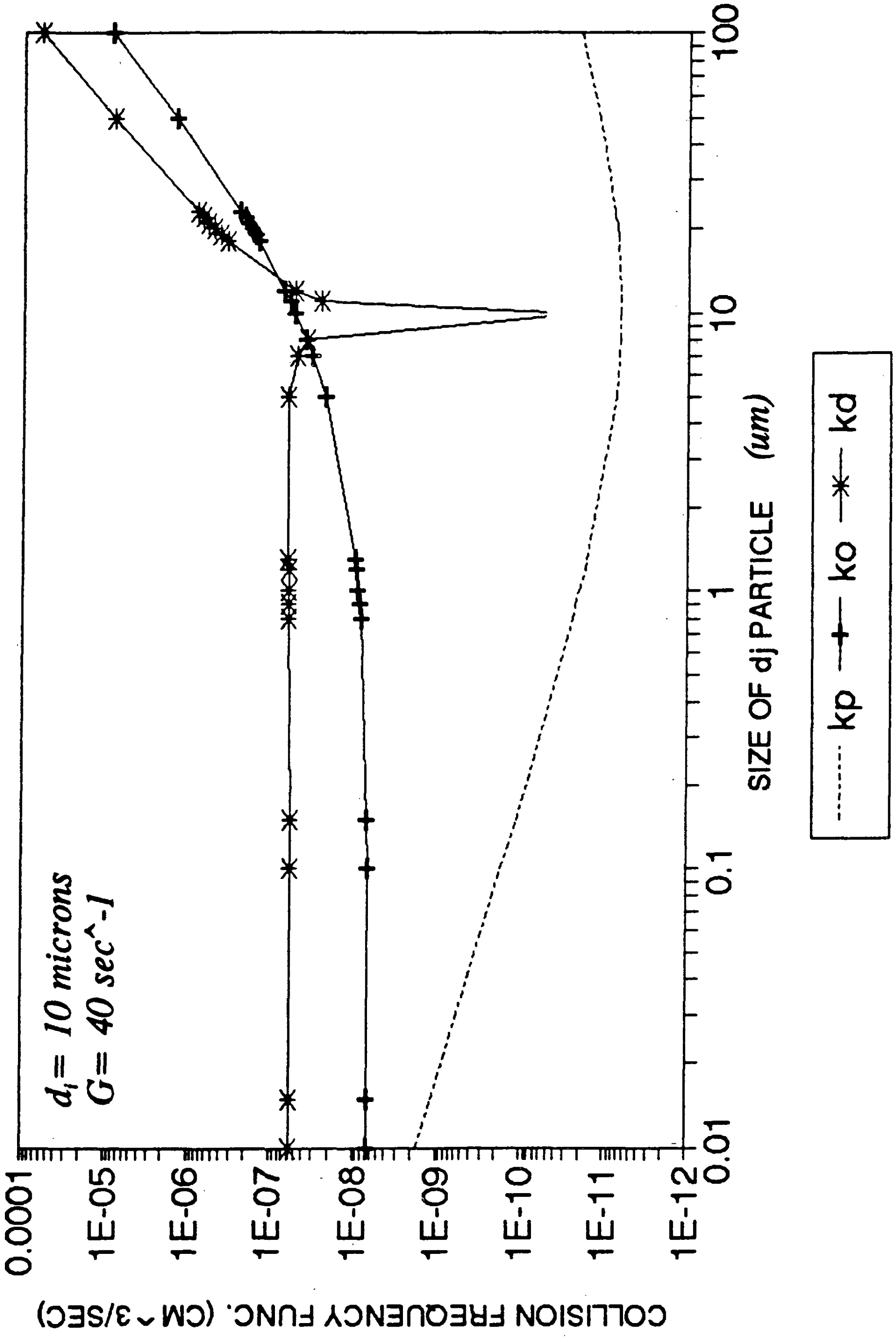
Collision Frequency Function
(d_i=1 micron, $\bar{G}$=40 sec^{-1})

Figure 11



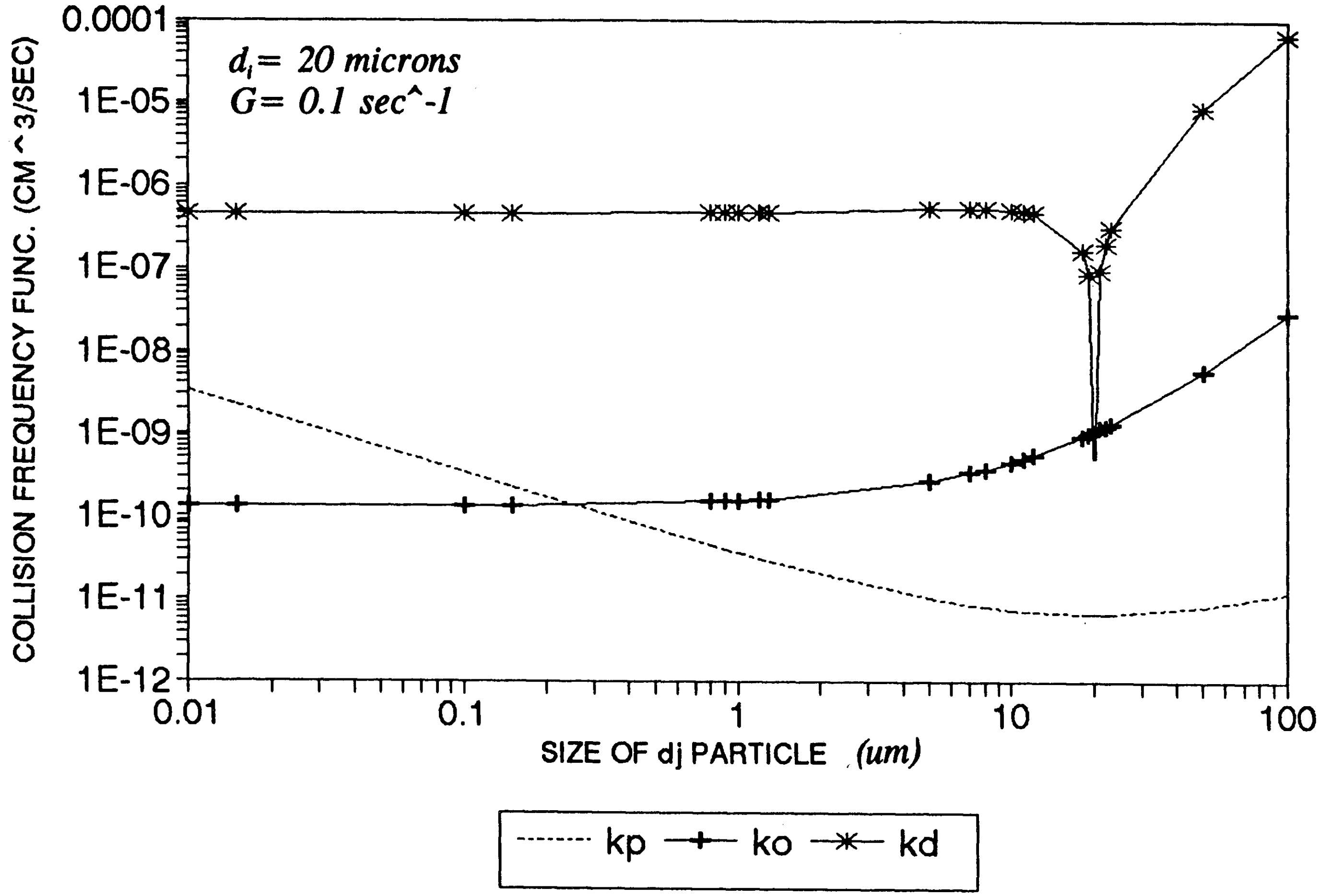
Collision Frequency Function
(d_i=10 microns, $\bar{G}$=0.1 sec^{-1})

Figure 12



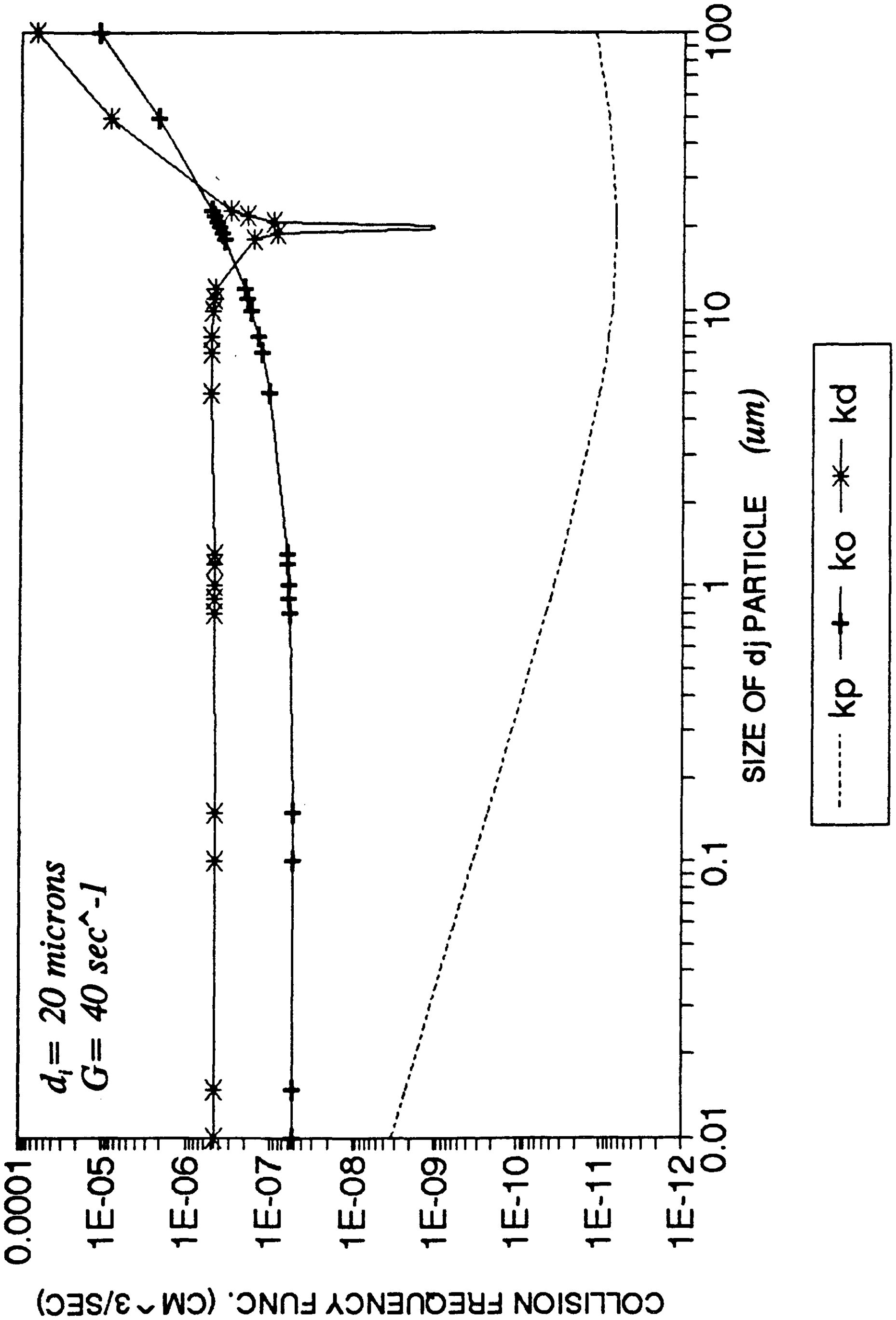
Collision Frequency Function
(d_i=10 microns, $\bar{G}$=40 sec^{-1})

Figure 13



Collision Frequency Function
(d_i=20 microns, $\bar{G}$=0.1 sec^{-1})

Figure 14



Collision Frequency Function
(d_i=20 microns, $\bar{G}$=40 sec^{-1})

Figure 15

Flocculation Mechanisms

Perikinetic Flocculation:

The driving force for this type of flocculation is the random thermal motion or Brownian motion of water molecules. This mechanism is only predominant for particles in the 1-2 micron size range[7]. In Figures 10 and 11 particle size d_i is set at 1 micron. As shown in Figures 10 and 11 perikinetic flocculation is one of the predominant transportation mechanisms when the particle size is less than 1 micron. For a coagulated water that has a large number of particles smaller than 1 micron this transport mechanism is critical because orthokinetic flocculation is not effective with particles smaller than 1 micron (see Figure 10). When a velocity gradient $\bar{G}$ is imparted to a water and the particle size is much greater than 1 micron this mechanism has only minor influence on particle transport (see Figures 13 & 15).

Orthokinetic Flocculation:

This transport mechanism is accomplished by imparting a velocity gradient to the water through mechanical or hydraulic mixing. Orthokinetic flocculation is the predominant transport mechanism used in flocculation for particles greater than 2 microns. Figures 11, 13, and 15, where the imparted G is 40 seconds^{-1}, show the predominance of orthokinetic flocculation when the particle size on the x-axis is greater than 2 microns. The value of the $\bar{G}$ imparted to the water cannot be so high as to cause excessive shearing of the floc. The generally acceptable range for the $\bar{G}$ imparted to the water is 30 to 50 seconds^{-1}.

Differential Settling:

The transport mechanism in differential settling is caused by different sized particles settling at different rates. The settling rate of particles with the same density is proportional to their diameter squared. The preceding statement is only true when the velocity of the settling particles is relatively slow, as is the case for settling floc particles. The settling velocity must be slow so that the Reynolds number which is a function of velocity is small. When the Reynolds number for a sphere settling in a liquid is between 0.5 to 2 Equation 7 holds true and allows the following derivation[9].

Equations 5-8 demonstrate why the settling rate of particles with the same density is proportional to their diameter squared.

The terminal velocity (V_t) of a sphere settling in a liquid is:

$$V_t = \left[\frac{4 * g * (\rho_s - \rho) * d}{3 * C_d * \rho}\right]^{0.5} \tag{5}$$

The Reynolds number (Re) for a settling sphere is:

$$R_e = \frac{\rho * V_t * d}{\mu} \tag{6}$$

According to Stokes Law the coefficient of drag for laminar flow is[9]:

$$C_d = \frac{24}{R_e} \tag{7}$$

Where the variables in Equations 5-7 are:

V_t= terminal velocity (m/sec)
g= gravity constant (m/sec^2)
ρ_s= density of the sphere (kg/m^3)
ρ= density of the liquid (kg/m^3)
d= diameter of the sphere (m)
C_d coefficient of drag (unitless)
μ= dynamic viscosity (N*sec/m^2)
R_e= Reynolds number (unitless)

Substituting Equation 6 into Equation 7 and then substituting that result into Equation 5 yields:

$$V_t = \frac{g * (\rho_s - \rho) * d^2}{18 * \mu} \tag{8}$$

Equation 8 shows that the settling velocity of two particles of the same density is proportional to their diameter squared.

Differential settling can be an effective transport mechanism in a flocculated water with a wide range of particle sizes. When the particle size d_j on the x-axis of Figures 10-15 approaches the particle size d_i found in the upper left hand corner of the graph, Figures 10-15 show that differential settling is not very effective. The reason for this is that two similar sized particles will settle at approximately the same rate and one particle will not overtake the other causing a collision.

Of the three types of transport mechanisms available for flocculation, orthokinetic flocculation is the only mechanism that can be controlled. Perikinetic and differential settling are simultaneously functioning along with orthokinetic flocculation but there is no control over their effectiveness. An energy gradient can be imparted to the water that is being flocculated to control the degree of orthokinetic flocculation. As can be seen in Equation 3 the collision frequency function can be increased by increasing the value of $\bar{G}$. The net effect of increasing $\bar{G}$ is an increase in the number of collisions between floc particles increasing their size which enhances sedimentation. If the imparted $\bar{G}$ is greater than the recommended value the floc particles will shear or break up and hinder sedimentation.

A G can be imparted to the water either hydraulically or mechanically. The equation to calculate the velocity gradient when it is hydraulically imparted to the water in a motionless static mixer is as follows[4]:

$$(\overline{G}) = 178 * \left(\frac{H}{t}\right)^{0.5} \tag{9}$$

Where: $\bar{G}$= velocity gradient ($seconds^{-1}$)
178= empirical constant
H= head loss (feet)
t= theoretical detention time (seconds)

Hydraulic flocculation has the advantage of not requiring any energy input but the velocity gradient is dependent upon the flow through the flocculation basin. For a water treatment plant with a constant flow hydraulic flocculation would be appropriate.

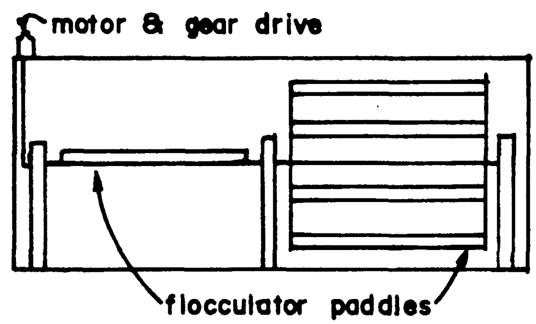
The equation used to calculate the velocity gradient when a mechanical flocculator is used is the same equation as that used for a back mixer and is as follows:

$$(\overline{G}) = \left(\frac{P}{\mu * V}\right)^{0.5} \tag{10}$$

Where: $\bar{G}$= velocity gradient ($seconds^{-1}$)
P= power input (ft*lb/second)
μ= dynamic viscosity ($lb*sec/ft^2$)
V= volume (ft^3)

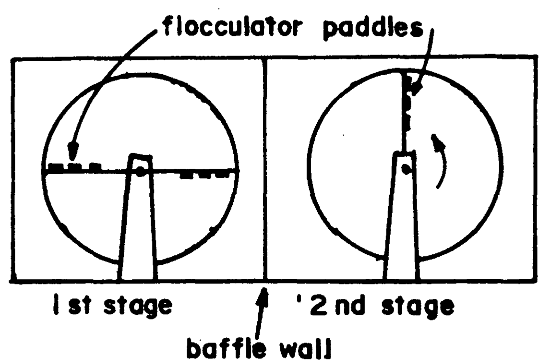
The flocculation mechanism type that is most appropriate for the NTMWTP is a mechanical flocculator for the following reasons:

1. Normal plant operations require that the flows through the plant be varied throughout the day. Since the velocity gradient for hydraulic flocculation is dependent upon the flow this type of flocculator would not be appropriate.
2. A variable speed mechanical flocculator (see Figures 16a & 16b) would allow the velocity gradient to be varied to optimize the flocculation step on a plant scale.



Mechanical Flocculator
(longitudinal view)

Figure 16a



Mechanical Flocculator
(cross section)

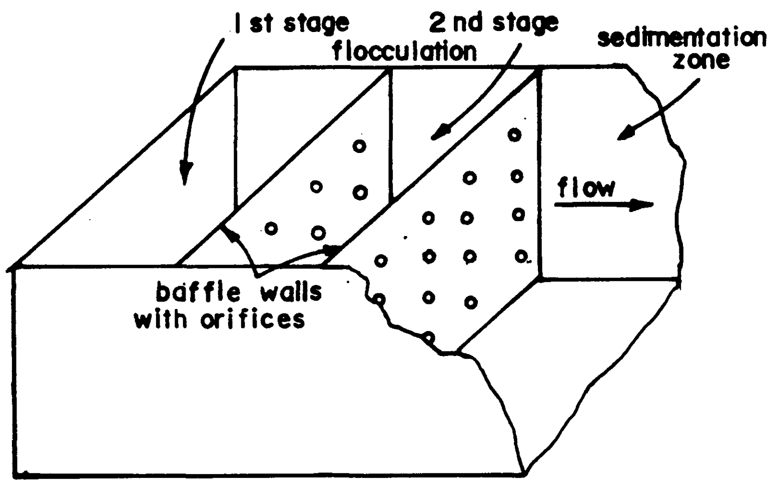
Figure 16b

Mechanical Flocculator Design

For effective sedimentation to occur the flocculation process must be optimally performed. To ensure effective flocculation minimum design constraints will be placed on the flocculator design such that cost optimization cannot be performed. Cost optimization will be done in a latter section of this thesis.

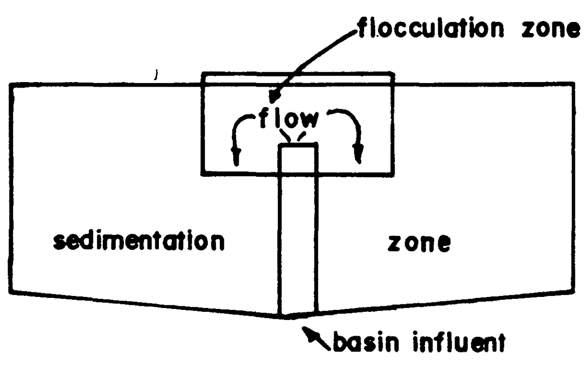
Pilot scale flocculation testing has shown that flocculation basins in series greatly enhance flocculation[10]. If a circular clarifier or sedimentation basin is used it is generally accepted that only one flocculation basin be used.

The suggested number of rectangular flocculation basins ranges from two to four basins and is dependent on plant size. For reasons of cost, two basins are generally acceptable for a small plant. The individual flocculation basins and the sedimentation basin are compartmentalized to prevent short circuiting. Short circuiting can be minimized by maintaining an orifice ratio between compartments of approximately 3 to 6 percent of the flow area[11] (see Figure 17a). In a circular basin the flocculated water is allowed to flow through the bottom of the flocculation basin so there are no orifices (see Figure 17b).



Flocculation Baffle Walls

Figure 17a



Circular Basin Flow
(elevation view)

Figure 17b

Tapering the velocity gradient in the flocculation compartments will form an optimum floc particle that will exhibit good settling characteristics[12]. The velocity gradient in the first compartment is higher than in the second to allow orthokinetic flocculation to form the initial floc as rapidly as possible. The velocity gradient is then lowered in the second compartment to prevent excessive floc shear but still allowing for orthokinetic flocculation. As previously stated the acceptable value for $\bar{G}$ ranges from 30 to 50 seconds^{-1} and any flocculator design should fall in this range.

Andreu-Villegas and Letterman have developed an empirical equation to calculate the optimum velocity gradient based on the coagulant dosage and the flocculation basin detention time and it is as follows[13]:

$$(\overline{G*})^{2.8}=\left(\frac{4.4*10^6}{C*Dt}\right) \qquad (11)$$

Where: $4.4*10^6$= empirical constant $(\text{sec*min*mg/L})^{1/2.8}$
$\bar{G}^*$= optimum mean velocity gradient
C= alum dosage concentration (10-50 mg/L)
Dt= theoretical detention time (minutes)

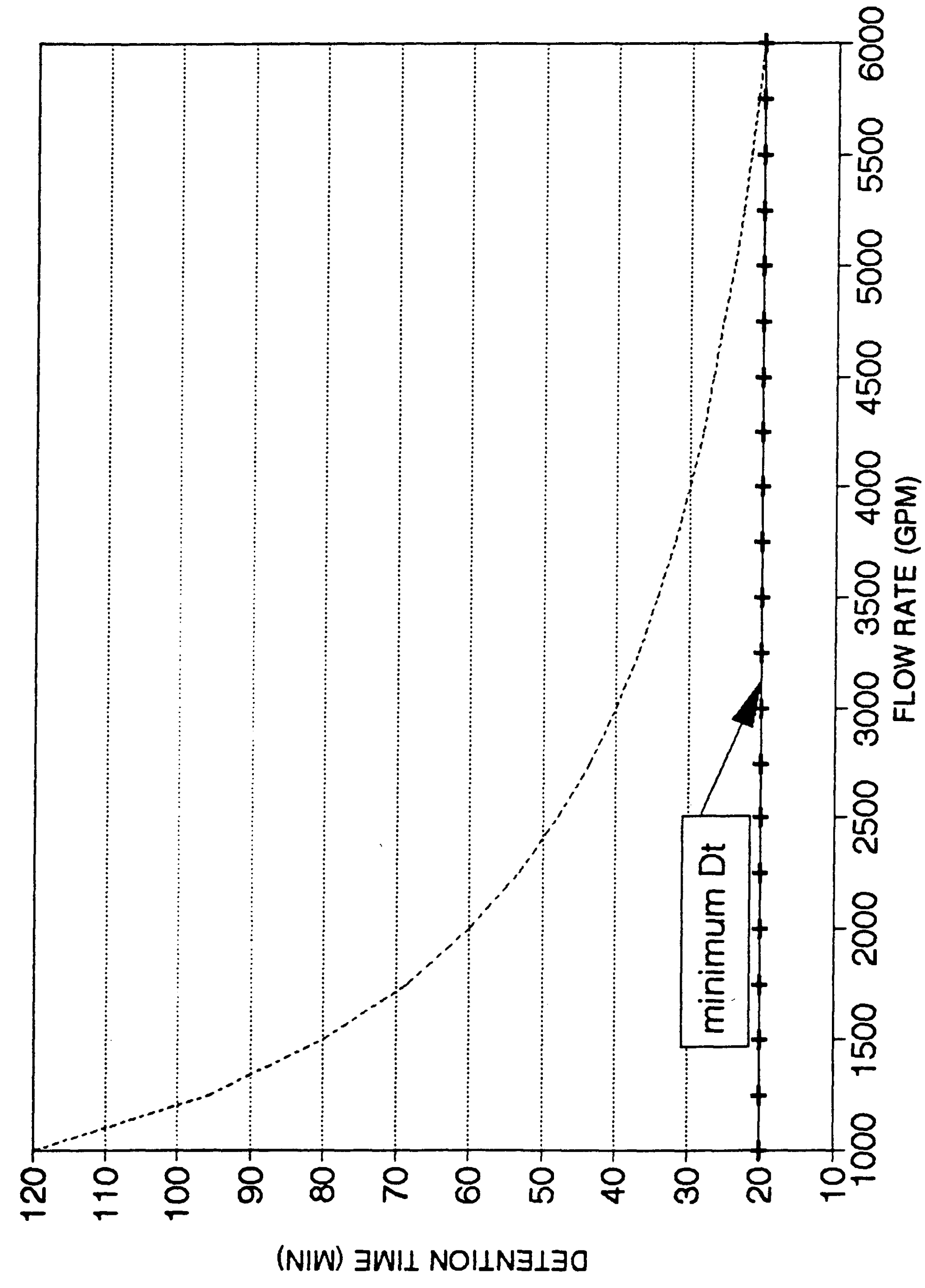
As Equation 11 is an empirical equation the units associated with the empirical constant are meaningless.

An over design factor for G^* of 1.4 is recommended[14]. The theoretical detention time for flocculation is generally accepted to be 20 minutes at maximum flow[15]. Longer detention times have been used in the past but no significant improvement in flocculation had been observed. For the NTMWTP the following design parameters will be used:

Parameter	Value
Average velocity gradient $\bar{G}$	30-50 $seconds^{-1}$
Total detention time	20 minutes
Velocity gradient * Detention time	greater than 30,000
Number of compartments	2 rectangular basin or 1 circular basin
Orifice ratio between compartments	3%-6%
Maximum velocity through orifices	less than 1 ft/sec

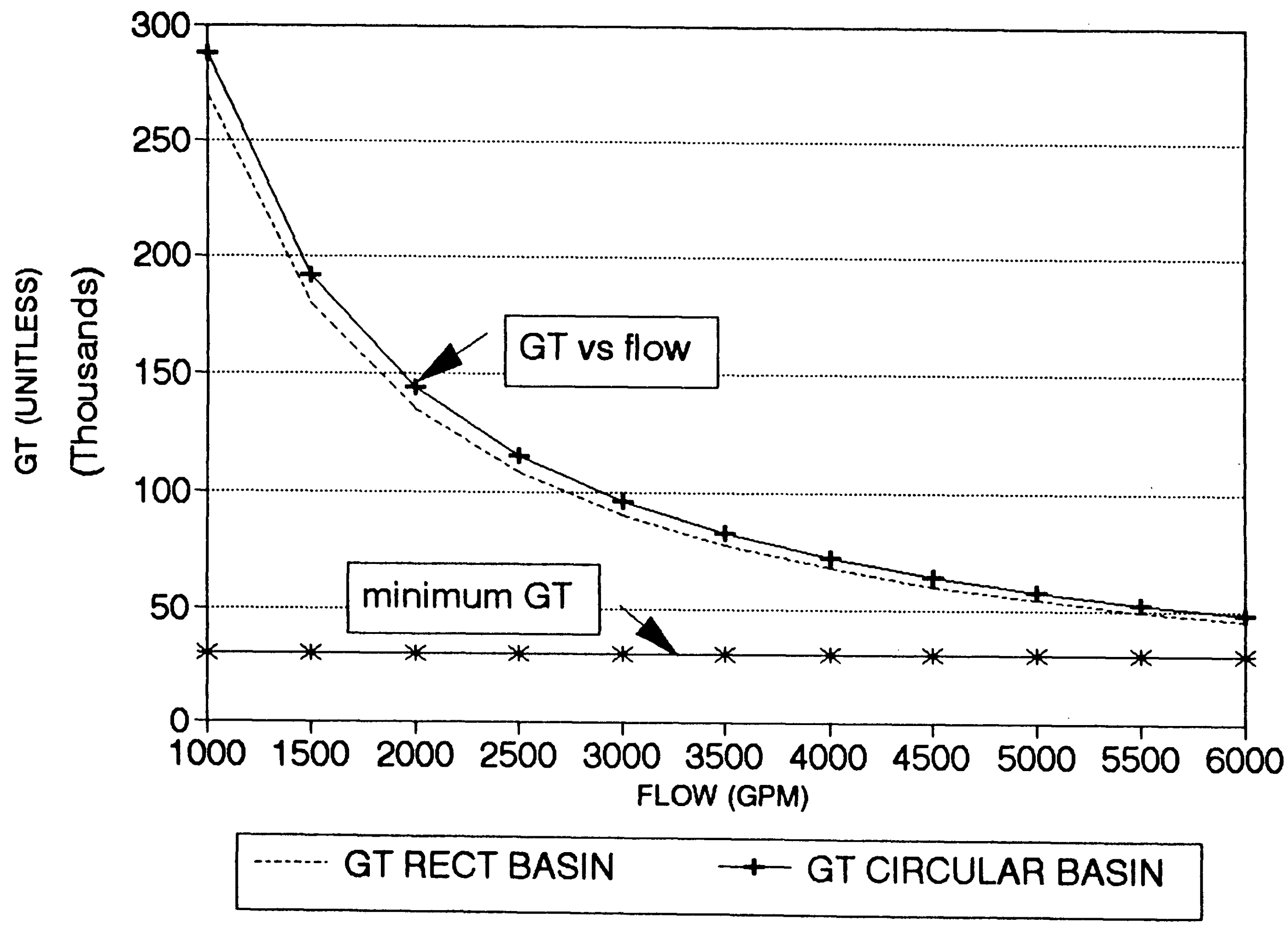
The following constraints are placed on the NTMWTP and will be used to ensure that the above parameters are met (see Figures 18 through 20). The minimum water temperature at the NTMWTP is 41° Fahrenheit (5° C) and the maximum water temperature is 68° Fahrenheit (20° C).

Constraint	Value
Maximum flow	6000 GPM
Minimum flow	2000 GPM
Maximum viscosity	$3.17*10^{-5}$ $lb*sec/ft^2$
Minimum viscosity	$2.20*10^{-5}$ $lb*sec/ft^2$



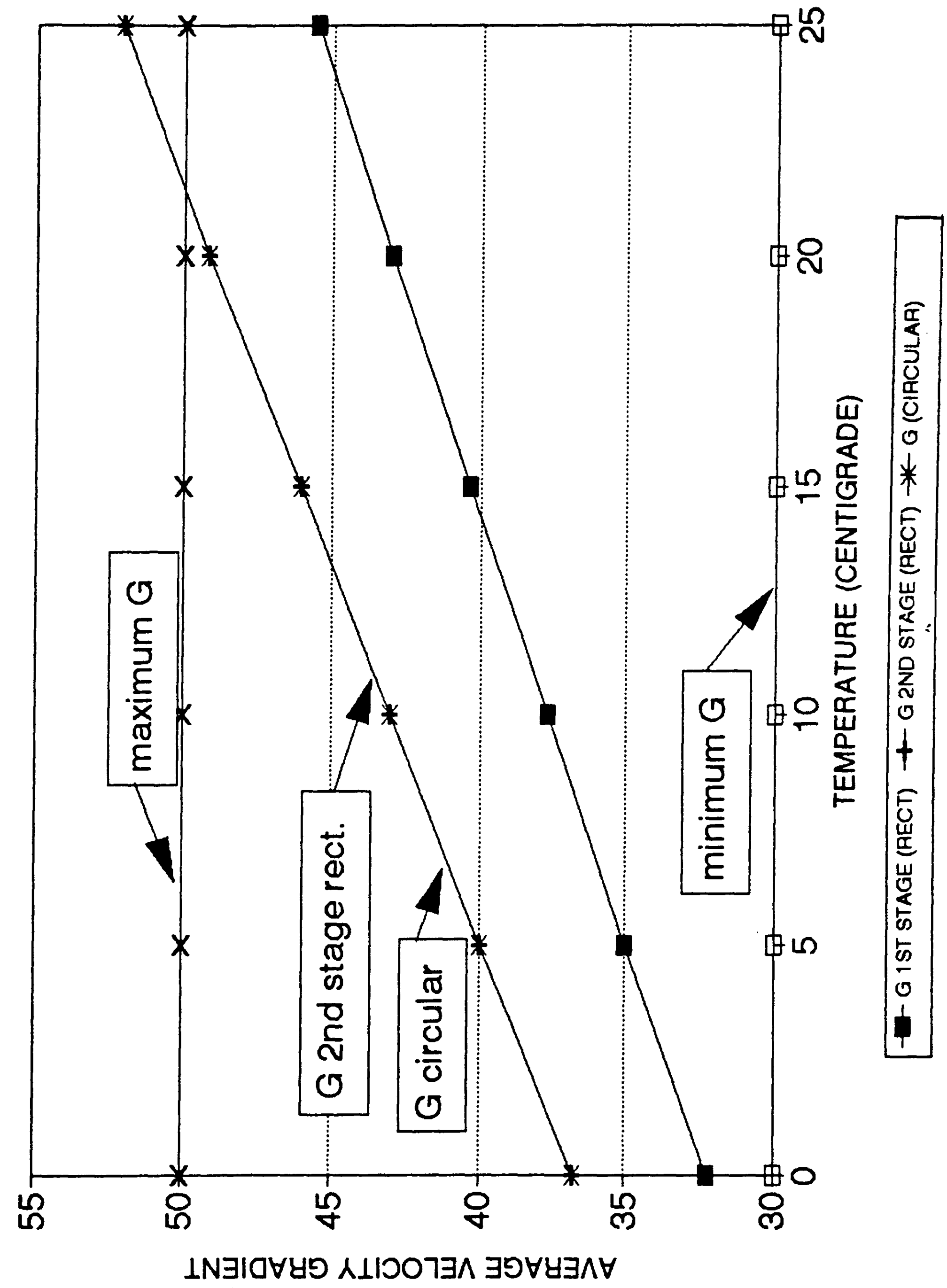
Flocculation Basin-Dt VS Flow

Figure 18



Flocculation Basin-Gt VS Flow

Figure 19



Flocculation basin
Water Temp. VS $\bar{G}$

Figure 20

The following calculations use the maximum flow and the maximum raw water viscosity since these two parameters will provide the minimum $\bar{G}$ value and the maximum power input. Unlike coagulation a $\bar{G}$ value above the recommended range will have a detrimental effect on flocculation in that it will cause excessive floc shear. After the flocculation motor has been sized, the maximum calculated power input and the minimum water viscosity will be used to calculate the maximum $\bar{G}$. This value will then be compared to the recommended $\bar{G}$ value range to ensure that excessive floc shear will not occur.

OPTIMUM MEAN VELOCITY CALCULATION:

The alum dosage for the NTMWTP ranges from 20-30 mg/L. The dosage varies with the quality of the raw water being treated. Using Equation 11 with a theoretical detention time of 20 minutes and an alum dosage range of 20-30 mg/L the optimum mean velocity gradient was calculated as follows:

Alum dosage of 20 mg/L:

$$(\overline{G*})^{2.8}=\left(\frac{4.4*10^{6}}{C*Dt}\right)$$

$$(\overline{G*})^{2.8}=\frac{4.4*10^{6}}{(20\ mg/L)\ *\ (20\ min)}=11000$$

$$(\overline{G*})=27.8\ seconds^{-1}$$

Alum dosage of 30 mg/L

$$(\overline{G*})^{2.8}=\left(\frac{4.4*10^{6}}{C*Dt}\right)$$

$$(\overline{G*})^{2.8}=\frac{4.4*10^{6}}{(30\ mg/L)\ *\ (20\ min)}=7733$$

$$(\overline{G*})=24.0\ seconds^{-1}$$

The recommended[14] over design factor for the optimal mean velocity gradient is 1.4. Therefore for an alum dosage of 20 mg/L the $\bar{G}$ is 38.9 seconds^{-1}, and for an alum dosage of 30 mg/L the $\bar{G}$ is 33.6 seconds^{-1}.

As tapered velocity gradient flocculation is recommended for a rectangular basin, a $\bar{G}$ for the first compartment will be 40 seconds^{-1} and the $\bar{G}$ value for the second compartment will be 35 seconds^{-1}. Separate calculations will be made for a circular basin with one flocculation compartment.

VOLUME CALCULATION (RECTANGULAR BASIN):

Two compartments will be used with an equal detention time between them for a total detention time of 20 minutes.

Compartments 1 & 2

$$Volume = flow * detention\ time$$

$$V = \left(\frac{6000\ gal}{min}\right) * \left(\frac{1 ft^3}{7.48\ gal}\right) * (10\ min) = 8021\ ft^3$$

VOLUME CALCULATION (CIRCULAR BASIN):

As previously stated a circular basin has only one compartment with a design detention time of 20 minutes at maximum flow.

$$Volume = flow * detention\ time$$

$$V = \left(\frac{6000\ gal}{min}\right) * \left(\frac{1 ft^3}{7.48\ gal}\right) * (20\ min) = 16042\ ft^3$$

POWER REQUIREMENT CALCULATION (RECTANGULAR BASIN):

1st compartment:

$$Power\ required = (\overline{G})^2 * \mu * volume$$

$$Power = (40\ sec^{-1})^2 * (3.17 * 10^{-5}\ \frac{lb*sec}{ft^2}) * (8021\ ft^3) = 406.8 \frac{ft*lb}{sec}$$

$$HP = \left((406.8\ \frac{ft*lb}{sec}\right) \div \left(\frac{550\ ft*lb/sec}{HP}\right) = 0.74 HP$$

Assuming an 80% efficient motor:

$$(.74\ HP)/(.80) = 0.92\ HP$$

This suggests that a readily available 1 HP motor is

2nd compartment:

$$Power = (35\ sec^{-}1)^2 * (3.17 * 10^{-5}\ \frac{lb*sec}{ft^2}) * (8021\ ft^3) = 311.5 \frac{ft*lb}{sec}$$

$$HP = \left((311.5\ \frac{ft*lb}{sec}\right) \div \left(\frac{550\ ft*lb/sec}{HP}\right) = 0.57 HP$$

Assuming an 80% efficient motor:

$$(.57\ HP)/(.80) = 0.71\ HP$$

This suggests that a readily available .75 HP motor is appropriate.

POWER REQUIREMENT CALCULATION (CIRCULAR. BASIN):

As previously calculated for an alum dosage of 20 mg/L and an over design factor of 1.4 use a $\bar{G}$ value of 40 seconds^{-1}.

$$Power\ required = (\overline{G})^2 * \mu * volume$$

$$Power = (40\ sec^{-}1)^2 * (3.17*10^{-5}\ \frac{lb*sec}{ft^2}) * (16042\ ft^3) = 813.6\frac{ft*lb}{sec}$$

$$HP = \left((813.65\ \frac{ft*lb}{\text{sec}}\right) \div \left(\frac{550\ ft*lb/\text{sec}}{HP}\right) = 1.5HP$$

Assuming an 80% efficient motor:

$$(1.5\ HP)/(.80) = 1.9\ HP$$

This suggests that a readily available 2 HP motor is appropriate.

GT CALCULATION (RECTANGULAR BASIN):

$$GT = (\overline{G}) * (Detention\ Time)$$

1st compartment:

$$GT = (40\ sec^{-1}) * (10\ min) * \left(60\ \frac{sec}{min}\right) = 24000$$

2nd compartment:

$$GT = (35\ sec^{-1}) * (10\ min) * \left(60\ \frac{sec}{min}\right) = 21000$$

Total GT:

$$24000 + 21000 = 45000$$

GT is greater than the suggested minimum value of 30,000 (see Figure 19).

GT CALCULATION (CIRCULAR BASIN):

$$GT = (40 sec^{-1}) * (20\ min) * \left(60\ \frac{sec}{min}\right) = 48000$$

GT is greater than the suggested minimum value of 30,000 (see Figure 19).

BASIN SIZE CALCULATION (RECTANGULAR BASIN):

To ensure thorough mixing the maximum water depth in the flocculation basins will be 10 feet. It is suggested that the length to width ratio be three[6].

$$(length) * (1/3\ length) = \left(\frac{volume}{depth}\right)$$

$$(L) * (1/3L) = \frac{(8021\ ft^3)}{(10\ ft)}$$

$$L = 49\ feet \qquad use\ 50\ feet$$

$$W = L/3 \qquad W = 49/3 = 16.3\ feet \qquad use\ 17\ feet$$

Actual water depth for a basin 50 feet long and 17 feet wide:

$$Depth = \frac{Volume}{L * W}$$

$$Depth = \frac{8021\ ft^3}{50\ ft * 17\ ft} = 9.4\ feet$$

Use a 50 X 17 X 10 foot basin with a 9.4 foot water depth which allows for a 1.6 foot freeboard.

BASIN SIZE CALCULATION (CIRCULAR BASIN):

To ensure thorough mixing the maximum water depth in the flocculation basin will be 10 feet.

$$Radius=\left(\frac{volume}{\pi * depth}\right)^{0.5}$$

$$Radius=\left(\frac{16042\ ft^3}{\pi * 10\ ft}\right)^{0.5} = 22.6\ ft \quad use\ 25\ ft$$

Actual water depth:

$$Depth=\frac{volume}{\pi * r^2}$$

$$Depth=\frac{16042\ ft^3}{\pi * (25\ ft)^2} = 8.2 ft$$

Use a 25 foot radius basin with a 10 foot depth and an 8.2 water depth which will allow for a freeboard of 1.8 feet.

AREA OF ORIFICES (RECTANGULAR BASIN ONLY):

The recommended velocity through the orifices between compartments is less than 1 ft/sec at maximum flow.

$$Orifice\ area=\frac{flow\ rate}{velocity}$$

$$Orifice\ area=\left(\frac{6000\ gal}{min}\right) * \left(\frac{ft^3}{7.48\ gal}\right) * \left(\frac{1\ min}{60\ sec}\right) \div \left(\frac{0.8\ ft}{sec}\right) = 16.7\ ft^2$$

PERCENT ORIFICE CALCULATION (RECTANGULAR BASIN ONLY):

$$\textit{Area of baffle wall} = (\textit{submerged depth}) * (\textit{length})$$

$$\textit{Area of baffle wall} = (9.4\ ft) * (50\ ft) = 470\ ft^2$$

$$\%\ \textit{Orifice area} = \frac{\textit{orfice area}}{\textit{baffle area}} * 100\%$$

$$\%\ \textit{Orifice area} = \frac{16.7\ ft^2}{470\ ft^2} * 100\% = 3.6\%$$

3.6% falls in the recommended percent orifice ratio of 3%-6%.

MAXIMUM G CALCULATION (RECTANGULAR BASIN):

The maximum value of $\bar{G}$ is calculated using the maximum power input taking into account motor efficiency and the minimum water viscosity.

1st compartment:

$$\bar{G}_{max} = \left(\frac{P_{max}}{\mu_{min} * V}\right)^{0.5}$$

$$P_{max} = (1\ HP*0.8) * \left(\frac{550\ ft*lb/sec}{HP}\right) = 440\frac{ft*lb}{sec}$$

$$\mu_{min} = 2.20*10^{-5}\ \frac{lb*sec}{ft^2}$$

$$\bar{G}_{max} = \left[\frac{440\ ft*lb/sec}{(2.2*10^{-5}\ lb*sec/ft^2) * (8021\ ft^3)}\right]^{0.5} = 49.9\ seconds^{-1}$$

49.9 $seconds^{-1}$ less than the recommended maximum $\bar{G}$ value of 50 $seconds^{-1}$.

2nd compartment:

$$P_{max} = (0.75\ HP*0.8) * \left(\frac{550\ ft*lb/sec}{HP}\right) = 330\frac{ft*lb}{sec}$$

$$\mu_{min} = 2.20*10^{-5}\ \frac{lb*sec}{ft^2}$$

$$\bar{G}_{max} = \left[\frac{330\ ft*lb/sec}{(2.2*10^{-5}\ lb*sec/ft^2)*(8021\ ft^3)}\right]^{0.5} = 43.2\ seconds^{-1}$$

43.2 $seconds^{-1}$ less than the recommended maximum $\bar{G}$ value of 50 $seconds^{-1}$.

MAXIMUM G CALCULATION (CIRCULAR BASIN):

The maximum value of $\bar{G}$ is calculated using the maximum power input taking into account motor efficiency and the minimum water viscosity.

$$P_{max} = (2\ HP*0.8) * \left(\frac{550\ ft*lb/sec}{HP}\right) = 880\frac{ft*lb}{sec}$$

$$\bar{G}_{max} = \left[\frac{880\ ft*lb/sec}{(2.2*10^{-5}\ lb*sec/ft^2)*(16042\ ft^3)}\right]^{0.5} = 49.9\ seconds^{-1}$$

49.9 $seconds^{-1}$ less than the recommended maximum $\bar{G}$ value of 50 $seconds^{-1}$.

The determination on whether to use a rectangular or circular basin will be made in a later section of this thesis. That determination will be made when considering the cost and efficiency of the two different basin configurations.

SEDIMENTATION

Sedimentation is the process of liquid-solid separation. The desired outcome of sedimentation is to have a water that is low in solids sent to the filtration process to minimize the amount of solids that the filters must remove.

Due to the complexity of the physical mechanisms affecting ideal sedimentation, accurate theoretical modeling of an actual sedimentation basin is difficult. Factors such as wind currents, temperature gradients, and inlet and outlet currents all affect ideal sedimentation. Floc particles are not all the same size, shape, or density, so the actual settling velocity cannot accurately be predicted.

For sedimentation the best approach toward designing a full-scale sedimentation basin is to use generally acceptable engineering design parameters.

Sedimentation Mechanisms

Sedimentation processes are classified into four regimes:

- Settling of nonflocculated particles
- Settling of flocculated particles
- Zone or hindered settling
- Compression settling

Settling of Nonflocculated Particles:

Settling of nonflocculated particles represents ideal theoretical settling. Ideal settling theory assumes that all particles settle discretely and are uniformly distributed throughout the vertical plane of the settling basin. This type of sedimentation is predominant when the shape of the particle is essentially spherical.

The settling velocity of a particle starting at rest will accelerate until the resistance to flow through the liquid is equal to the effective weight of the particle. The effective weight of the particle is the weight of the particle minus its buoyancy force. After acceleration ceases the particle reaches its terminal settling velocity. The terminal settling velocity is a function of the properties of the particle and the liquid in which it is settling. The terminal settling velocity of a sphere in a liquid is defined by the following equation:

$$V_t = \left[\frac{4*g*(\rho_s-\rho)*d}{3*C_d*\rho}\right]^{0.5} \tag{12}$$

Where: g= gravity constant (m/sec^2)
ρ_s= density of the settling particle (kg/m^3)
ρ= density of the water (kg/m^3)
C_d= coefficient of drag

Equation 12 is derived by summing the forces acting on a settling sphere. As this type of settling does not occur in water treatment, it will not be discussed any further.

Settling of Flocculated Particles:

As sedimentation in a water treatment plant does not follow ideal settling theory, settling of flocculated particles is the predominant settling mechanism in water treatment.

Settling of flocculated particles recognizes the following:

1. Particles do not settle discretely.
2. Differential settling flocculation will occur, resulting in an increase in settling particle size and thus settling velocity.
3. Inlet and outlet conditions will affect settling.
4. Thermal and basin currents will affect settling.

Equation 12 is still the governing equation for the velocity of a settling particle, with a modification to the coefficient of drag to account for the shape of the floc particle.

In order to account for the above listed non-ideal conditions engineering design parameters such as flow, basin depth, basin surface area, and the properties of the settling particle are utilized to obtain a reasonably designed sedimentation basin. These parameters will be examined, in detail, in a later part of this section.

Zone or Hindered Settling:

As a particle settles it displaces the liquid in which it is settling, resulting in an upward flow of the liquid. When the concentration of the settling particles is small this upward velocity is small and can be ignored. When the concentration of the settling particles becomes large enough the upward velocity of the displaced water begins to approach the velocity of the settling particles. Since the settling velocity is relative to the movement of the water, the upward velocity of the water reduces the downward velocity of the settling particle. The net effect is that the time necessary for settling is increased. To compensate for this the hindered settling velocity, as opposed to the terminal settling velocity, is used to design the sedimentation basin. For water treatment plants the effects of hindered settling are negligible and can be ignored as, the influent solids loading is relatively low[16].

Compression Settling:

Compression settling is the settling that occurs under the layer of hindered settling particles and is caused by the weight of the solids above. The effects of compression settling are minor and will be ignored.

Theory of Settling Flocculated Particles

Terminal Settling Velocity:

The terminal settling velocity of a sphere in a liquid is defined by Equation 12. The coefficient of drag (C_d) is a function of the Reynolds number (R_e) which is calculated as follows:

$$R_e = \frac{\rho * V_t * d}{\mu} \qquad (13)$$

Where: Re= Reynolds number (unitless)
ρ= density of the liquid (kg/m^3)
d= diameter of the sphere (m)
V_t= terminal settling velocity (m/sec)
μ= dynamic viscosity (N*sec/m^3)

When the Reynolds number is less than 500, the coefficient of drag is[17]:

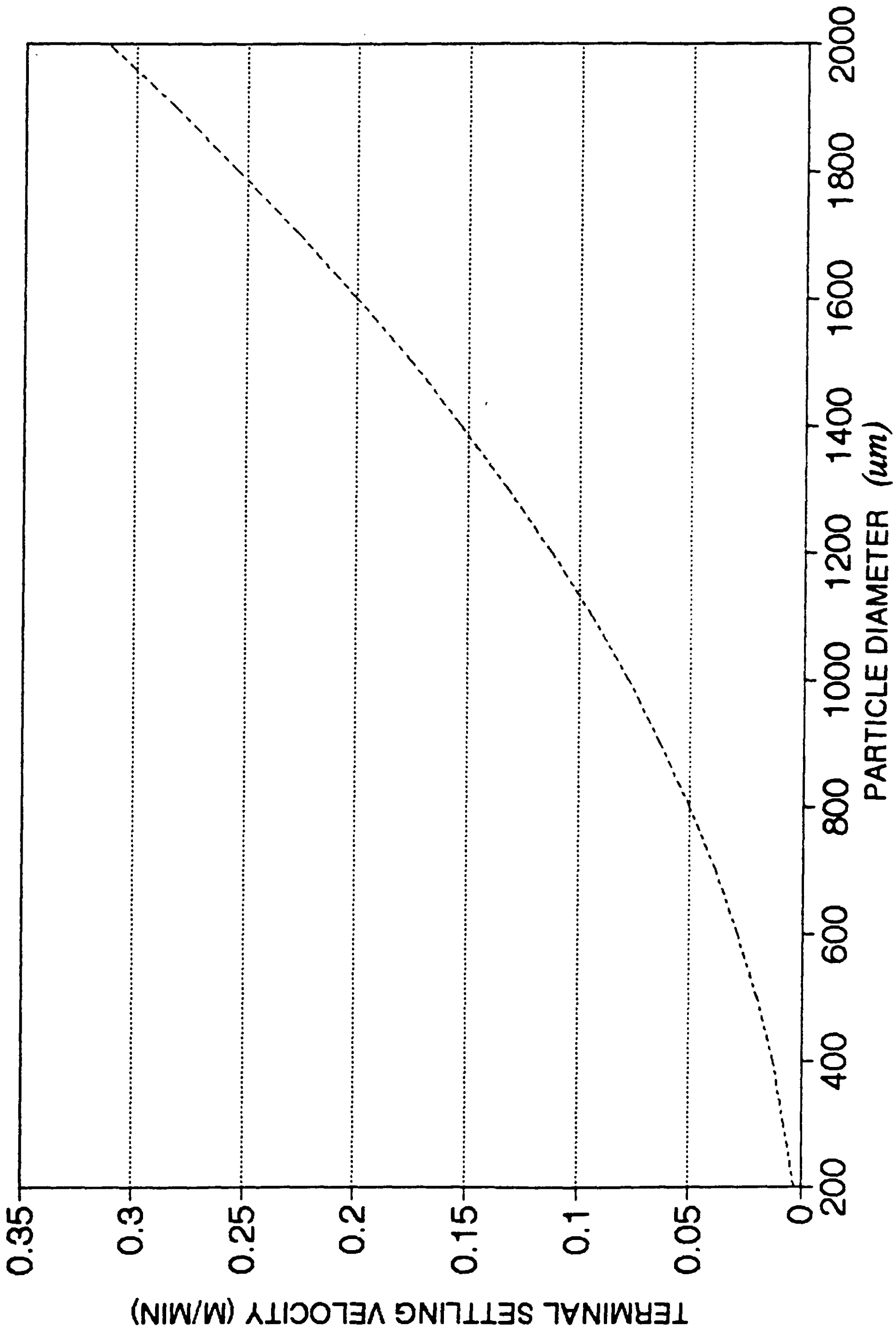
$$C_d = \frac{24 * \phi}{R_e} \qquad (14)$$

Where: 24= empirical constant
Φ= shape factor
R_e= Reynolds number

Because of the varying size and shape of floc particles, the shape factor is difficult to determine but it is generally accepted[18] to be 22. Therefore the terminal settling velocity for floc particles settling in water is:

$$V_t = \frac{g * (\rho_s - \rho) * d^2}{396 * \mu} \tag{15}$$

Figure 21 shows how the terminal settling velocity varies with the particle diameter.



Terminal Settling Velocity
VS Particle Diameter

Figure 21

Transitional Settling Velocity:

Transitional settling velocity is the velocity of a settling particle that is accelerating until it reaches the terminal settling velocity. For any given transitional settling distance x, where the particle is still accelerating, the square of the transitional velocity is defined as[19]:

$$V^2 = \frac{8*g*r*(\rho_s-\rho_w)}{3*C_d*\rho_w} * \left[1-\exp\left(\frac{-3*\rho_w*C_d*x}{4*\rho_s*r}\right)\right] \tag{16}$$

Where: x= transitional settling distance (m)
r= particle radius (m)

Table 1 shows the relationship between the size of a settling particle and the distance it takes to reach ninety percent of the terminal settling velocity. Since a settling particle asymptotically approaches the terminal settling velocity the distance to reach ninety percent of the terminal settling velocity was calculated[20]. Table 1 shows that the terminal settling velocity is rapidly reached as the distance the particle must fall is minute.

Table 1

Transitional Settling Velocity

PARTICLE RADIUS (M^-6)	SETTLING VELOCITY (M/SEC)	90% OF Vt^2 (M/SEC)	SETTLING DISTANCE TO REACH 90% Vt (M)
100	0.000052	2.20E-09	3.11E-09
200	0.000209	3.53E-08	4.97E-08
300	0.000469	1.78E-07	2.52E-07
400	0.000834	5.64E-07	7.96E-07
500	0.001304	0.000001	0.000002
600	0.001878	0.000003	0.000004
700	0.002556	0.000005	0.000007
800	0.003338	0.000009	0.000013
900	0.004224	0.000014	0.00002
1000	0.005215	0.000022	0.000031
1100	0.006311	0.000032	0.000046
1200	0.00751	0.000046	0.000064
1300	0.008814	0.000063	0.000089
1400	0.010222	0.000085	0.000119
1500	0.011734	0.000112	0.000157
1600	0.013351	0.000144	0.000204
1700	0.015072	0.000184	0.00026
1800	0.016898	0.000231	0.000326
1900	0.018827	0.000287	0.000405
2000	0.020861	0.000353	0.000497

Surface Loading Rate:

In a continuous horizontal flow rectangular tank a settling particle has both horizontal and vertical velocity components as is shown in Figure 22.

The length that a settling particle will travel across a rectangular basin is defined as follows:

$$l = \frac{t * Q}{H * W} \tag{17}$$

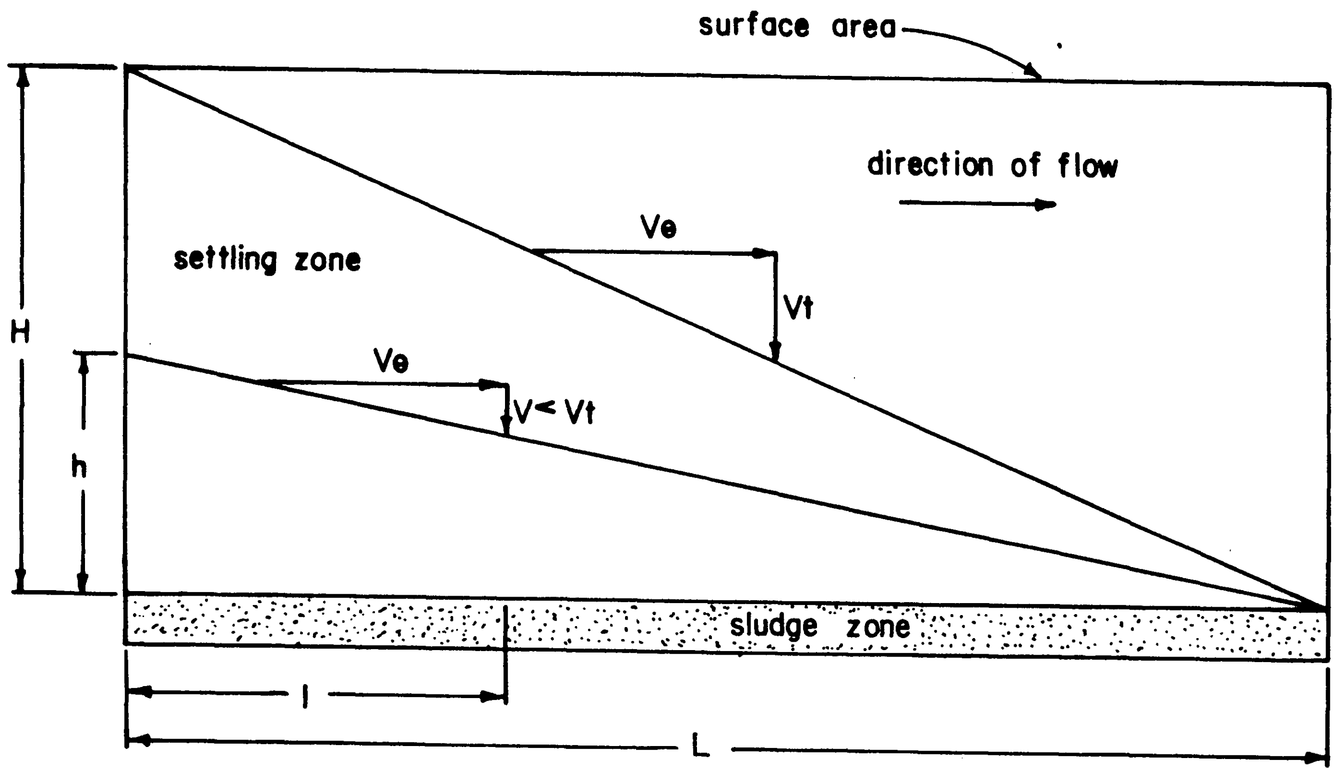
Where: (see Figure 22)
- l= horizontal distance traveled (ft, m)
- t= time of travel (sec)
- H= depth of sedimentation basin (ft, m)
- W= width of sedimentation basin (ft, m)
- Q= flow (ft^3/sec, m^3/sec)

The vertical distance (h) settled is:

$$h = V_t * t \tag{18}$$

The time for a particle to settle a distance h therefore would be:

$$t = \frac{h}{V_t} \tag{19}$$



Ideal Sedimentation Basin

Figure 22

Substituting Equation 19 into 17 yields:

$$l = \frac{h * Q}{V_t * H * W} \tag{20}$$

or:

$$V_t = \frac{h * Q}{H * l * W} \tag{21}$$

If all of the particles entering the basin, with a given V_t, were to settle out then l=L and h=H (see Figure 22) and Equation 21 becomes:

$$V_t = \frac{Q}{L * W} \tag{22}$$

Where: L*W= sedimentation basin surface area (ft^2, m^2)

This defines an important sedimentation basin design parameter called the surface loading rate which is as follows:

$$V_t^* = \frac{Q}{L * W} = \frac{Q}{A^*} \tag{23}$$

Where: V_t^*= surface loading rate (GPM/ft^2, $m^3/min/m^2$)
A^*= sedimentation basin surface area (ft^2, m^2)

Particles with a settling velocity less than V_t^* would not settle out and would require a larger basin to allow for the necessary extra settling time.

Equation 23 shows that the settling efficiency is dependent on the sedimentation basin surface area and independent of the depth or flow area. The preceding statement is known as Hazen's Law[7] and is true in theory, but in practice the depth of a sedimentation is important. The preceding derivation was for a rectangular basin but the same principles hold true for a circular basin.

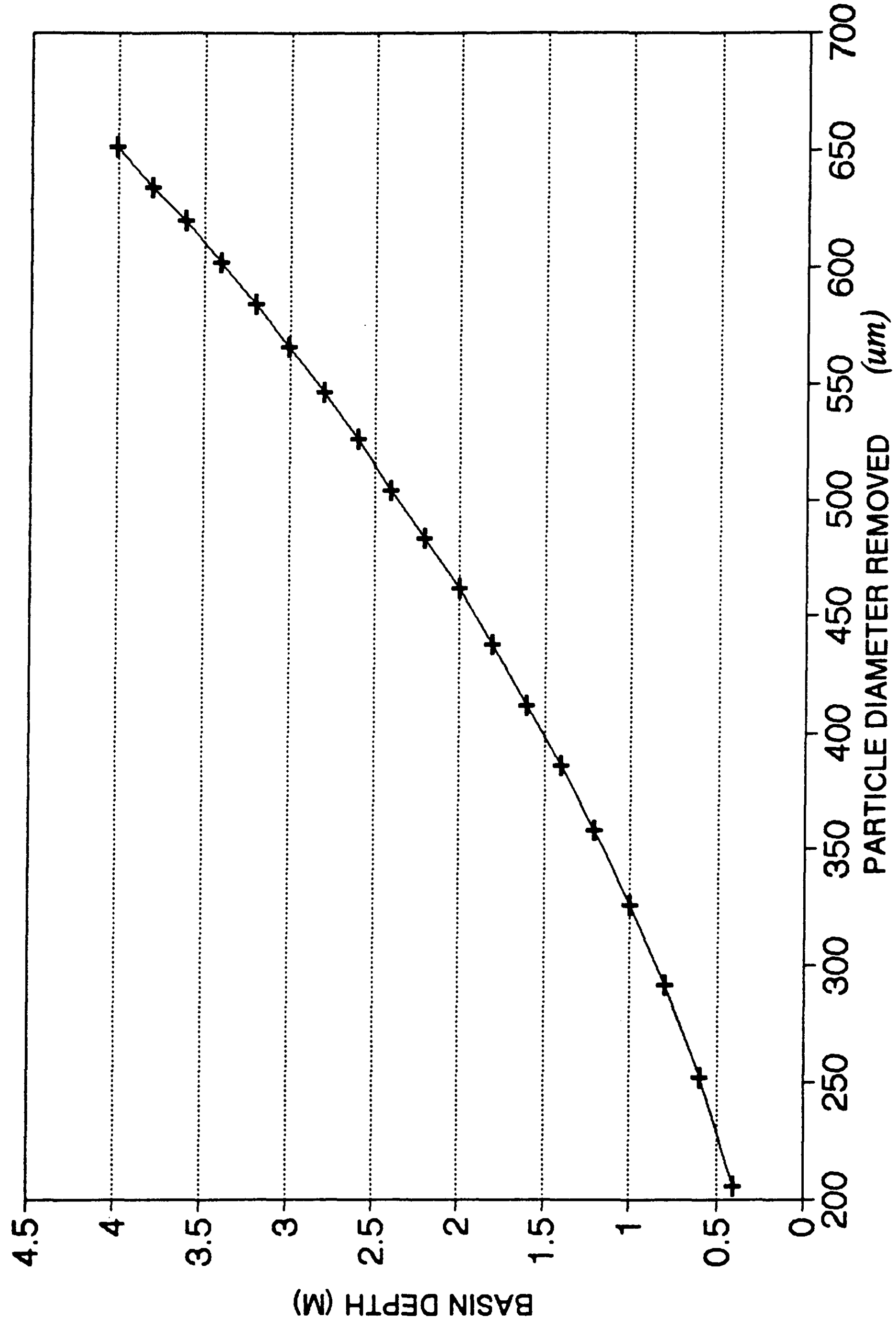
Theoretical Basin Configuration

Several spreadsheets were developed varying the geometrical configuration of the sedimentation basin to see if an optimal configuration existed. An optimum basin would have the minimum basin area occurring at the highest removal efficiency. The basin area is defined as the submerged surface area including the walls and floor. The reason behind minimizing the basin area is that fewer materials are necessary to construct the basin, thus reducing the cost.

The removal efficiency is a function of the size of the particle that the basin is capable of removing. The smaller the particle removed the more efficient the basin. The assumption made here is that all particles larger than the smallest one removed will also be removed.

In order to compare different basin configurations, the volume of the basin was fixed as well as the flow through the basin. In a later section of this thesis the volume of the basin will be varied, and a relationship between efficiency and cost will be developed. Fixing the volume and flow also fixes the detention time through the basin. The detention time is defined as the basin volume divided by the flow to the basin. If the detention time through a basin is fixed then, the time that a particle has to settle is also fixed.

As shown in Figure 21 a smaller diameter particle settles at a slower rate. If the time given for different sized particles to settle out is fixed, a smaller particle would have to settle a shorter distance meaning the basin would have to be shallower. The removal efficiency of a sedimentation basin is a function of the size of particles that it can remove; therefore theoretically, the removal efficiency of a sedimentation basin is a function of its depth. This is true in theory, but a shallow basin would be more susceptible to wind and thermal currents that would hinder settling. In order to compare the efficiency of basins with different configurations the depth will be used as an indication of its efficiency. Figure 23 shows the relationship between the depth of a basin and the size of particle removed.



Sedimentation basin-Basin Depth
VS Particle Diameter Removed

Figure 23

In Table 2 the length and volume are fixed and the width and depth are allowed to vary. In Table 3 the width and volume are fixed and the length and depth are allowed to vary.

Keeping in mind that the efficiency of the basin decreases with an increasing depth, and that the cost is based on the basin area, Tables 2 & 3 show that the minimum basin cost occurs at the minimum rather than the maximum basin efficiency.

In Table 4 the volume and depth are fixed and the width and length are allowed to vary. Since the basin depth is fixed, the efficiency of the basin is constant with the varying widths and lengths. Table 4 shows that the cost of the basin is minimized when the length-to-width ratio is one. The generally acceptable length-to-width ratio[21] is in the range of 3:1 to 5:1. The reason for the above length-to-width ratio is to minimize cross currents and to maintain plug flow as best as possible.

In Table 5 a circular basin is modeled fixing the basin volume and allowing the depth and radius to vary.

Table 2

Rectangular Basin-Length and Volume Fixed

BASIN PARAMETERS

BASIN LENGTH (M)	61
BASIN VOL (M^3)	2745
FLOW (M^3/MIN)	22.71

WIDTH (M)	DEPTH (M)	LENGTH TO WIDTH RATIO	FLOW AREA (M^2)	HORZ VELOCITY (M/MIN)	SURFACE AREA (M^2)	BASIN AREA (M^2)
10	4.50	6.10	45	0.50	610	1249
11	4.09	5.55	45	0.50	671	1260
12	3.75	5.08	45	0.50	732	1279
13	3.46	4.69	45	0.50	793	1305
14	3.21	4.36	45	0.50	854	1334
15	3.00	4.07	45	0.50	915	1371
16	2.81	3.81	45	0.50	976	1403
17	2.65	3.59	45	0.50	1037	1449
18	2.50	3.39	45	0.50	1098	1493
19	2.37	3.21	45	0.50	1159	1537
20	2.25	3.05	45	0.50	1220	1584
21	2.14	2.90	45	0.50	1281	1632
22	2.05	2.77	45	0.50	1342	1681
23	1.96	2.65	45	0.50	1403	1731
24	1.88	2.54	45	0.50	1464	1782
25	1.80	2.44	45	0.50	1525	1834
26	1.73	2.35	45	0.50	1586	1887
27	1.67	2.26	45	0.50	1647	1940
28	1.61	2.18	45	0.50	1708	1994
29	1.55	2.10	45	0.50	1769	2048
30	1.50	2.03	45	0.50	1830	2103

Table 3

Rectangular Basin-Width and Volume Fixed

BASIN PARAMETERS
BASIN WIDTH (M) 15
BASIN VOL (M^3) 2745
FLOW (M^3/MIN) 22.71

DEPTH (M)	LENGTH (M)	LENGTH TO WIDTH RATIO	FLOW AREA (M^2)	HORZ VELOCITY (M/MIN)	SURFACE AREA (M^2)	BASIN AREA (M^2)
0.40	457	30.50	6	3.79	6860	7240
0.60	305	20.33	9	2.52	4570	4950
0.80	228	15.25	12	1.89	3435	3825
1.00	183	12.20	15	1.51	2740	3140
1.20	152	10.17	18	1.26	2280	2680
1.40	130	8.71	21	1.08	1961	2361
1.60	114	7.63	24	0.95	1713	2123
1.80	101	6.78	27	0.84	1520	1940
2.00	91	6.10	30	0.76	1370	1790
2.20	83	5.55	33	0.69	1243	1673
2.40	76	5.08	36	0.63	1145	1585
2.60	70	4.69	39	0.58	1057	1497
2.80	65	4.36	42	0.54	986	1436
3.00	61	4.07	45	0.50	910	1370
3.20	57	3.81	48	0.47	851	1311
3.40	53	3.59	51	0.45	805	1275
3.60	50	3.39	54	0.42	760	1230
3.80	48	3.21	57	0.40	727	1207
4.00	45	3.05	60	0.38	685	1175
4.20	43	2.90	63	0.36	657	1147

Table 4

Rectangular Basin-Fixed Depth and Volume

BASIN PARAMETERS
SETTLING DEPTH (M) 3
BASIN VOL (M^3) 2745
FLOW (M^3/MIN) 22.71

WIDTH (M)	LENGTH (M)	LENGTH TO WIDTH RATIO	FLOW AREA (M^2)	HORZ VELOCITY (M/MIN)	SURFACE AREA (M^2)	BASIN AREA (M^2)
10	91.50	9.15	30	0.76	915	1524
11	83.18	7.56	33	0.69	915	1489
12	76.25	6.35	36	0.63	915	1440
13	70.38	5.41	39	0.58	915	1411
14	65.36	4.67	42	0.54	915	1394
15	61	4.07	45	0.50	915	1371
16	57.19	3.57	48	0.47	915	1353
17	53.82	3.17	51	0.45	915	1334
18	50.83	2.82	54	0.42	915	1328
19	48.16	2.53	57	0.40	915	1315
20	45.75	2.29	60	0.38	915	1300
21	43.57	2.07	63	0.36	915	1303
22	41.59	1.89	66	0.34	915	1295
23	39.78	1.73	69	0.33	915	1290
24	38.13	1.59	72	0.32	915	1285
25	36.60	1.46	75	0.30	915	1280
26	35.19	1.35	70	0.29	915	1285
27	33.89	1.26	80	0.28	915	1283
28	32.68	1.17	80	0.27	915	1277
29	31.55	1.09	80	0.26	915	1271
30	30.50	1.02	90	0.25	915	1270

Table 5

Circular Basin-Fixed Volume

BASIN PARAMETERS				
BASIN VOLUME (M^3)	2745			
FLOW (M^3/MIN)	22.71			
INSIDE RADIUS (M)	7.62			
DEPTH (M)	OUTSIDE RADIUS (M)	AVG HORZ VELOCITY (M/MIN)	SURFACE AREA (M^2)	BASIN AREA (M^2)
0.40	46.74	0.33	6680	6979
0.60	38.16	0.25	4392	4718
0.80	33.05	0.21	3248	3597
1.00	29.56	0.18	2562	2930
1.20	26.98	0.16	2105	2490
1.40	24.98	0.14	1778	2180
1.60	23.37	0.13	1533	1950
1.80	22.03	0.12	1342	1774
2.00	20.90	0.11	1190	1635
2.20	19.93	0.10	1065	1523
2.40	19.08	0.10	961	1431
2.60	18.33	0.09	873	1355
2.80	17.67	0.08	797	1291
3.00	17.07	0.08	732	1236
3.20	16.52	0.07	675	1190
3.40	16.03	0.07	624	1149
3.60	15.58	0.07	580	1114
3.80	15.16	0.06	539	1084
4.00	14.78	0.06	503	1057
4.20	14.42	0.06	471	1034
4.40	14.09	0.05	441	1013
4.60	13.78	0.05	414	995
4.80	13.49	0.05	389	978
5.00	13.22	0.05	366	964

As can be seen in Table 5 the most efficient basin also has the highest basin area or cost.

The preceding four tables show that an optimal basin with maximum efficiency occurring at the minimum cost does not exist. Therefore, a trade-off between efficiency and cost will have to be made. This trade-off between efficiency and cost will be examined in a future section of this thesis

Types of Sedimentation Basins

There are several types of sedimentation basins commonly in use in water treatment:

- Conventional rectangular or circular basin
- Upflow clarifier
- Conventional basin with tube settlers

Conventional Basin:

A conventional sedimentation basin is a concrete basin that allows water to flow horizontally through it. The basin geometry is either rectangular or circular. As the water flows through the basin, the floc particles are allowed to settle out with the settling velocity as calculated in Equation 15. For a rectangular basin the water flows over a weir into the basin to assure an even distribution of water into the basin. In a circular basin the water is fed in the center of the basin and below the water surface.

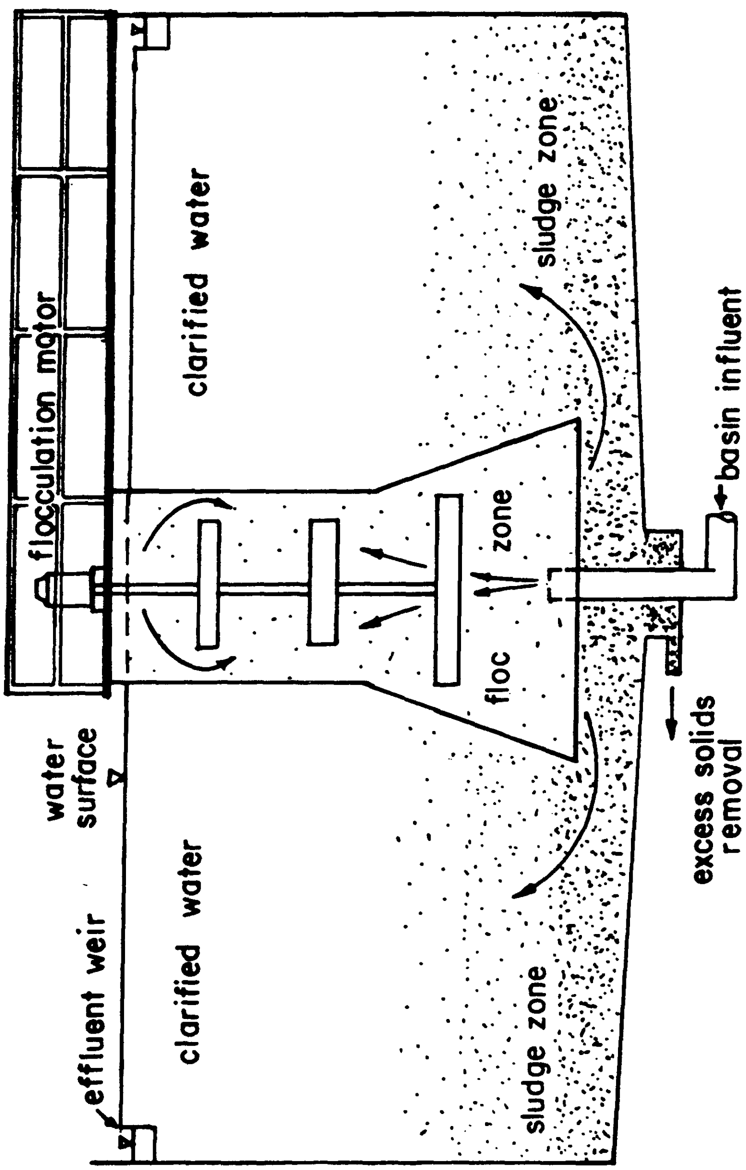
<u>Upflow Clarifiers:</u>

The terminal settling velocity for an upflow clarifier is dictated by Equation 15, except that the actual settling velocity is relative to the flow of water which is in the opposite direction of the settling particle. The equation used to calculate the settling velocity in an upflow clarifier is as follows:

$$V_t = \frac{g * (\rho_s - \rho) * d^2}{396 * \mu} - V\,clarifier \qquad (24)$$

Where: V clarifier= upward velocity through the upflow clarifier (m/min)
All other parameters as in Equation 15

The influent feed to an upflow clarifier is at the bottom and the water is allowed to flow up through the clarifier (see Figure 24). To facilitate settling of floc particles, the upward flow through the clarifier is less than the velocity of the settling particles. The idea behind an upflow clarifier is that a blanket of solids will accumulate toward the bottom of the clarifier and will act to filter or trap the incoming floc particles. This type of clarifier is more effective with water that is high in turbidity, as a higher turbidity water will form the blanket



Upflow Clarifier

Figure 24

of solids quicker. An upflow clarifier also needs to be continually operated to prevent the solids blanket from settling out.

Tube Settlers:

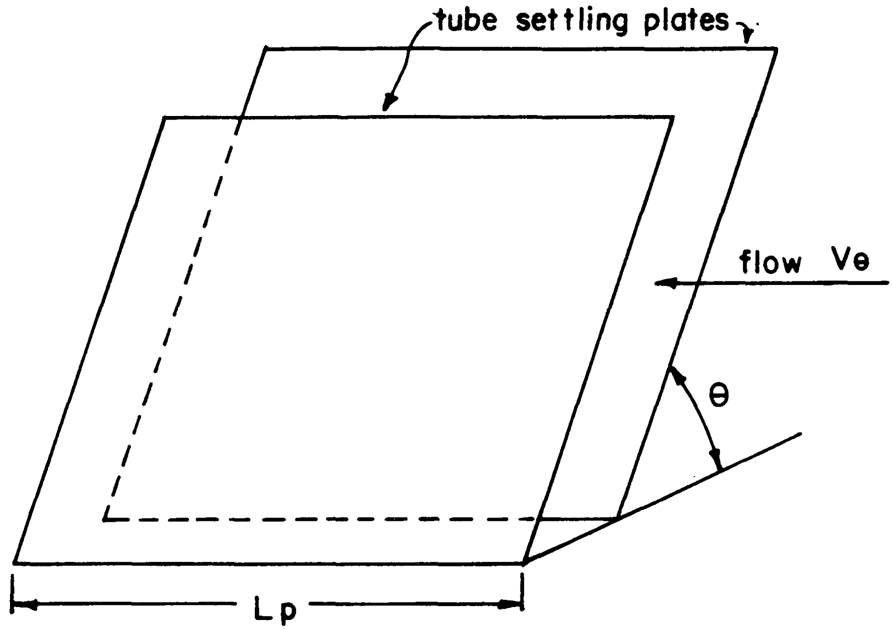
As previously explained, the settling efficiency of particles in a liquid is dependent on the area available for settling. Tube settlers take advantage of this fact by using a series of closely spaced inclined plates (see Figures 25 and 26).

Yao developed the following equations for calculating the length of surface necessary to allow a particle enough time to settle the vertical distance between plates[22]:

$$t= \frac{w}{V_t * \cos\theta} \tag{25}$$

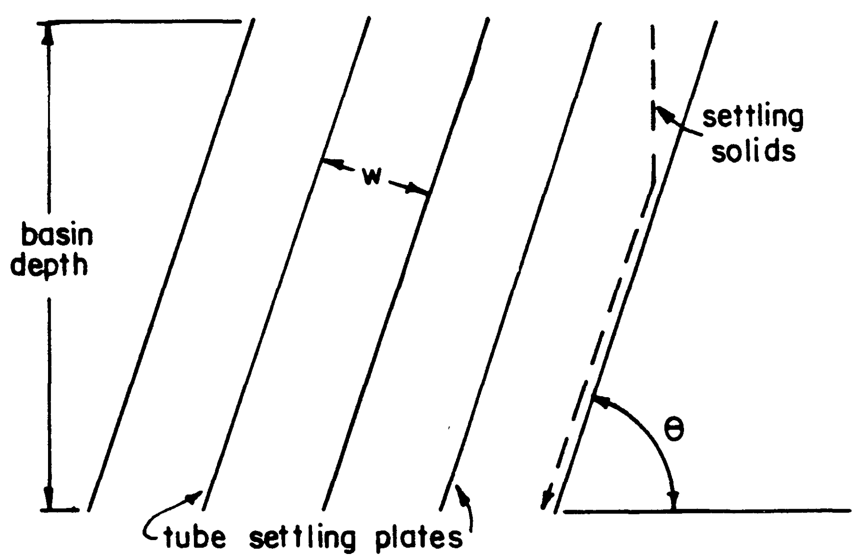
$$Lp= \frac{V_\theta * w}{V_t * \cos\theta} \tag{26}$$

Where: t= time (seconds)
w= perpendicular distance between plates (ft, m)
V_t= terminal settling velocity (ft/min, m/min)
Lp= length of plate (ft, m)
V_θ= velocity through clarifier (ft/min, m/min)
θ= angle of tube settlers (see Figure 25)



Tube Settler Schematic

Figure 25



Tube Settler Schematic

Figure 26

Sedimentation Basin Design

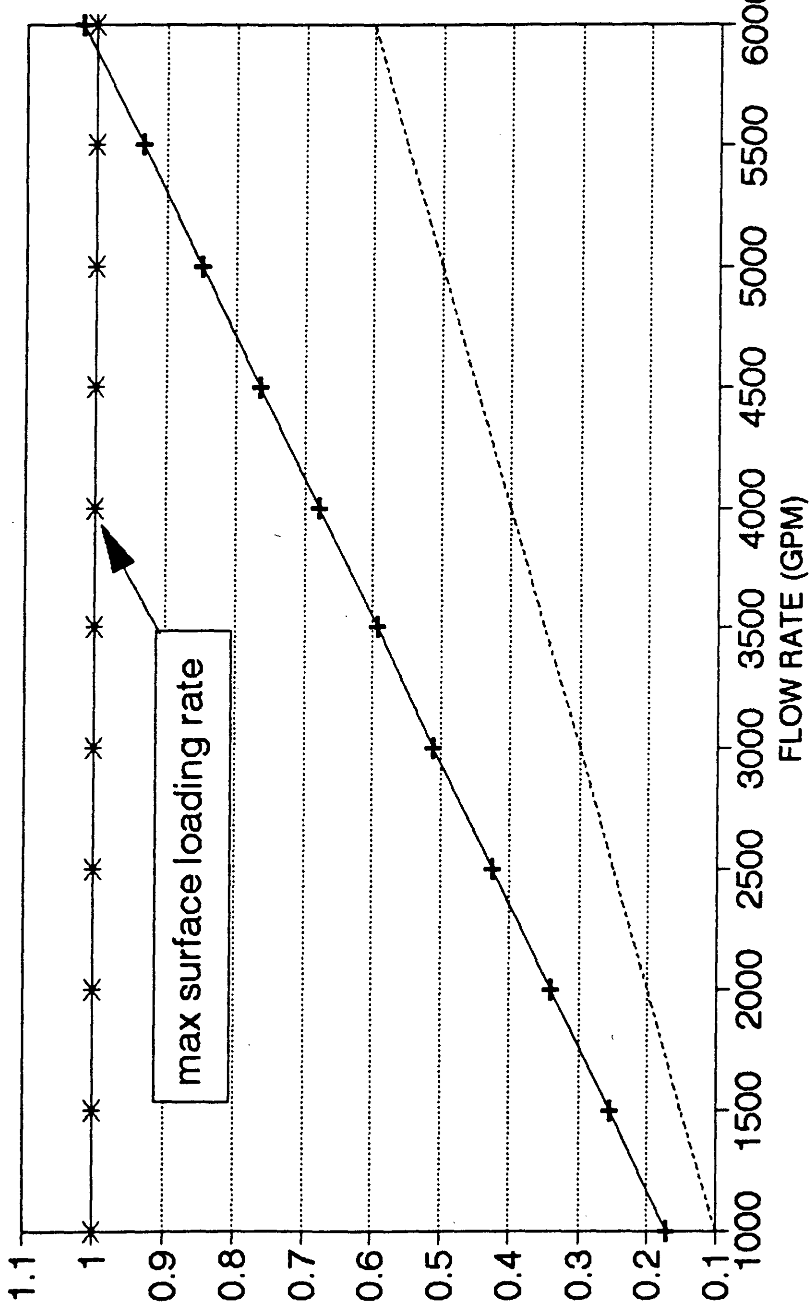
The sedimentation basin that is most appropriate for the NTMWTP is either a rectangular or circular conventional settling basin for the following reasons:

1. The plant influent turbidity is usually low; as previously stated an upflow clarifier functions better with a high influent turbidity.
2. The NTMWTP does not operate 24 hours per day; an upflow clarifier needs to operate continuously.
3. A plant scale test utilizing tube settlers at the NTMWTP proved ineffective.

The following sedimentation basin design is based on recommended design parameters. For the NTMWTP the following design parameters will be used to size a sedimentation basin:

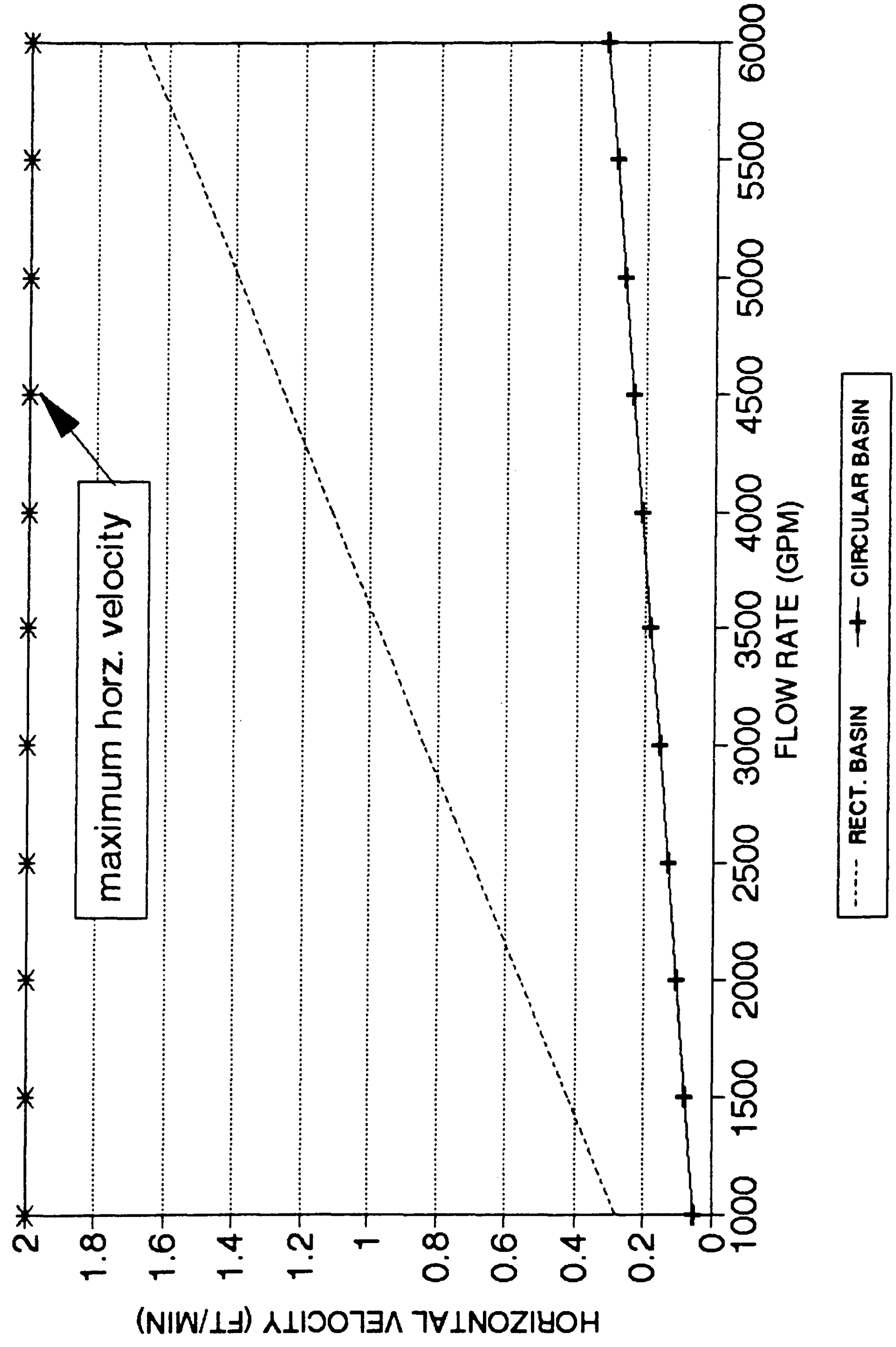
Parameter	Value
Surface loading	0.4 to 1.0 GPM/ft^2
Detention time (Dt)	2-4 hours
Max. horizontal velocity	0.5 to 2 ft/min
Max. length (rect. basin)	200 ft
Depth (rect. basin)	10-15 ft
Length to width ratio	3:1 to 5:1
Max. diameter	100 ft
Depth (circular basin)	15-18 ft
Weir overflow rate	10-20 GPM/ft

Values above the recommended range for surface loading, horizontal velocity, and weir overflow rate would have a detrimental effect on sedimentation. Values for these parameters below the recommended range would not effect sedimentation but would show that the basin was over designed (see Figures 27-29). For detention time the opposite is true and a detention time less than two hours at maximum flow would not be acceptable (see Figure 30).



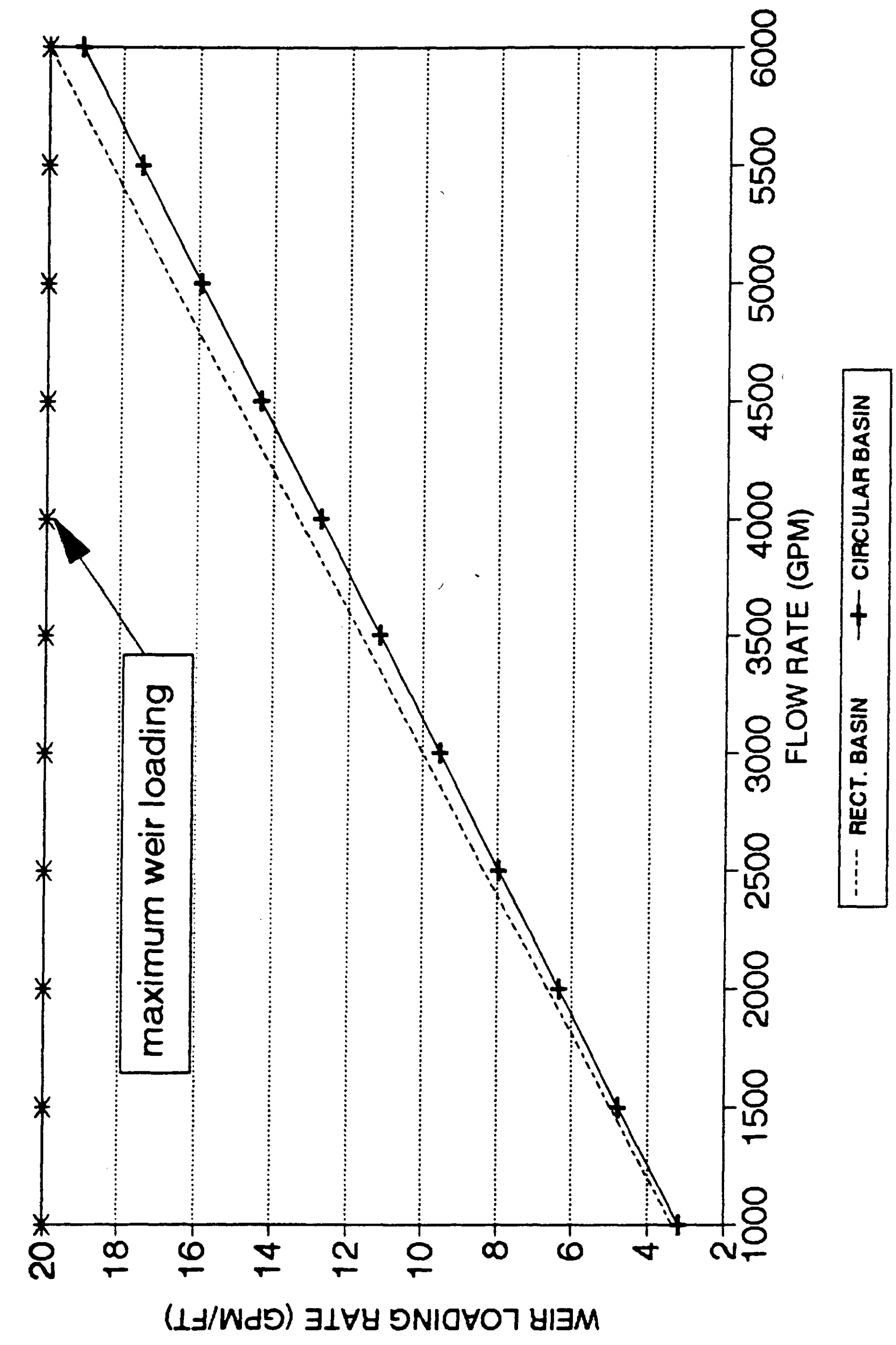
Sedimentation Basin
Surface Loading Rate VS Flow

Figure 27



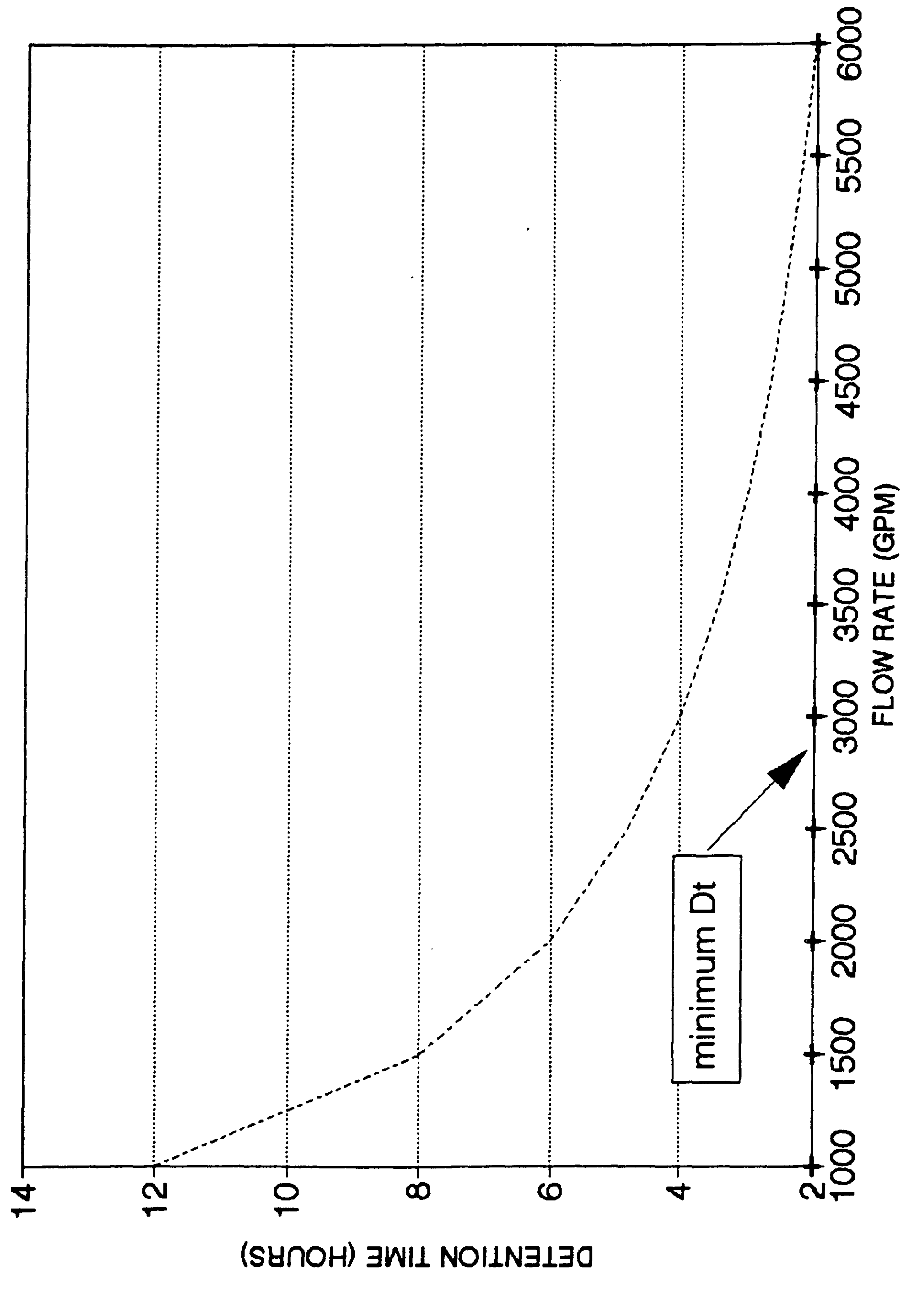
Sedimentation Basin
Max. Horizontal Velocity VS Flow

Figure 28



Sedimentation Basin
Weir Loading Rate VS Flow

Figure 29



Sedimentation Basin
Detention Time VS Flow

Figure 30

SEDIMENTATION BASIN VOLUME CALCULATION:

A detention time of two hours will be used, as this is the minimum acceptable detention time at the maximum design flow of 6000 GPM.

The following calculation is the same for either a rectangular or circular basin:

$$Volume = flow * detention\ time\ (Dt)$$

$$Volume = \left(\frac{6000\ gal}{\min}\right) * \left(\frac{60\ \min}{1\ hr}\right) * (2\ hr) * \left(\frac{1\ ft^3}{7.48\ gal}\right) = 96257\ ft^3$$

BASIN SIZE CALCULATION (RECTANGULAR BASIN):

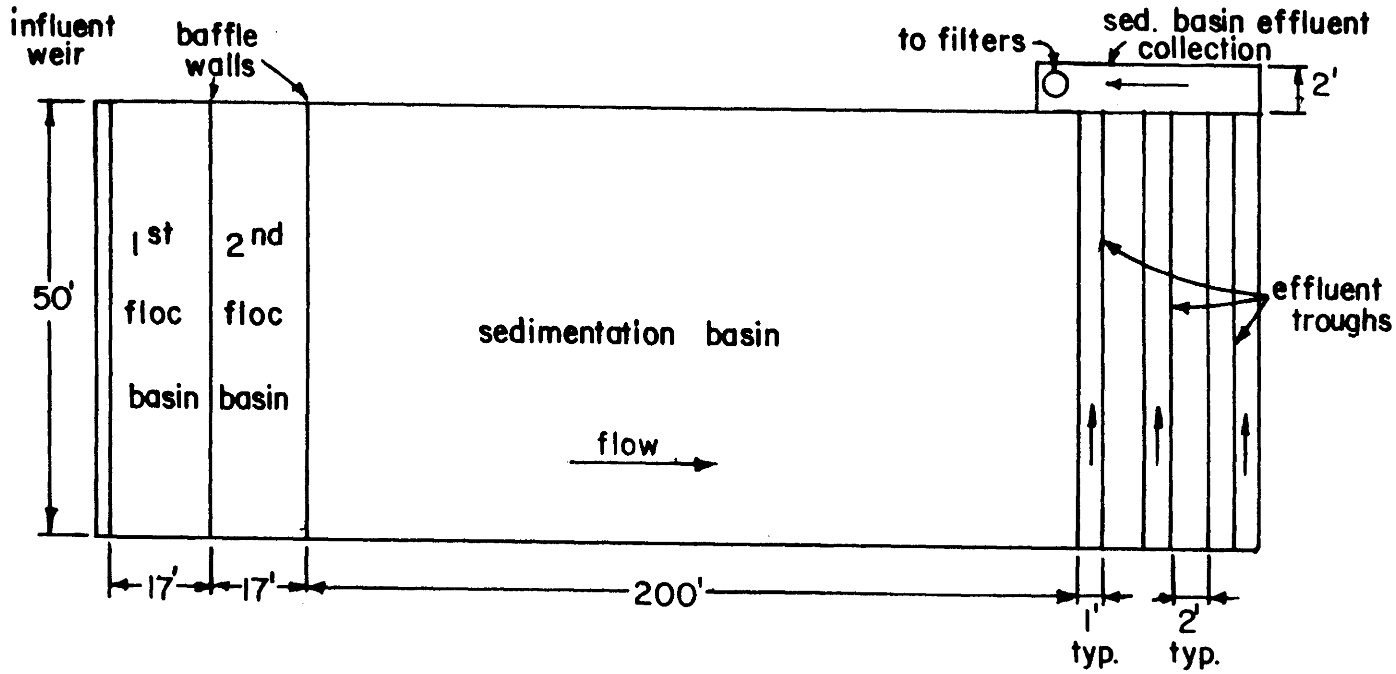
As the flow from the flocculation basin flows directly into the sedimentation basin, the width of the two basins will be the same (see Figure 31 & 32). As previously calculated in the section on flocculation the width of the flocculation basin was found to be 50 feet.

As the volume calculation was based on the maximum flow, the maximum acceptable length of 200 feet will be used in conjunction with a width of 50 feet. This length-to-width ratio of 4:1 ratio also falls in the acceptable range of 3:1 to 5:1.

$$Depth = \frac{volume}{length * width}$$

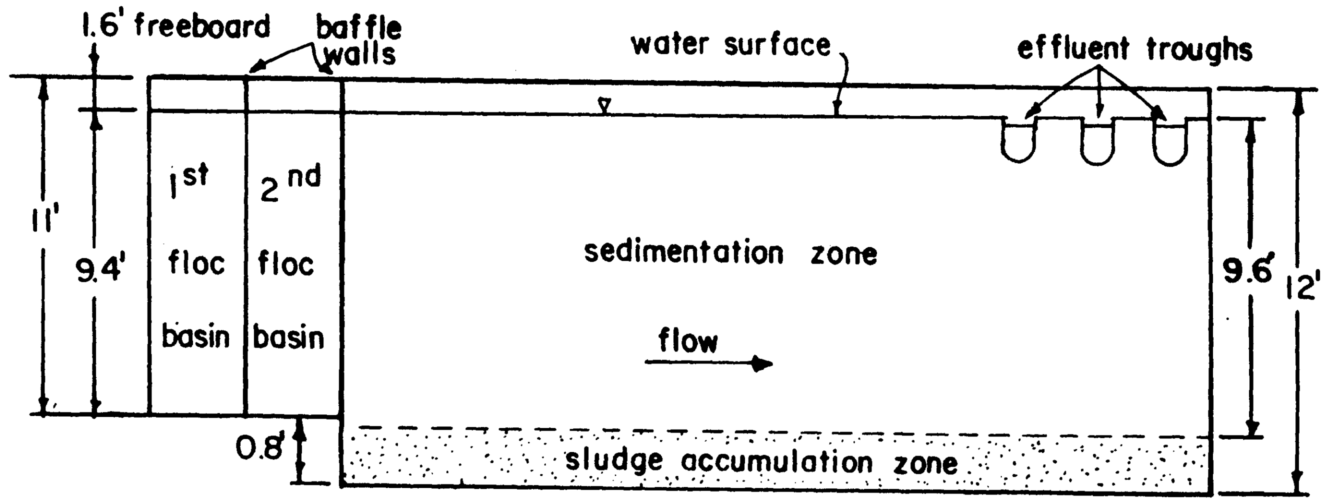
$$Depth = \frac{96257\ ft^3}{50\ ft * 200\ ft} = 9.6\ ft$$

The depth of the flocculation basin was calculated in the section on flocculation to be 9.4 feet, which makes the depth of the sedimentation basin slightly deeper. This will not pose a problem, as the actual depth of the sedimentation basin will be 12 feet to allow for a 1.6 foot freeboard and 0.8 feet of accumulated settled solids (see Figures 31 & 32). The accumulated solids will be removed using a manufactured vacuum system or mechanical scrapers will move the sludge to a sump where it will be pumped out of the basin.



Rectangular Sedimentation Basin
Design (Plan View)

Figure 31



Rectangular Sedimentation Basin
Design (Elevation View)

Figure 32

BASIN SIZE CALCULATION (CIRCULAR BASIN):

As the volume calculation was based on the maximum flow, the maximum acceptable diameter of 100 feet will be used (see Figure 33). In a circular basin the flocculation area cannot be used for the sedimentation area and must be subtracted out. The radius of the flocculation basin was previously calculated to be 25 feet.

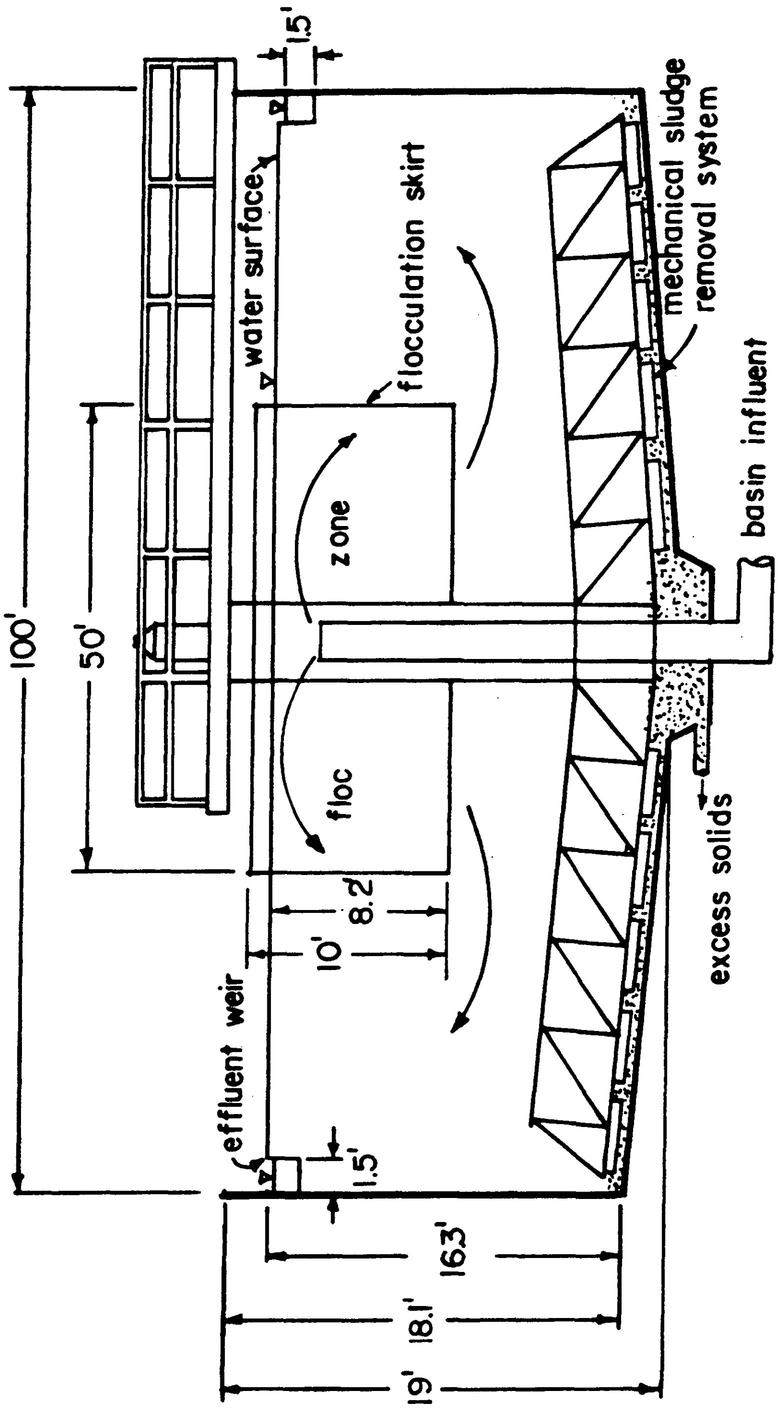
$$Sedimentation\ Area = total\ area - flocculation\ area$$

$$Area = [(50\ ft)^2 * \pi] - [(25\ ft)^2 * \pi] = 5890.5\ ft^2$$

$$Depth = \frac{volume}{area}$$

$$Depth = \frac{96257\ ft^3}{5890.5\ ft^2} = 16.3\ ft$$

The actual depth of the basin will be 19 feet to allow for a 1.8 foot freeboard and 0.9 feet of accumulated solids (see Figure 33). Accumulated solids will be scraped toward the center of the basin and pumped out of the basin (see Figure 33).



Circular Sedimentation Basin
Design (Elevation View)

Figure 33

SURFACE LOADING RATE CALCULATION (RECTANGULAR BASIN):

$$\textit{Maximum Surface Loading Rate} = \frac{\textit{maximum flow}}{\textit{surface area}}$$

$$\textit{Surface Loading} = \left(\frac{6000\ gal}{\min}\right) \div (50\ ft * 200\ ft) = 0.6\ GPM/ft^2$$

The calculated surface loading rate of 0.6 GPM/ft^2 falls within the acceptable range of 0.4 to 1.0 GPM/ft^2.

SURFACE LOADING RATE CALCULATION (CIRCULAR BASIN):

$$\textit{Surface Loading} = \left(\frac{6000\ gal}{\min}\right) \div [(50\ ft)^2 * \pi - (25\ ft)^2 * \pi] = 1.0\ GPM/ft^2$$

The calculated surface loading rate of 1.0 GPM/ft^2 is at the top of the acceptable range of 0.4 to 1.0 GPM/ft^2.

HORIZONTAL VELOCITY CALCULATION (RECTANGULAR BASIN):

$$\textit{Maximum Horizontal Velocity} = \textit{maximum flow} \div \textit{flow area}$$

$$\textit{Hor Vel} = \left(\frac{6000\ gal}{\min}\right) * \left(\frac{ft^3}{7.48\ gal}\right) \div (9.6\ ft * 50\ ft) = 1.67\ ft/\min$$

The calculated value for the maximum horizontal flow velocity is less than the maximum recommended value of 2 ft/min.

HORIZONTAL VELOCITY CALCULATION (CIRCULAR BASIN):

For a circular basin the maximum horizontal flow velocity occurs at the smallest radius of the sedimentation basin.

$$Vel=\left(\frac{6000\ gal}{\min}\right)*\left(\frac{ft^3}{7.48\ gal}\right)\div(2*\pi*25\ ft*16.3\ ft) = 0.31\ ft/\min$$

The calculated value for the maximum horizontal flow velocity is less than the maximum recommended value of 2 ft/sec.

WEIR OVERFLOW RATE (RECTANGULAR BASIN):

The overflow weir length for a rectangular basin is twice the length of the overflow trough. The maximum design flow rate will be used to calculate the maximum weir overflow rate.

Three troughs with a length based on the sedimentation basin width of 50 feet will be used.

$$Overflow\ Weir\ Length= 3\ troughs * \frac{100\ ft}{trough} = 300\ ft$$

$$Max\ Weir\ Overflow\ Rate= maximum\ flow\ rate \div length\ of\ weir$$

$$Weir\ Overflow\ Rate=\left(\frac{6000\ gal}{\min}\right) \div 300\ ft= 20\ GPM/ft$$

The calculated value for the maximum weir overflow rate is at the recommended limit.

WEIR OVERFLOW RATE (CIRCULAR BASIN):

The overflow weir for a circular basin is at the outside radius of the basin.

$$Overflow\ Weir\ Length = \pi * outside\ diameter = 314\ ft$$

$$Weir\ Overflow\ Rate = \left(\frac{6000\ gal}{\text{min}}\right) \div (\pi * 100\ ft) = 19.1\ GPM/ft$$

The calculated maximum weir overflow rate is less than the recommended maximum weir overflow rate.

COLLECTION CHANNEL DEPTH CALCULATION:

The flow into the sedimentation basin effluent channel is spatially varied. To calculate the maximum depth for a channel with spatially varied flow, the "Method of Singular Point" was used[23]. Since the calculations used in the "Method of Singular Point" are extensive they are omitted here.

For a circular basin with a collection channel two feet wide the maximum water depth was found to be 0.8 feet. To allow for freeboard the collection channel will be two feet wide by one and one half feet deep.

For a rectangular basin with three collection channels with a width of one foot (see Figure 32) the maximum water depth was found to be 0.5 feet. To allow for freeboard the collector channel will be one foot by one foot.

TURBIDITY REMOVAL COST OPTIMIZATION

The ultimate goal of the plant design is to optimize the cost to remove the turbidity after it has been coagulated and flocculated. It is assumed that the coagulation and flocculation steps have been performed to optimum standards by following recommended design parameters. The turbidity will then be removed by two processes. Those processes are sedimentation and filtration.

The cost of filtration is a function of the amount of turbidity that the filter must remove. The more turbidity removed the higher the filtration costs. The effluent turbidity of a sedimentation basin is a function of the basin size. The larger the basin size the less the effluent turbidity but the higher the cost. Therefore, an optimal level of turbidity in the water leaving the sedimentation basin exists such that the total cost of turbidity removal is minimized.

Sedimentation

To calculate the cost of sedimentation as a function of the sedimentation basin effluent turbidity a function was derived relating the size of the basin to the amount of turbidity it could remove. As the size of the basin increases so does the cost, with the benefit of increased removal efficiency. Both a rectangular and a circular basin will be analyzed to see if there is any cost advantage between the two configurations.

Since a sedimentation basin is constructed of concrete the cost of cast in place concrete will be used to determine the cost of a sedimentation basin. The cost of cast in place concrete was taken from Means Construction Guide[24] and includes material and labor costs. The costs for the formed walls and poured floor will be separated, as there is a significant difference between the two.

The terminal settling velocity for floc particles (see Equation 15) will be used to determine the time it takes a particle to settle a fixed distance. The smallest particle removed is the particle with a settling time equal to the time the particle moves across the basin. The smaller the size of the particle removed the more efficient the basin.

of the smallest particle removed to the sedimentation basin effluent turbidity.

$$Effluent\ NTU = EXP\left[(d)^2 * 4.332*10^{-6}\right] \qquad (27)$$

Where: Eff. NTU= sed. basin effluent turb. (NTU)
d= particle diameter (microns)
4.332×10^{-6}= empirical constant (1/particle diameter^2)

As Equation 27 is an empirical equation the units associated with the empirical constant are meaningless.

Equation 27 is based on the authors practical experience in operating the NTMWTP and the equations that govern the settling of particles. For that reason Equation 27 is only valid for the NTMWTP. Equation 27 is also valid for a particle diameter of up to 800 microns, above that the sedimentation basin effluent turbidity would be excessively high. To further strengthen this argument it would take a 800 micron particle only 1 hour to settle 10 feet, which is well under the minimum recommended sedimentation detention time of 2 hours.

Personal experience has shown that when the NTMWTP is operating at low flows or below maximum design the effluent turbidity is approximately 2.0 Nephelometric Turbidity Units (NTU's), which when using Equation 27 equates to being able to settle out a 400 micron particle. It would take a 400 micron particle approximately 4 hours to settle 10 feet micron particle approximately 4 hours to settle 10 feet which equates to the detention time in a basin operating at low flows. Therefore, the constant in Equation 27 was set a 4.332×10^{-6} so that a basin that was capable of removing a 400 micron particle would have an effluent turbidity of 2 NTU's.

As was previously discussed the settling velocity is a function of the particle diameter squared which accounts for the particle diameter squared in the exponent.

The exponential term reflects the relationship between a high effluent turbidity of the sedimentation basin and the ineffectiveness of the basin in settling out large particles. The assumption here is that the smaller particles are also not being removed.

Use of this equation in theoretically sizing a sedimentation basin has shown a close correlation between a theoretically sized basin and a basin sized in the previous section using generally acceptable design parameters. Therefore, Equation 27 is believed to reasonably equate the sedimentation basin effluent turbidity with the smallest particle that a particular basin could remove.

COSTS FOR SEDIMENTATION CALCULATIONS:

Basic Assumptions:
Expected Basin Life: 30 years
Basin Wall Thickness: 1 foot
Basin Flow: 8.6 MGD

To account for the time value of money the present value of the sedimentation basin was amortized over the expected life period of thirty years. An interest rate of 5.5% compounded continuously was used as that is the interest rate that is currently available for a thirty year bond. The assumption made here is that the cost for the water treatment plant will be financed through bond money.

The following equation was used to calculate the cost of sedimentation in dollars per 1000 gallons treated.

$$(cost)*\left(\frac{1\ day}{8600000\ gal}\right)*\left(\frac{1\ yr}{365\ days}\right)*\left(\frac{A}{P},r,n,\right)=\frac{\$}{1000\ gal} \qquad (28)$$

Where: (A/P,r%,n)= present worth to equal annual series
A= equal annual worth
P= present worth
n= number of periods
r= annual percent interest

$$\frac{A}{P}\ \text{compounded continuously}=\frac{(e^{r}-1)}{(1-e^{-r*n})}$$

For the purposes of this design the specific gravity of the settling particle was taken to be 1.08 which falls into the generally acceptable range[25] of 1.06 to 1.10. The minimum water temperature at the NTMWTP of 40° F was used when determining the maximum water viscosity. The maximum viscosity was used as it would cause the minimum settling velocity for any given particle size.

A spreadsheet was developed that showed the increase in volume of a rectangular basin against the smallest diameter particle that the basin could remove. The basin's volume was varied by allowing the length to increase and fixing the depth at 10 feet and the width at 50 feet. The preceding two fixed parameters were chosen because they were the depth and width calculated in the previously section using generally acceptable design parameters. The flow through the basin was fixed at the maximum flow rate of 6000 GPM.

The smallest particle removed is the particle whose settling time equals the time across the basin. Equation 27 was used to calculate the sedimentation basin effluent turbidity. Equation 28 was used to calculate the sedimentation cost. Output from the spreadsheet is shown in Table 6.

Table 6

Rectangular Basin Turbidity Removal Costs

SMALLEST DIAMETER PARTICLE REMOVED (M^-6)	LENGTH NECESSARY TO REMOVE PARTICLE (FT)	COST ($/1000 GAL)	EFFLUENT TURBIDITY (NTU)
40	37212	0.792431	1.0
401	370	0..008562	2.0
504	234	0.005671	3.0
566	185	0.004639	4.0
610	160	0.004089	5.0
644	143	0.003739	6.0
671	132	0.003498	7.0
693	123	0.003322	8.0
713	117	0.003176	9.0
730	111	0.003061	10.0
744	107	0.002973	11.0
758	103	0.002889	12.0
770	100	0.002821	13.0
781	97	0.002761	14.0
791	95	0.002709	15.0
801	92	0.002659	16.0
809	90	0.002620	17.0
817	89	0.002582	18.0
825	87	0.002545	19.0
832	86	0.002514	20.0

The same procedure was used for a circular basin with a fixed depth of 16.3 feet and the depth that the particle must settle of 10 feet. The depth of 16.3 feet was used as it was the depth calculated in the previous section using generally acceptable design parameters. The influent portal to the sedimentation section of a circular basin is beneath the water surface and about 8 feet above the bottom of the basin (see Figure 33 in the previous section). For this reason the depth that the particle must settle was set at 10 feet which allows for some mixing into the layer above the sedimentation influent level which is at 8 feet. The maximum velocity in a circular basin with a 25 foot radius flocculation zone was calculated to be 0.31 feet/minute so the previous assumption is reasonable as the velocity is slow enough that an excessive amount of mixing will not occur. Table 7 shows the output form the spreadsheet for a circular basin.

Table 7

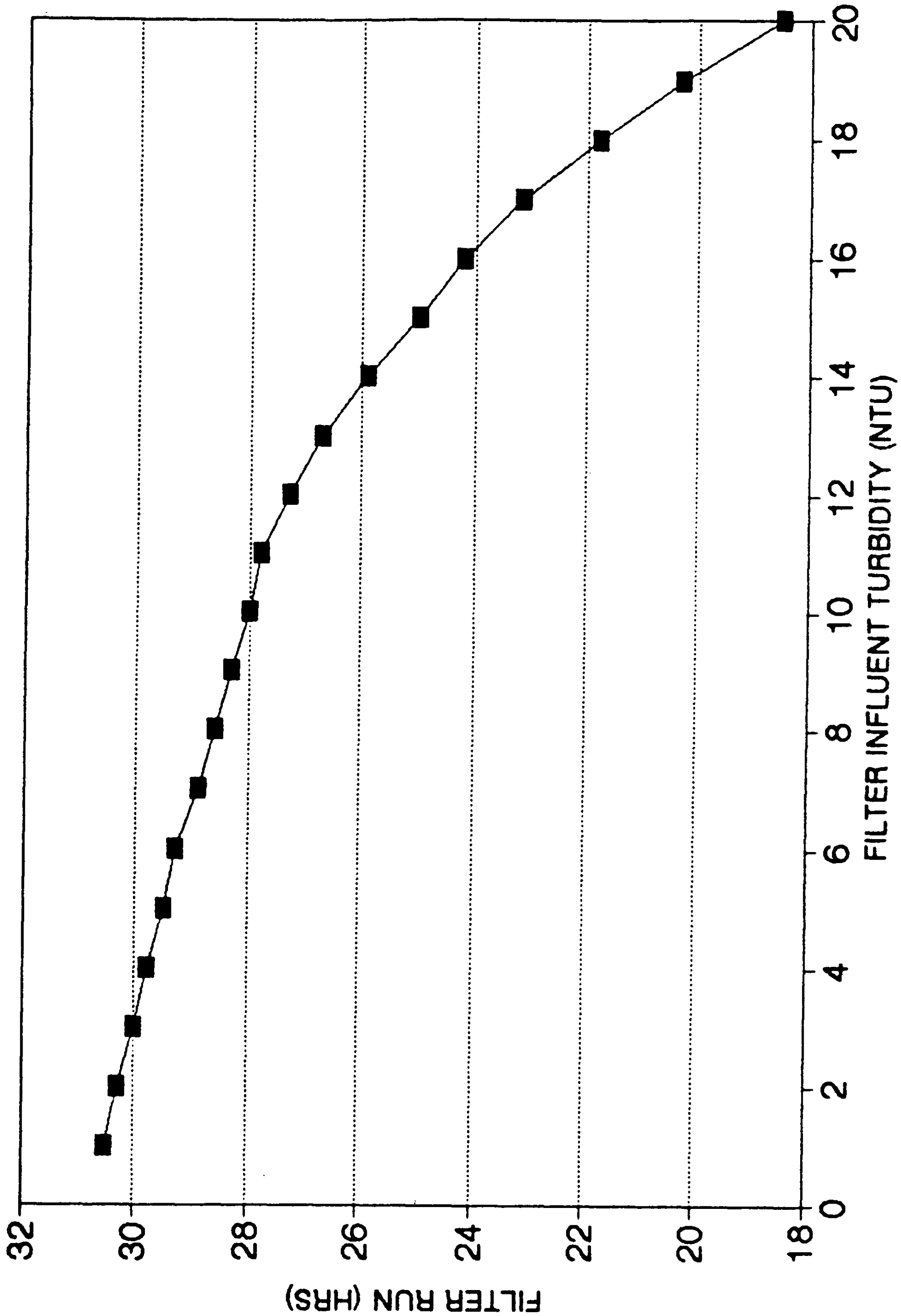
Circular Basin Turbidity Removal Costs

SMALLEST DIAMETER PARTICLE REMOVED (M^-6)	RADIUS NECESSARY TO REMOVE PARTICLE (FT)	COST ($/1000 GAL)	EFFLUENT TURBIDITY (NTU)
40	583	0.21523	1.0
401	63	0.00663	2.0
504	53	0.005223	3.0
566	48	0.004678	4.0
610	46	0.004376	5.0
644	44	0.004178	6.0
671	43	0.004039	7.0
693	42	0.003936	8.0
713	41	0.003849	9.0
730	40	0.003780	10.0
744	40	0.003727	11.0
758	39	0.003676	12.0
770	39	0.003634	13.0
781	38	0.003597	14.0
791	38	0.003565	15.0
801	38	0.003534	16.0
809	38	0.003509	17.0
817	37	0.003486	18.0
825	37	0.003463	19.0
832	37	0.003443	20.0

Filtration

COSTS FOR FILTRATION:

Data from the NTMWTP of the filter influent turbidity and the length of the filter run were compiled and are shown in Table 8 and graphed in Figure 34. A filter run is the length of time a filter is capable of filtering water. The filter run is limited because the filter's porosity is reduced by the filtered solids causing an excessive headloss through the filters. As expected the higher the filter influent turbidity the shorter the filter run (see Figure 34).



Filter Run VS
Filter Influent Turbidity

Figure 34

Table 8

Filter Influent Turbidity VS Filter Run

FILTER INF. TURB	FILTER RUN (HRS)
1.0	30.5
2.0	30.3
3.0	30.0
4.0	29.8
5.0	29.5
6.0	29.3
7.0	28.9
8.0	28.6
9.0	28.3
10.0	28.0
11.0	27.8
12.0	27.3
13.0	26.7
14.0	25.9
15.0	25.0
16.0	24.2
17.0	23.2
18.0	21.8
19.0	20.3
20.0	18.5

In order to convert this data into the cost for filtration the capital and fixed costs for filtration were determined. The capital costs account for the costs of installing the filtration facility and the fixed costs are the cost of backwashing the filter to remove the filtered turbidity so it can begin another filter run.

In order to develop comparable capital and operating costs, the costs were broken down into dollars per thousand gallons of water filtered. Capital costs were then amortized over the reasonably expected life period of thirty years.

To accommodate for inflation, cost indexes published by the Richardson Construction Cost Trend Reporter[26] were used. These indexes are provided by the Bureau of Census, Department of Commerce. This index is fixed in 1983 at a value of 100 and for other years the index is varied to compensate for inflation. To utilize these indexes the capital cost as found from the year of the source is multiplied by the 1993 cost index and then divided by the index from the year of the source.

The source used to determine the capital costs for filtration facilities were the cost to have two filters installed at the NTMWTP in 1985.

CAPITAL COSTS FOR FILTRATION:

Basic Assumptions:
- Expected life: 30 years
- Bid price: \$655,000
- Flow rate: 3.89 MGD per filter
- Source year: 1985
- 1985 Cost Index: 96.2
- 1993 Cost Index: 121.5

$$\$655{,}000 * \left(\frac{1\ day}{3888000\ gal}\right) * \left(\frac{1\ year}{365\ days}\right) * \left(\frac{A}{P}, r\%, n\right) * \left(\frac{121.5}{96.2}\right) = \frac{\$0.041}{1000\ gal}$$

Where: (A/P,r%,n)= present worth to equal annual series
A= equal annual worth
P= present worth
n= number of periods
r= annual percent interest

$$\frac{A}{P}\text{ compounded continuously} = \frac{(e^{r}-1)}{(1-e^{-r*n})}$$

OPERATING COSTS FOR FILTRATION:

Operating cost for filtration come from the cost of backwash water and is highly site specific. The cost of the raw water and the amount of backwash water necessary to properly maintain the filters have a profound impact on the operating costs for filtration.

Data from the NTMWTP:

Backwash water cost: \$0.65/1000 gal (raw water cost)

$$\left(\frac{\$0.65}{1000\ gal}\right) * \left(\frac{80{,}000}{backwash}\right) = \$52.00/backwash \qquad (28)$$

Using these cost data a spreadsheet was developed to calculate the cost of filtration for a range of filter influent turbidities (see Table 9).

First the amount of water treated for each filter run was calculated by the following equation (see Table 9):

$$1000\ gallons = (filter\ run) * (80\ thousand\ gal/hr) \qquad (29)$$

Where: 80 thousand gallons/hour= typical filter flow rate
Filter run= length of time the filter is in service in hours

The capital cost for each filter influent turbidity or filter run was calculated by multiplying the gallons of water treated in units of 1000 gallons by the previously calculated cost of \$0.041/1000 gallons (see Table 9)

Table 9

Filtration Costs in $/1000 gallons as a Function of Filter Influent Turbidity

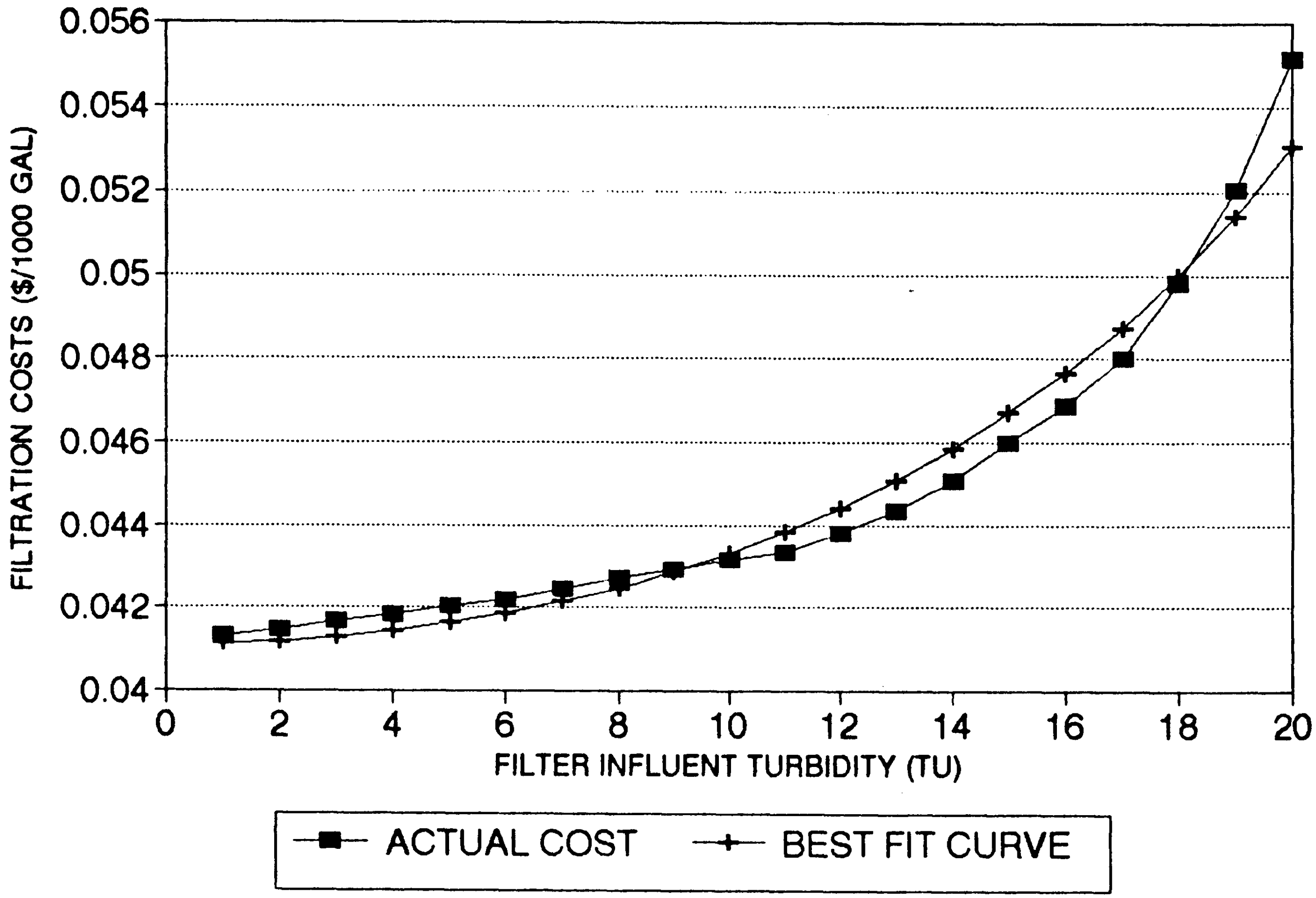
INF TURB (NTU)	FILTER RUN (HOURS)	1000 GAL PRODUCED	CAPITAL COSTS ($)	FIXED COSTS ($)	$/1000 GALLONS	BEST FIT CURVE
1	30.5	2440	100.04	52.00	0.0623	0.061
2	30.3	2424	99.38	52.00	0.0624	0.062
3	30	2400	98.40	52.00	0.0626	0.062
4	29.8	2384	97.74	52.00	0.0628	0.062
5	29.5	2360	96.76	52.00	0.0630	0.062
6	29.3	2344	96.10	52.00	0.0631	0.062
7	28.9	2312	94.79	52.00	0.0634	0.063
8	28.6	2288	93.80	52.00	0.0637	0.063
9	28.3	2264	92.82	52.00	0.0639	0.063
10	28	2240	91.84	52.00	0.0642	0.064
11	27.8	2224	91.18	52.00	0.0643	0.064
12	27.3	2184	89.54	52.00	0.064	0.065
13	26.7	2136	87.57	52.00	0.0653	0.065
14	25.9	2072	84.95	52.00	0.06607	0.066
15	25	2000	82.00	52.00	0.067	0.067
16	24.2	1936	79.37	52.00	0.06787	0.068
17	23.2	1856	76.09	52.00	0.0690	0.068
18	21.8	1744	71.50	52.00	0.0707	0.069
19	20.3	1624	66.58	52.00	0.0732	0.071
20	18.5	1480	60.68	52.00	0.0761	0.072

As previously stated, the operating costs for backwashing are fixed at \$52.00 and are independent of the filter influent turbidity or filter run.

To calculate the total cost for filtration at each filter influent turbidity the capital costs and operating were summed (see Table 9).

Figure 35 shows a plot of the \$/1000 gallons of filtered water versus the filter influent turbidity. An equation was derived to best fit the data from Table 7 and graphed in Figure 35 and is as follows:

$$\frac{\$}{1000\ gal} = 2.4 * \frac{1}{[1500 - (NTU)^2]^{0.5}} \qquad (30)$$



Filtration Cost VS
Filter Influent Turbidity

Figure 35

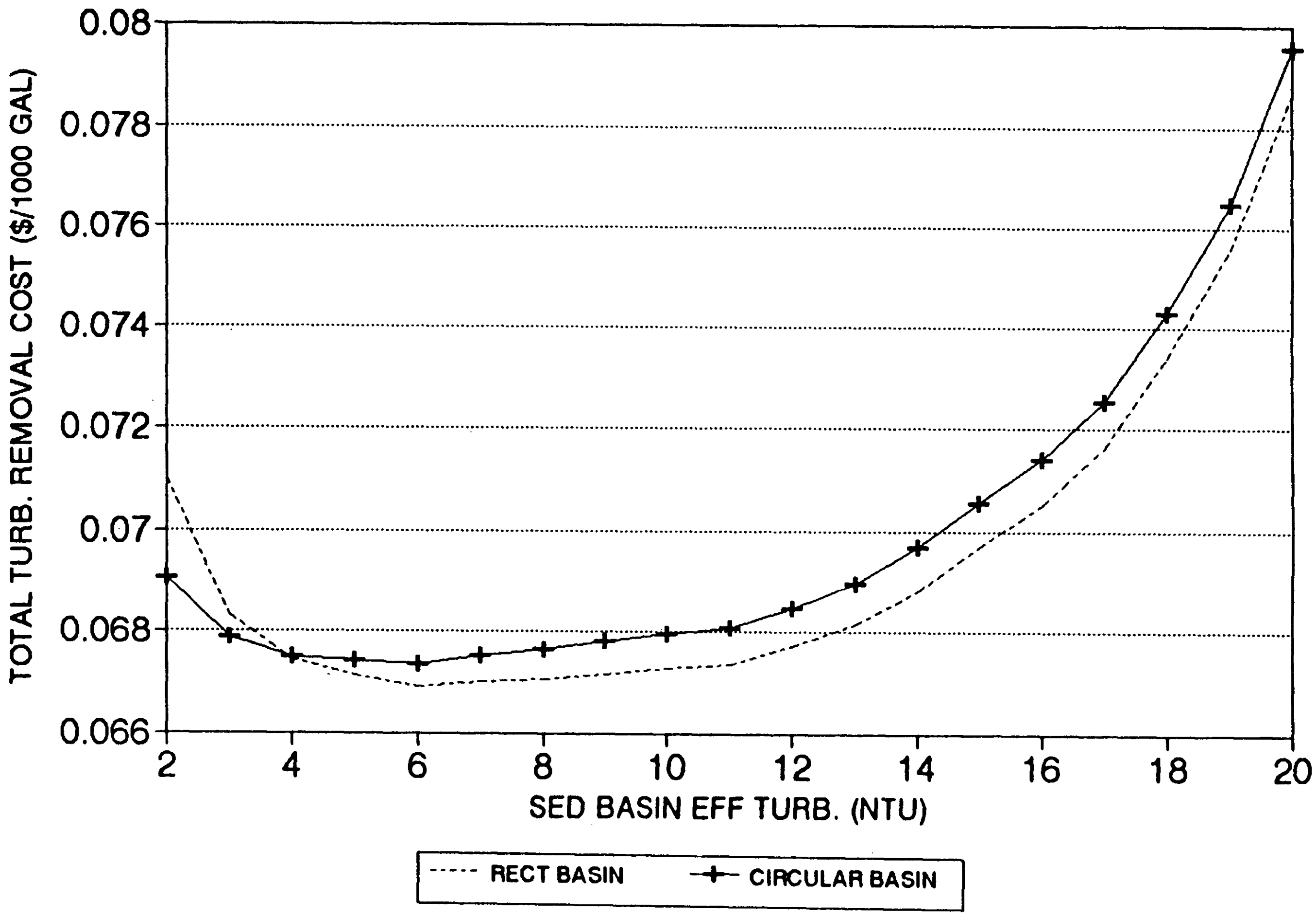
Total Cost for Turbidity Removal

For each sedimentation basin effluent turbidity a cost for that basin was calculated. Also for each filter influent turbidity a cost was calculated to remove the turbidity to a predetermined level of 0.05 NTU's. The cost of sedimentation and filtration for the same turbidity were summed to come up with the total cost for turbidity removal. This data is shown in Table 10 and graphed in Figure 36.

Table 10

Total Turbidity Removal Costs

SED BASIN EFF. TURB. (NTU)	FILTER COST ($1000 GAL)	RECT SED COST ($/1000 GAL)	RECT TOTAL COST ($/1000 GAL)	CIRCULAR SED COST ($/1000 GAL)	CIRCULAR TOTAL COST ($/1000 GAL)
1	0.062311	0.792431	0.854742	0.215299	0.27761
2	0.062452	0.008562	0.071014	0.00663	0.069082
3	0.062667	0.005671	0.068338	0.005223	0.06789
4	0.062812	0.004639	0.067451	0.004678	0.06749
5	0.063034	0.004089	0.067123	0.004376	0.06741
6	0.063184	0.003739	0.066923	0.004178	0.067362
7	0.063491	0.003498	0.066989	0.004039	0.06753
8	0.063727	0.003322	0.067049	0.003936	0.067663
9	0.063968	0.003176	0.067144	0.003849	0.067817
10	0.064214	0.003061	0.067275	0.00378	0.067994
11	0.064381	0.002973	0.067354	0.003727	0.068108
12	0.06481	0.002889	0.067699	0.003676	0.068486
13	0.065345	0.002821	0.068166	0.003634	0.068979
14	0.066097	0.002761	0.068858	0.003597	0.069694
15	0.067	0.002709	0.069709	0.003565	0.070565
16	0.06786	0.002659	0.070519	0.003534	0.071394
17	0.069017	0.00262	0.071637	0.003509	0.072526
18	0.070817	0.002582	0.073399	0.003486	0.074303
19	0.07302	0.002545	0.075565	0.003463	0.076483
20	0.076135	0.002514	0.078649	0.003443	0.079578



Total Turbidity Removal Cost
VS Sed. Basin Effluent Turbidity

Figure 36

From Table 10 and Figure 36 the least expensive total turbidity removal cost for a rectangular or a circular basin is when the sedimentation basin effluent is 6 NTU's. For a sedimentation basin effluent turbidity less than 6 NTU's the cost for sedimentation becomes excessive causing an increase in the total turbidity removal cost. For a sedimentation basin effluent turbidity greater than 6 NTU's the cost of filtration becomes excessive and the total turbidity removal costs increase rapidly. The graph in Figure 36 is skewed to the right because the cost of filtration is much greater than the cost of sedimentation.

For a rectangular basin the optimal theoretical length was found to be 144 feet which is close to the length of 200 feet that was calculated in the previous section using acceptable design parameters.

For a circular basin the optimal theoretical radius was found to be 44 feet and corresponds well with the design radius of 50 feet as calculated in the previous section.

The theoretical design was smaller the practical design which should be true as the theoretical design does not take into account inlet, and outlet conditions, as well as thermal and wind currents.

The difference in cost between a rectangular basin and a circular basin is small enough that from a cost standpoint no distinction will be made. Thus for the purposes of a preliminary design either a circular or a rectangular basin would be appropriate. The choice of basin configuration will be made in the final design phase and based on space considerations.

SUMMARY OF THE PRELIMINARY PLANT DESIGN

The preliminary plant design is based on the design using generally acceptable engineering design parameters as previously calculated.

Coagulation

Basin Volume: 400 ft^3

Basin Dimensions: 7'-6" long
7'-6" wide
7'-6" deep
7'-1" water depth with 5" freeboard

Motor: 30 HP with a 80% efficiency rating

Basin will contain baffles to aid in mixing

Coagulant to be introduced at the mixing impeller

Flocculation

Two stage flocculation will be used for a rectangular basin and single stage for a circular basin.

Volume (rectangular basin 1st & 2nd stage): 8021 ft^3

Dimensions (rectangular basin 1st & 2nd stage):
50′ long
17′ wide
11′ deep
9.4′ water depth with 1.6′ freeboard

Volume (circular): 16042 ft^3

Dimensions (circular): 25′ radius
10′ deep
8.2′ water depth with 1.8′ freeboard

Motor (rectangular): 1st stage: 1 HP with an 80% efficiency rating
2nd stage: 0.75 HP with an 80% efficiency rating

Motor (circular): 2 HP with an 80% efficiency rating

Sedimentation

Volume (rectangular of circular): 96257 ft^3

Dimensions (rectangular):
200′ long
50′ wide
12′ deep
9.6′ water depth with 1.6′ freeboard and 0.8′ accumulated solids

Dimensions (circular):
50′ radius
19′ deep
16.3′ water depth with 1.8′ freeboard and 0.9′ accumulated solids

The decision on whether to use a rectangular or circular basin will be made during the final design. The decision will be based on space constraints and future expansion.

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