Filtration

Porous Media Filtration

Definition: Removal of colloidal (usually destabilized) and suspended material from water by passage through layers of porous media.

Water treatment: turbidity removal

Wastewater treatment: tertiary filtration (removal of very fine suspended particles)

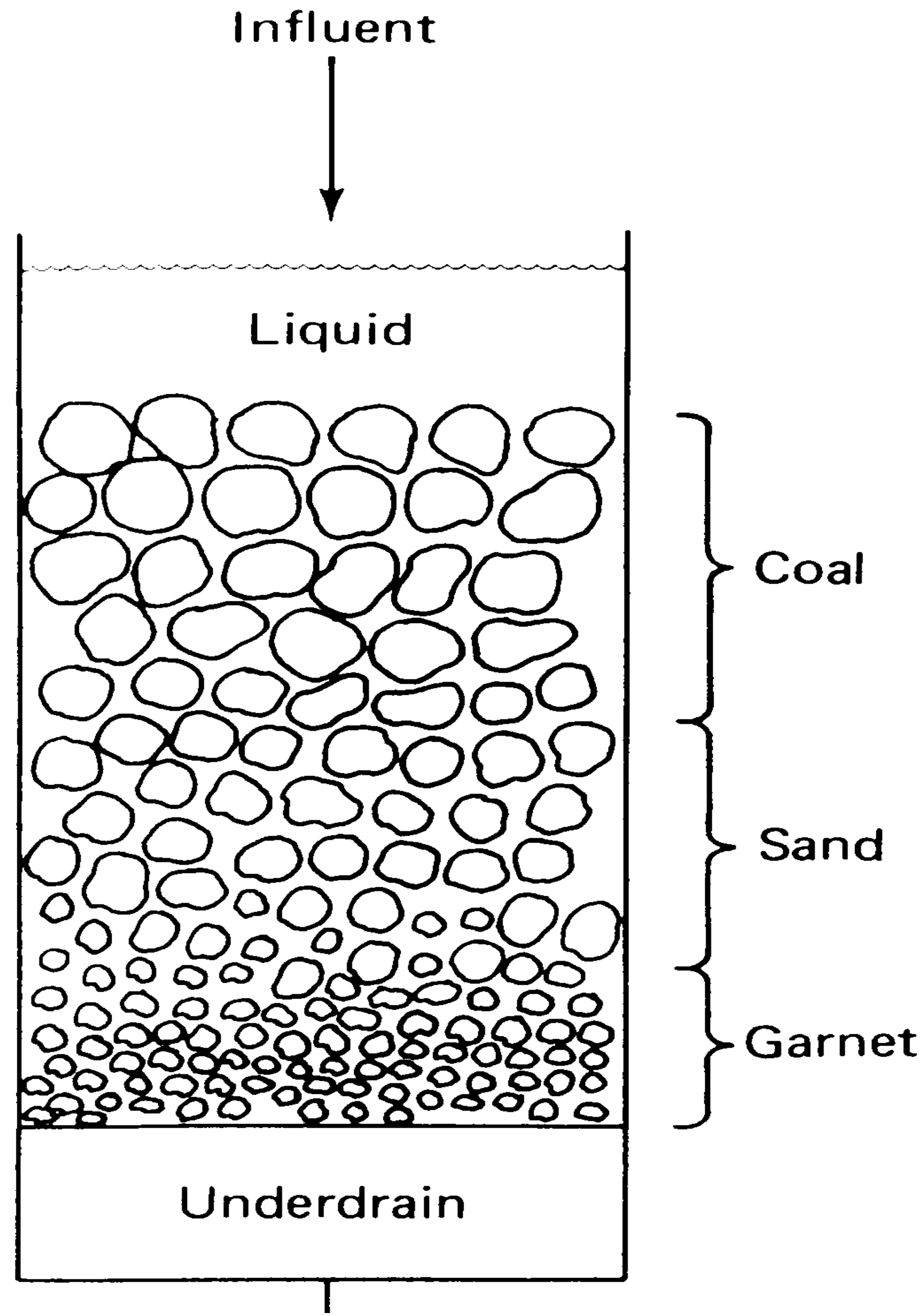
TYPES OF FILTERS

The filters described below are modern types used for water/wastewater treatment purposes. Variations of these filter types and other types are discussed at the end of this section. In all filters the primary design/operating parameters are:

- quality (SS concentration) of the effluent.
- headloss through the filter and appurtenances.

Deep Granular Filters

Deep granular filters are made of granular material (sand, anthracite, garnet) arranged in a bed to provide a porous media as shown in the figure below. Filter bed is supported by gravel bed as also shown below. Flow is typically in the downflow mode. Upflow mode is used but much less frequently.

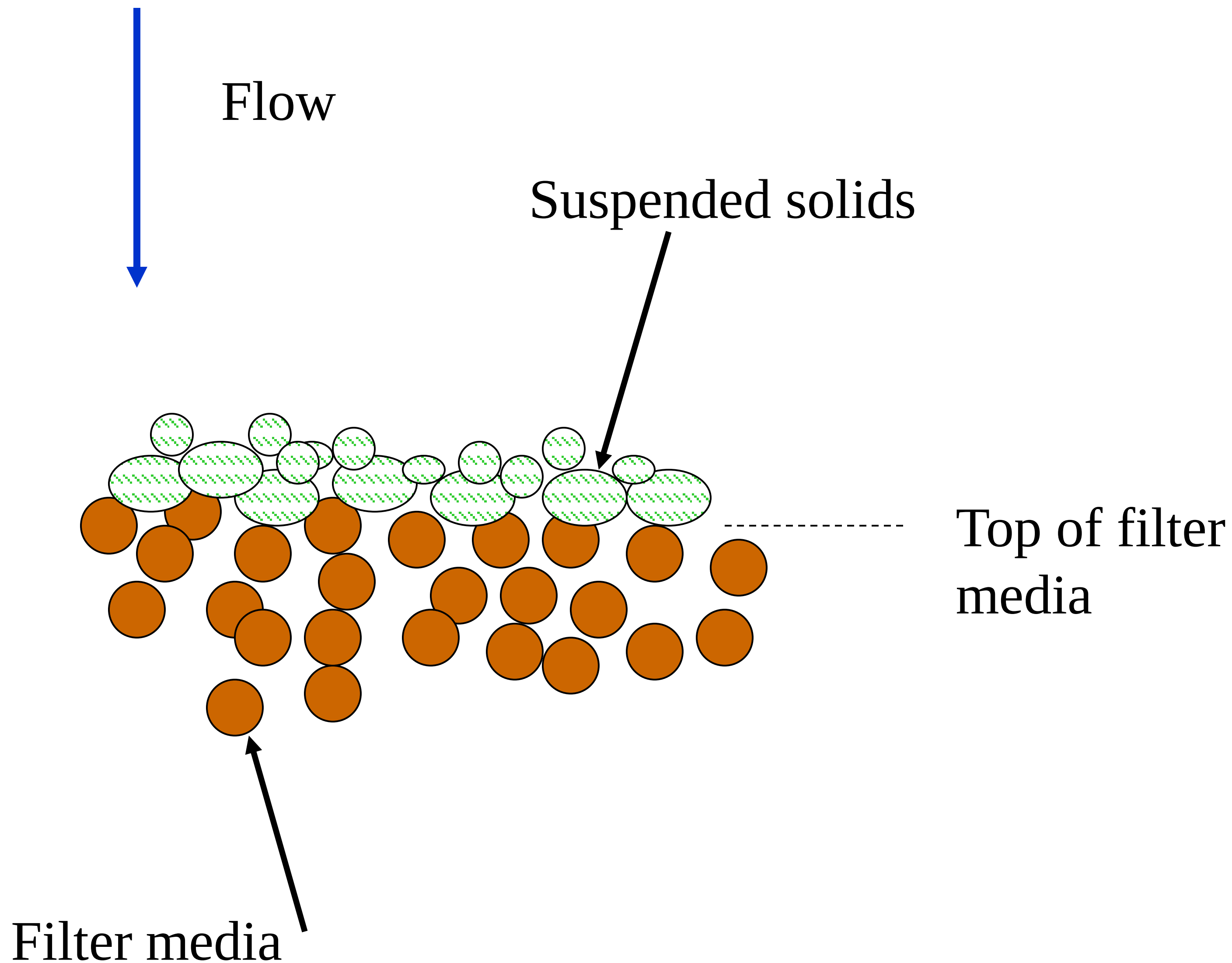

Influent
Liquid
Coal
Sand
Garnet
Underdrain

Mechanisms of suspended solids removal

There are several mechanisms of SS removal in deep granular filters.

•*Surface removal (straining)*

Mechanical straining caused by a layer of suspended solids (from the feed water) which builds up on the upper surface of the porous media. This type of removal is to be avoided because of the excessive head loss that results from the suspended solids layer's compressibility.

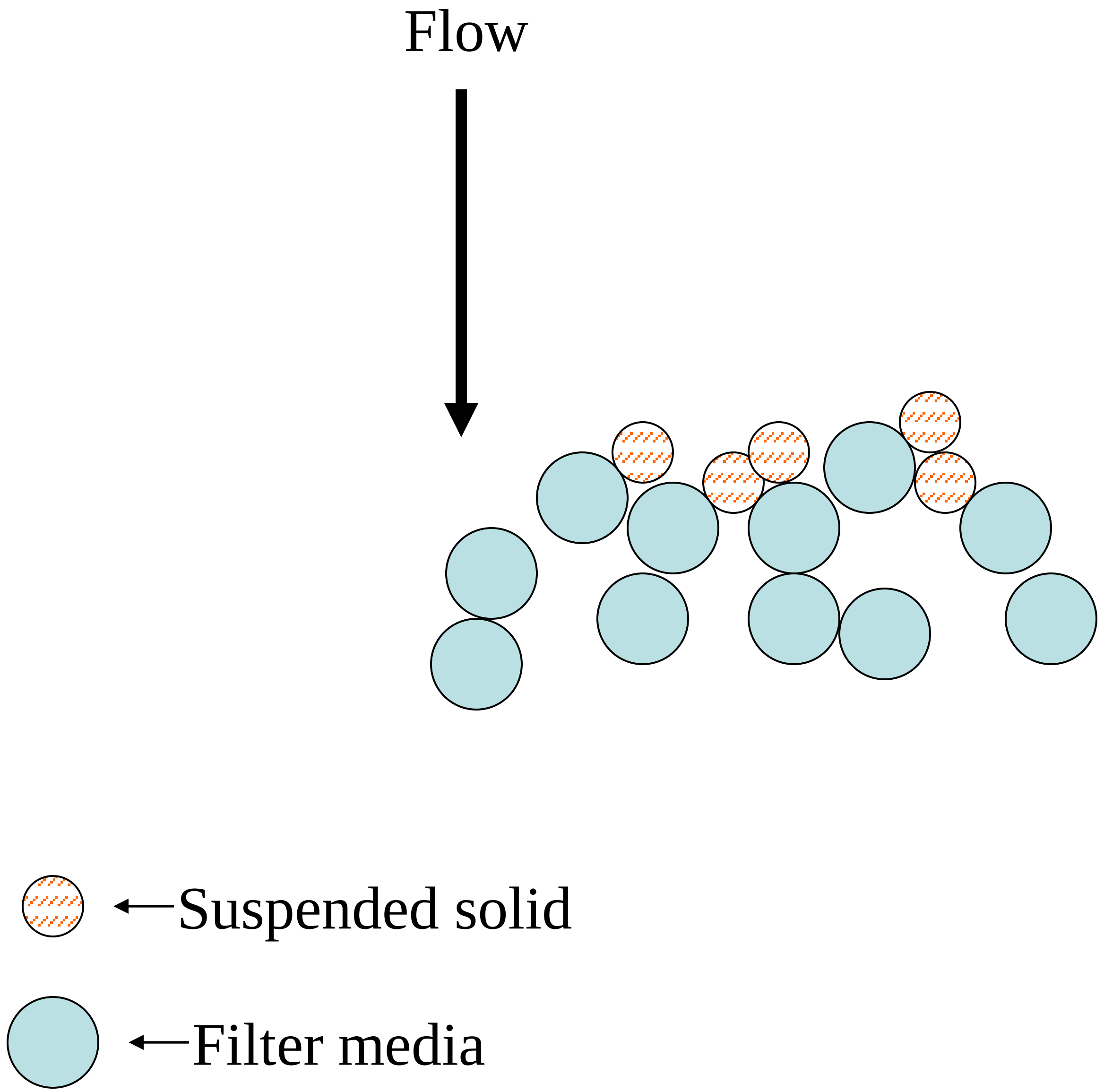

Flow
Suspended solids
Top of filter media
Filter media

• *Depth removal*

Depth removal refers to SS removal below the surface of the filter bed. There are two types of "depth removal".

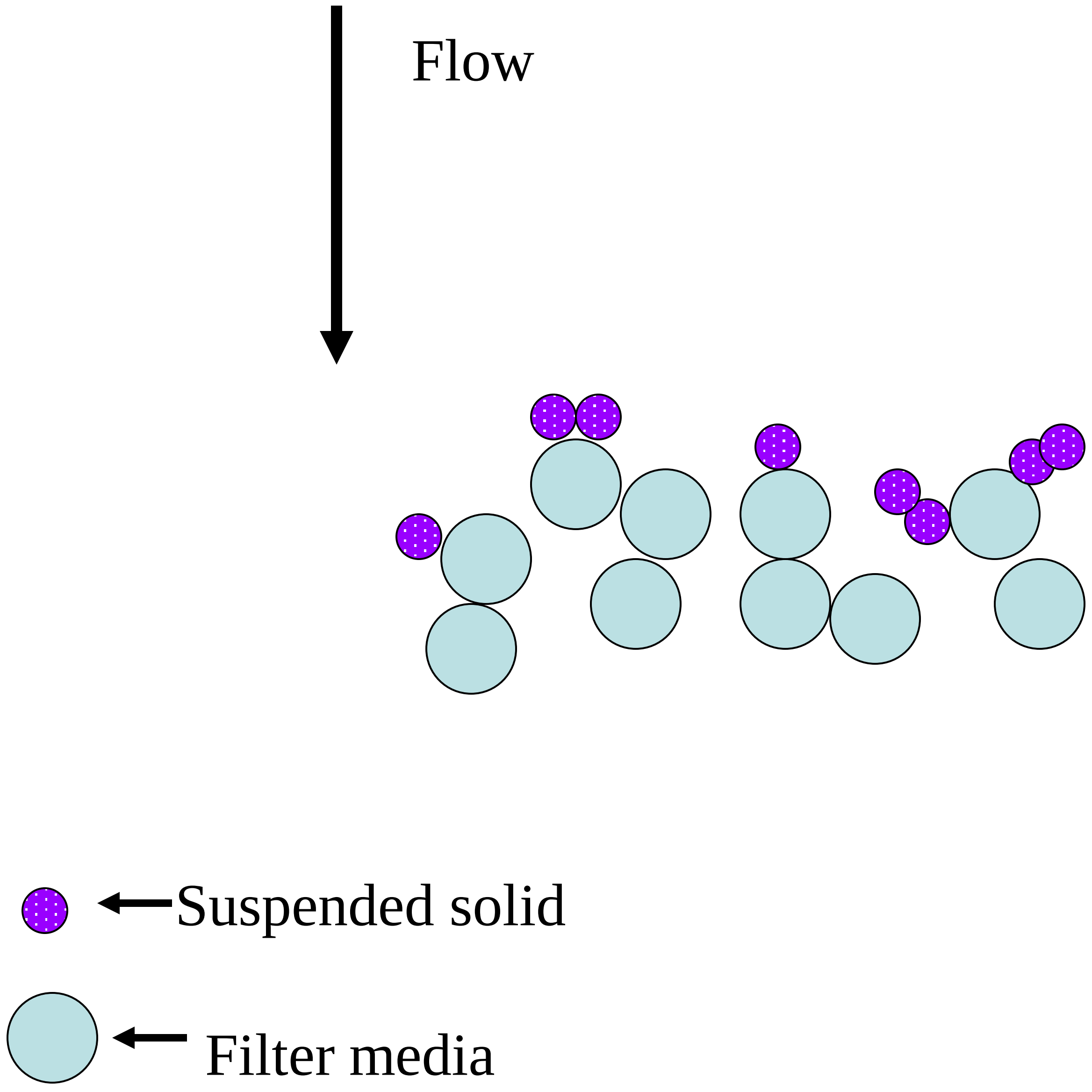
Interstitial straining

Larger particles become trapped in the void space between granular media particles.


Flow
Suspended solid
Filter media

Attachment

Suspended solids are typically flocculent by design (filter often follows coagulation/flocculation) or by nature (clays, algae, bacteria). Therefore, attachment or adsorption of suspended solids is a good possibility. Attachment can be electrostatic, chemical bridging or specific adsorption. Attachment is enhanced by addition of small amount of coagulant and as the filter bed becomes coated with suspended solids ("ripened" filter). It is easier for suspended solids to attach to other SS that are already attached to the filter media.


Flow
Suspended solid
Filter media

In general all three mechanisms of removal are occurring at the same time during a filter run. The relative predominance of these mechanisms depends on:

- character of media
- character of SS
- temperature
- flow rate
- bed depth
- time (throughput volume)

Filter Cycle

As filter run proceeds deposits build up in the upper portion of the filter bed. As a consequence void volume decreases, interstitial flow velocity increases with more hydraulic shear on the trapped and attached SS. This drives some of the filtered SS deeper into the filter bed.   Ultimately the SS get washed into the effluent.

At this point the filter must be backwashed to clean the filter bed surfaces. The filter is then put in the forward flow mode again.   It is possible (and likely) that the head loss through the bed becomes high enough that the bed has to be backwashed before the effluent quality becomes unacceptable. Head loss builds because the void space shrinks with time. Head loss is usually what determines time to backwashing. Therefore, it is important to know the hydraulics of granular filters.

Hydraulics of Deep Granular Filters

Hydraulics of flow through porous media can be described by D'Arcy's law if flow is laminar.

$$V = K_p S_l = K_p \frac{h_f}{L}$$

V = superficial approach velocity (ft/min).

K_p = coefficient of permeability (ft/min). This will change with time in the filter.

S_l = hydraulic gradient (h_f/L) dimensionless

h_f = frictional head loss (ft)

L = depth of filter (ft).

Alternatively, the empirical Carmen- Kozeny , Fair-Hatch, Rose or other equations are more appropriate because the pore volume will continually change as suspended solids are removed.

For example the Carmen-Kozeny equation is often used:

$$\frac{h_f}{L} = J \frac{\nu}{g} \frac{(1-\varepsilon)^2}{\varepsilon^3} V \left(\frac{\sigma_s}{d_p} \right)^2$$

ν = kinematic viscosity (ft^2/sec)

J = packing factor (empirical) ~ 6 for laminar flow

ε = porosity = void volume fraction of filter bed

g = acceleration of gravity (ft/sec^2)

d_p = measured particle dia (ft).

d_p is commonly taken as geometric mean of adjacent sieve sizes that pass and retain the particles. For non-uniform size media particles divide the bed into incremental layers and use geometric mean size in each layer $(d_1 \times d_2)^{0.5} = d_p$ for that layer. Compute h_f/L for each layer and sum for total bed. d_1 is size passed d_2 is size retained for a particular layer. Sometimes the **effective size** of the particles is used here.

effective size = size for which 10% of sample (by wt.) is smaller (d_{10}) .

σ_s= shape factor, measure of particle irregularity

= 6 for spheres

= 8.5 for crushed granular media.

Typical sand filter media:

effective size = 0.5mm

uniformity coefficient. = 1.75

uniformity coefficient. = size for which 60% of sample (wt) is smaller (d_{60})/effective size. = d_{60}/d_{10}.

Headloss Development in Granular filter

The Bernoulli equation (conservation of energy) can be used to model head loss through a granular filter:

$$\frac{V_i^2}{2g} + \frac{p}{\gamma} + Z + h_f = \text{constant}$$

V_i = interstitial flow velocity, ft/sec.

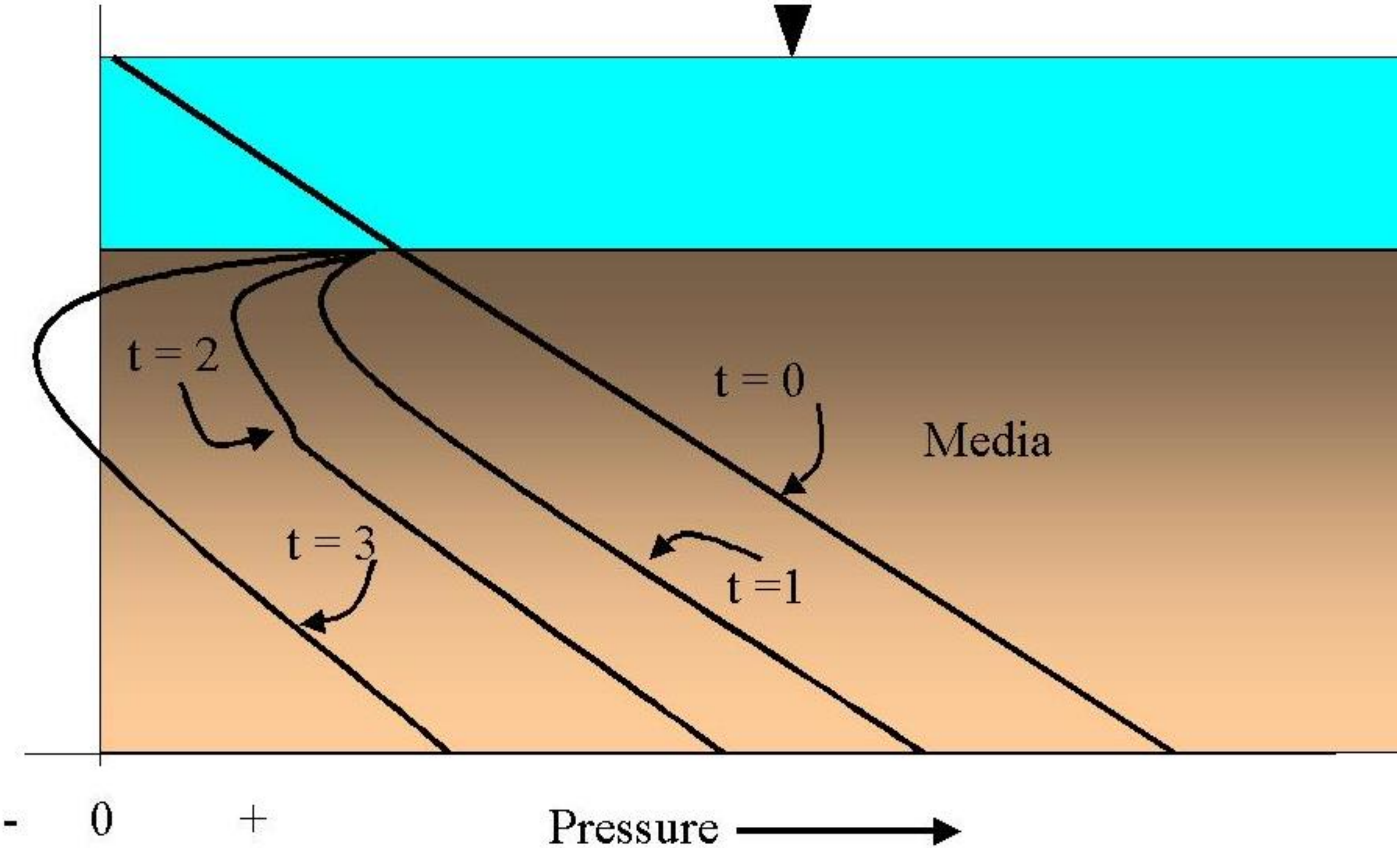
γ= specific wt of water = 62.4 lbs/ft

Z = depth measured from datum, ft.

For no flow (hydrostatic conditions):

$$\frac{p}{\gamma} + Z = \text{constant}$$

Head loss patterns change as the filter run proceeds; interstitial velocity increases as pore size decreases (first in the upper portions of filter) and as the velocity increases the frictional head loss increases. Since the progression of head loss increases non- uniformly throughout the filter we get the following head loss pattern.



t = 2
t = 0
Media
t = 3
t =1
-
0
+
Pressure

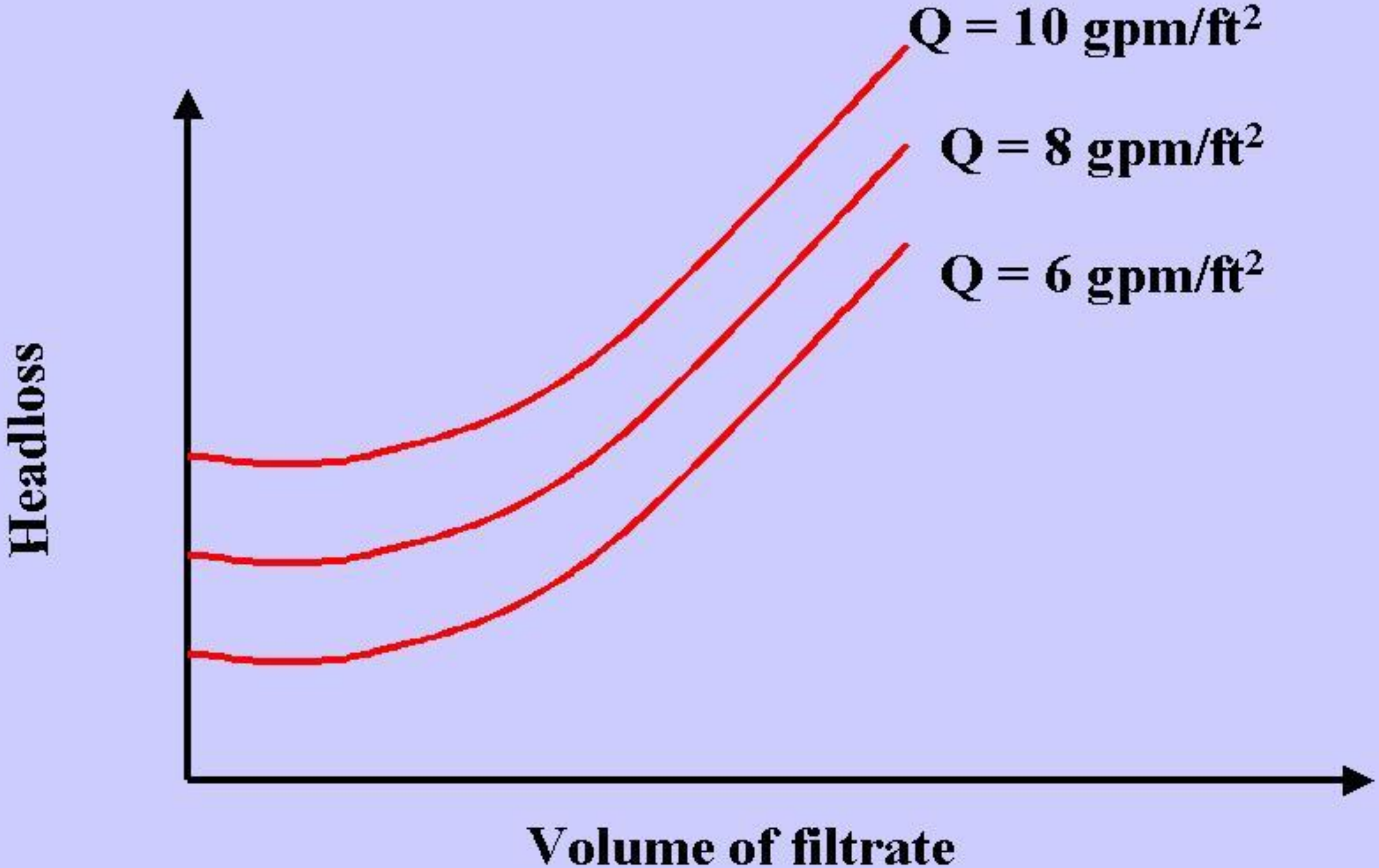
In the region of negative pressure degasification of the water can occur. This may cause air binding and reduction in the effective filter surface area. Negative pressure regions can also cause cracking of the filter (results in fissures in bed that allow unfiltered water to pass through to effluent).

Pretreatment

If the sand filter is not preceded with a coagulation/flocculation process (as is typically the case for water treatment systems), pretreatment of the suspended solids is often employed, particularly if the water contains fine clays. Pretreatment is usually the addition of coagulant just before the filter. The filter acts as a flocculation process as described earlier.

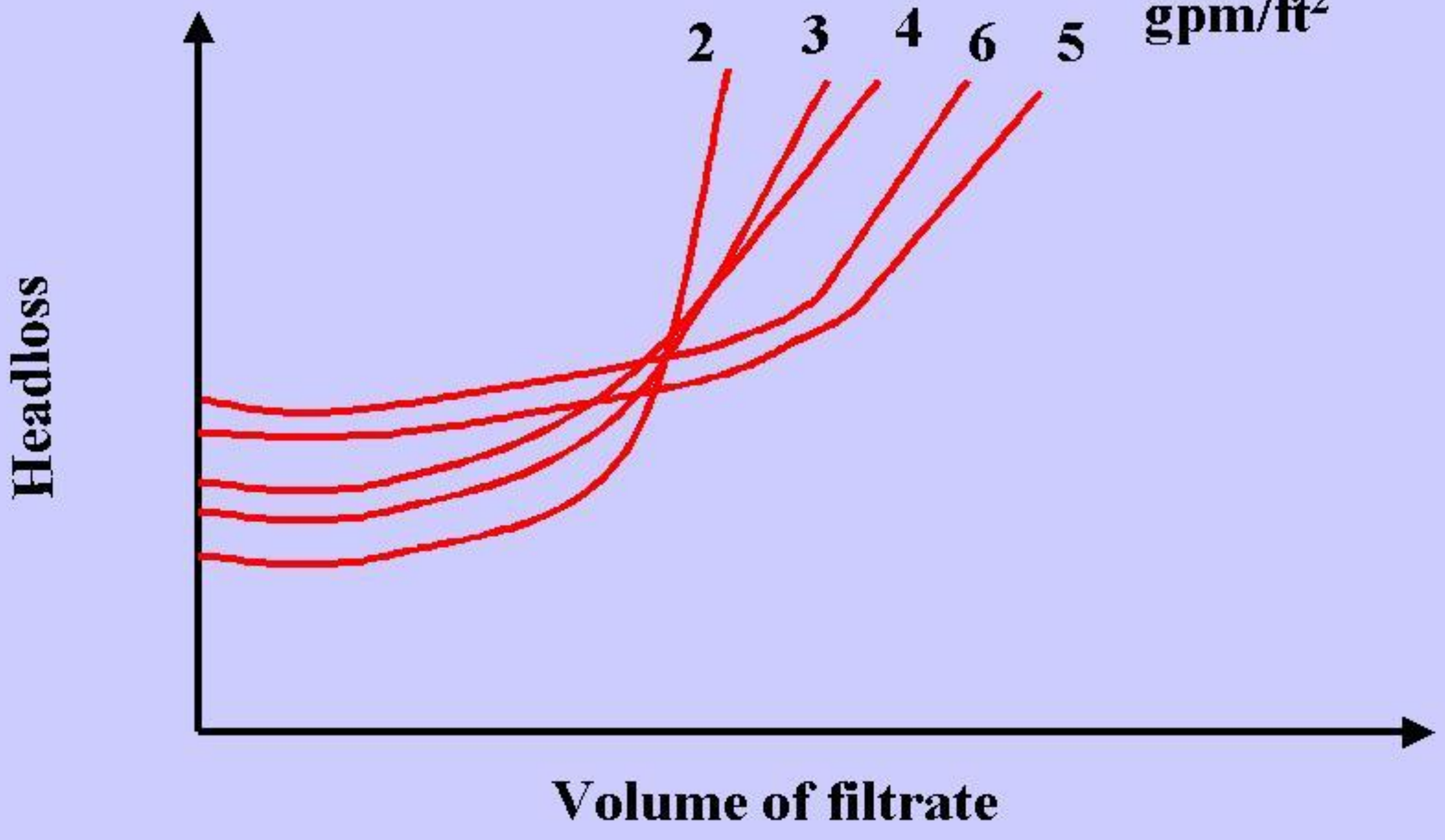
Head loss patterns

When depth removal is the primary mechanism for SS removal the head loss pattern is shown in the following figure. Head loss increases with surface loading rate due to higher solids loading rate as well as higher frictional head loss.

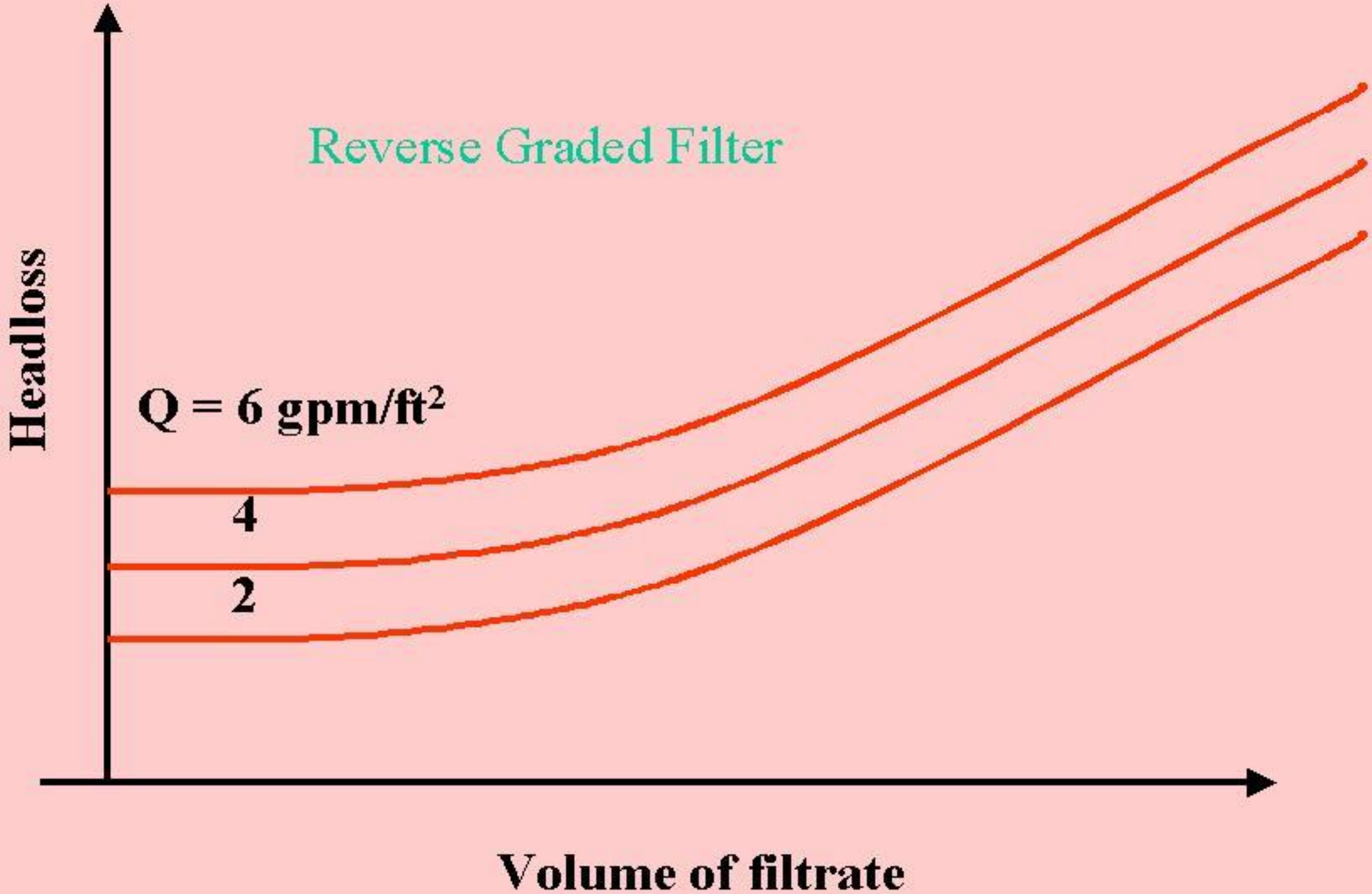


Q = 10 gpm/ft²
Q = 8 gpm/ft²
Q = 6 gpm/ft²
Headloss
Volume of filtrate

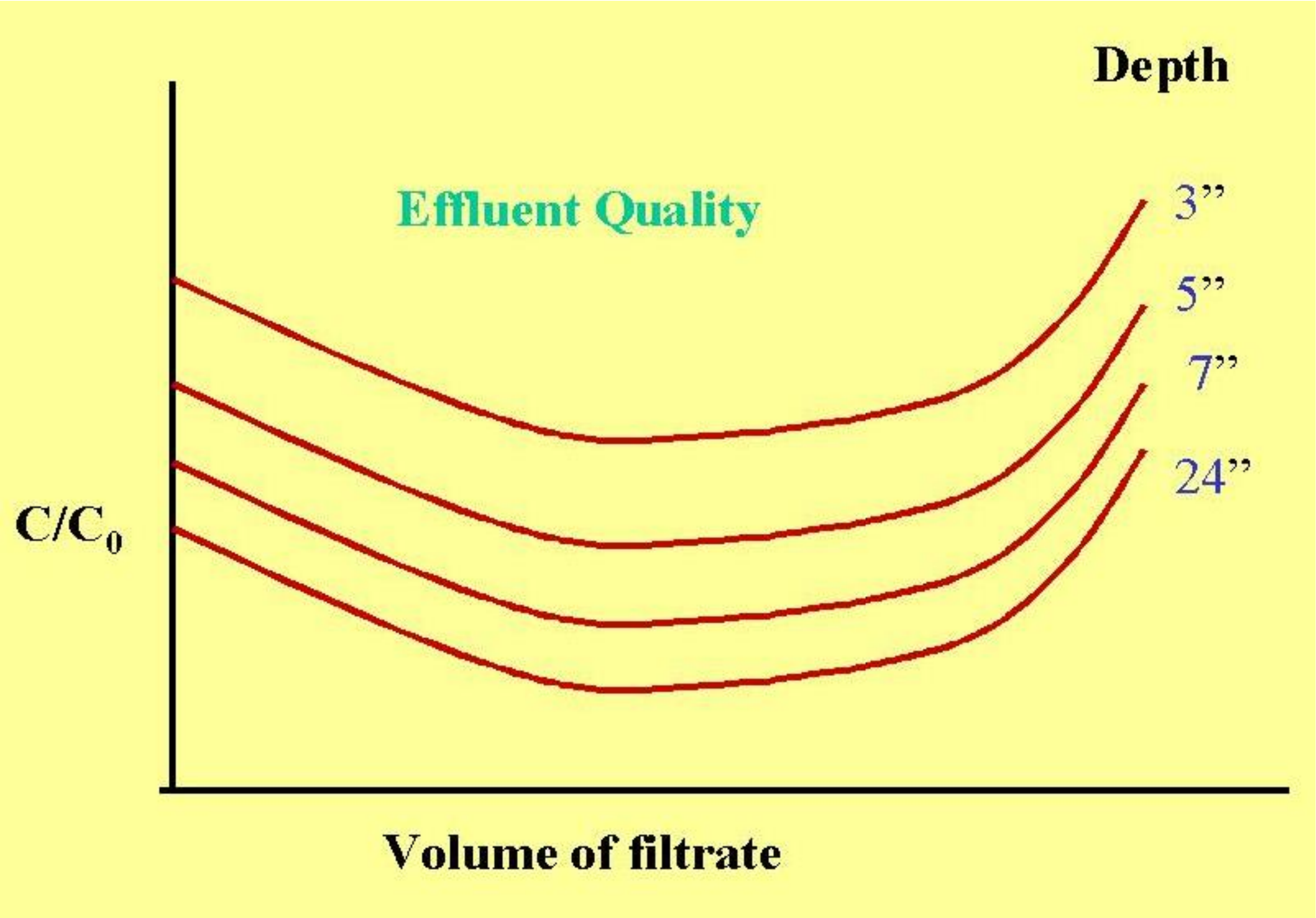
At lower surface loading rates surface removal is significant because the velocity is not high enough to drive the SS into the media (At higher loading rates the suspended solids are driven into the media). The compressibility of the surface layer results in higher headloss at higher velocity.


gpm/ft²
2
3
4
6
5
Headloss
Volume of filtrate

Reverse graded filters are used to enhance depth removal and reduce surface removal. A more uniform solids distribution results. Thus longer filter runs can be attained. Filter runs of 2-5 times longer than single media filters are attainable. Head loss patterns for a reverse graded filter are shown in this figure.


Reverse Graded Filter
Headloss
Q = 6 gpm/ft²
4
2
Volume of filtrate

Effluent Quality (turbidity) patterns for various depths in a granular filter are shown here:



Effluent quality at any layer tends to improve initially, then get worse with time or throughput. As SS are removed by adsorption and straining the media surface area increases (giving better adsorption/attachment) and the void spaces become smaller (giving better straining). As the channels become smaller interstitial velocity increases and we get greater shear which results in sloughing to lower layers of media.

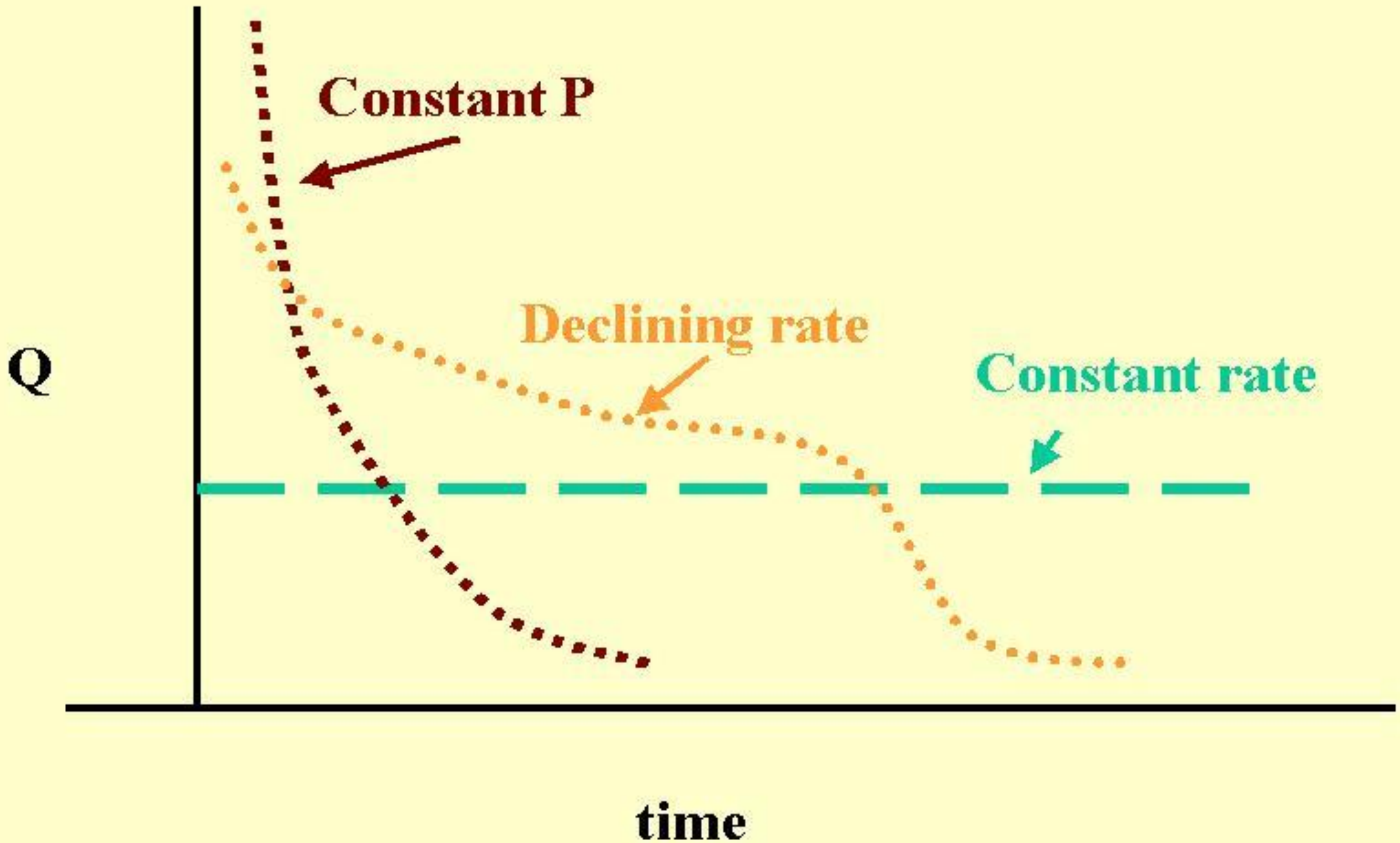
DESIGN OF GRANULAR FILTERS

Modes of Flow Control

Possibilities:

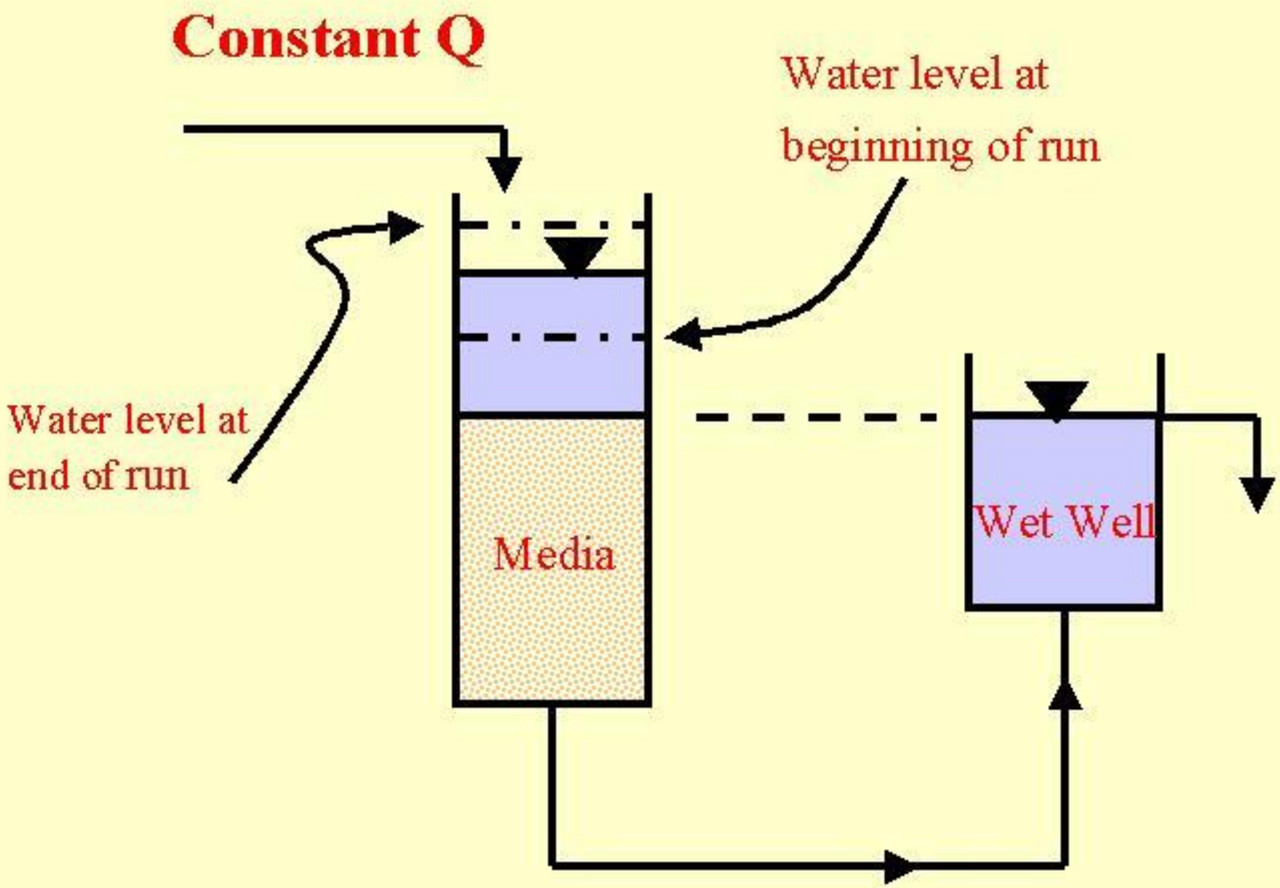
- **Constant pressure,**
- **constant rate,**
- **declining rate**

Only constant rate and declining rate are reasonable from economic, practical point of view. So these are the only two we will discuss.


Constant P
Declining rate
Constant rate
Q
time

Constant rate filtration

In this mode of operation a constant flow rate (Q) is applied to the filter. This constant Q is usually controlled by a system of weirs. This usually requires a wet well (storage) if we are treating a wastewater. As the filter run proceeds and head loss increases, water level in the filters increases to compensate for greater head requirement. A schematic of a constant flow filter is shown here.


Constant Q
Water level at
beginning of run
Water level at
end of run
Media
Wet Well

Declining rate filtration

In this scheme the filters are operated in parallel with common influent header. 4 parallel filters are operated so that one filter is down and being backwashed and the other filters take up the slack. When one filter is down the flow increases to the other 3. The head in each of the other three increases somewhat to force more flow (to accommodate the extra flow from the down filter).

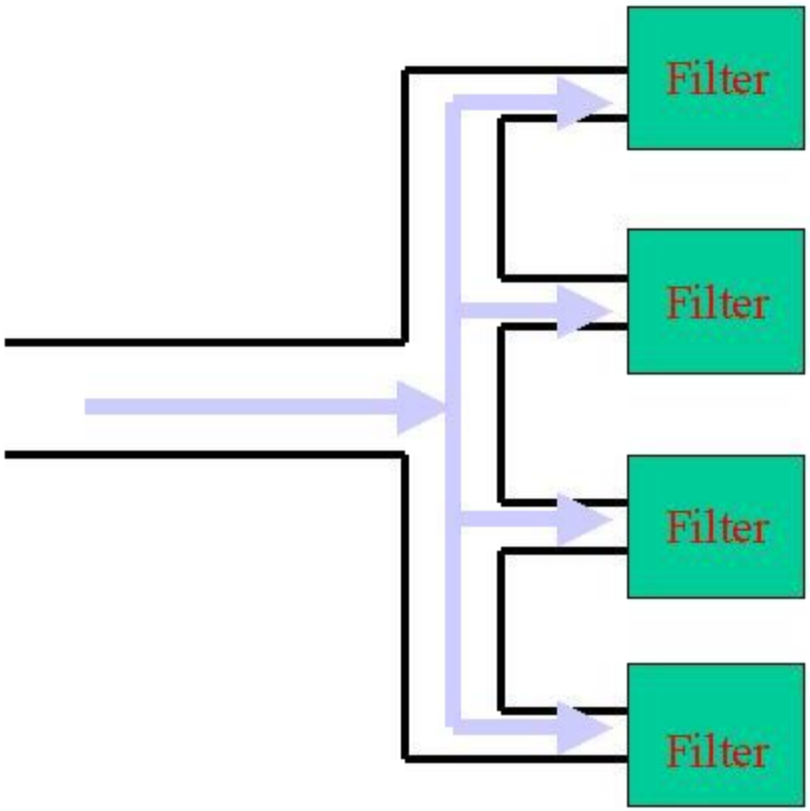
Distributing the flow across all the filters evens out the cycle and produced a declining rate (but a gradual declining rate). Net result is that headloss is the same in all filters but Q is not. In fact the individual filter rate declines gradually.

Advantages:

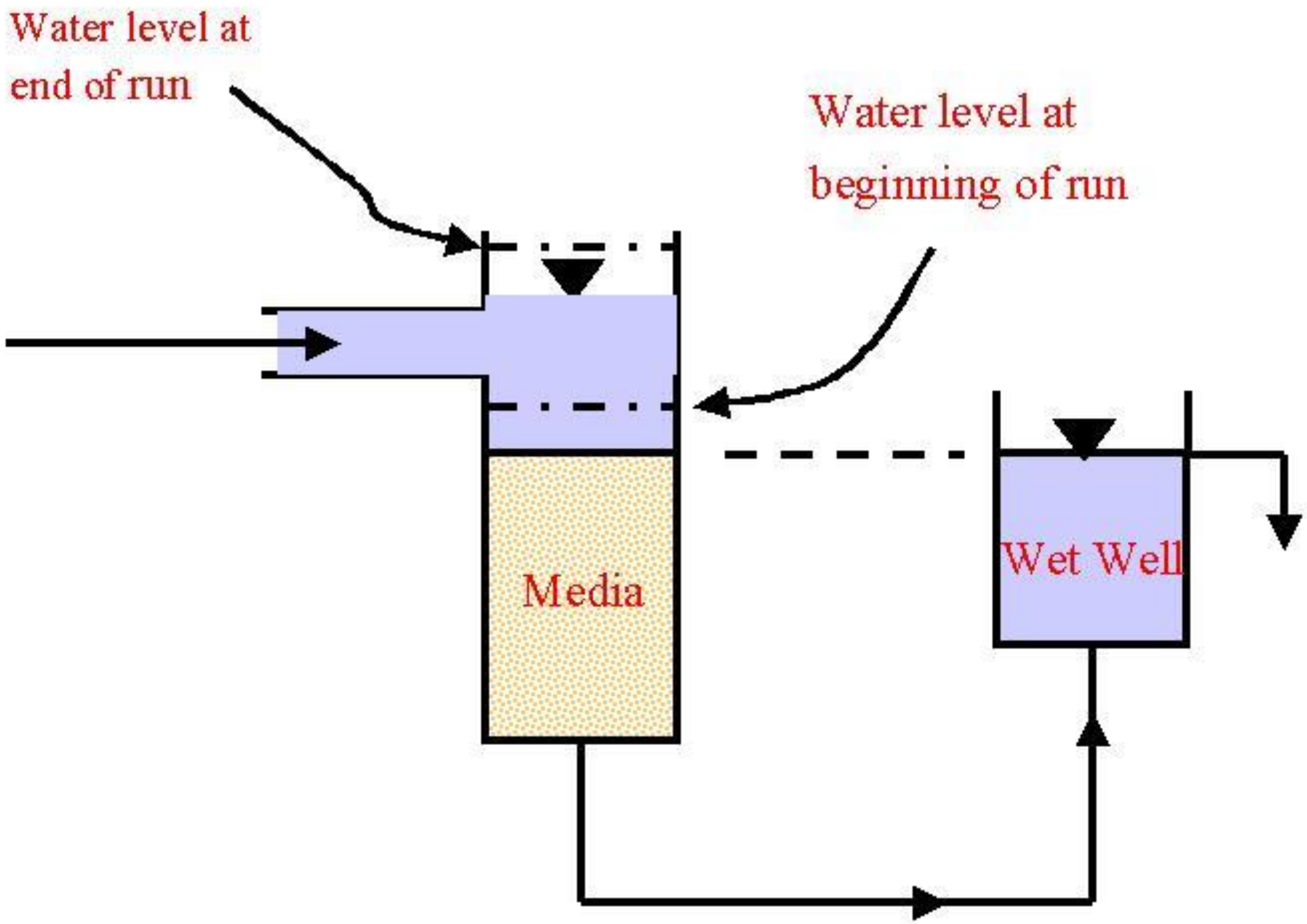
•· Better filtrate quality since we don't try to force high velocity through a clogged filter as in the constant rate system.

•· Lower headloss in influent since there are no weir losses involved.

Declining Rate Filters



Declining rate filter



Filter Configuration

Downflow filters are the most conventional. In the upflow mode the filter needs a grid on surface to prevent sand from flowing out of the top. One advantage of upflow filter is that we can take advantage of the reverse grading of the filter bed (coarser particles at bottom or influent side). In the downflow mode sand is usually naturally graded in the opposite direction relative to flow.

Reverse grade helps to extend filter run. Another disadvantage of the upflow mode is that hydraulic perturbations can lift the bed allowing suspended solids to escape. Usually the depth of upflow bed is 6 -10 ft. as compared to 1 to 3 ft in the downflow mode.

Filter media

Important considerations in selecting media:

- • too fine - surface straining which results in high head loss and short filter runs.

- • too coarse - poor filtrate quality , high backwash flow required.

Single media:

Sand :

•24"-30" depth

•Effective size = 0.4-1.0 mm.

•Uniformity coefficient < 1.65

•Density = 2.65.

•Porosity = 0.43

Dual media:

To compensate for the unfavorable gradation that occurs in the single media filters we can use dual media (reverse graded) filters. Place a less dense, larger diameter media on top of sand. This results in a higher porosity (0.55) at top of filter. Sand has porosity of about 0.4. Lower density also allows the less dense media to remain on top after backwashing.

Media	**Depth (in)**	**Eff Size (mm)**	**Uniformity Coefficient**
Anthacite	12 – 20	0.9 – 1.0	< 1.8
Sand	12 – 16	0.5 – 0.55	<1.65

This combination will allow about 6" of intermixing so that there will not be an accumulation of suspended solids at a sudden porosity change interface.

This will also allow fluidization (backwash) for both layers at approximately the same Q.

Trimedia Filters

Garnet sand (density = 4-4.2 g/cc) is some times used below the sand in a third layer. Difficulties arise in keeping the sand and garnet from intermixing during backwash. In general this extra layer not worth the extra trouble.

Filtration rate

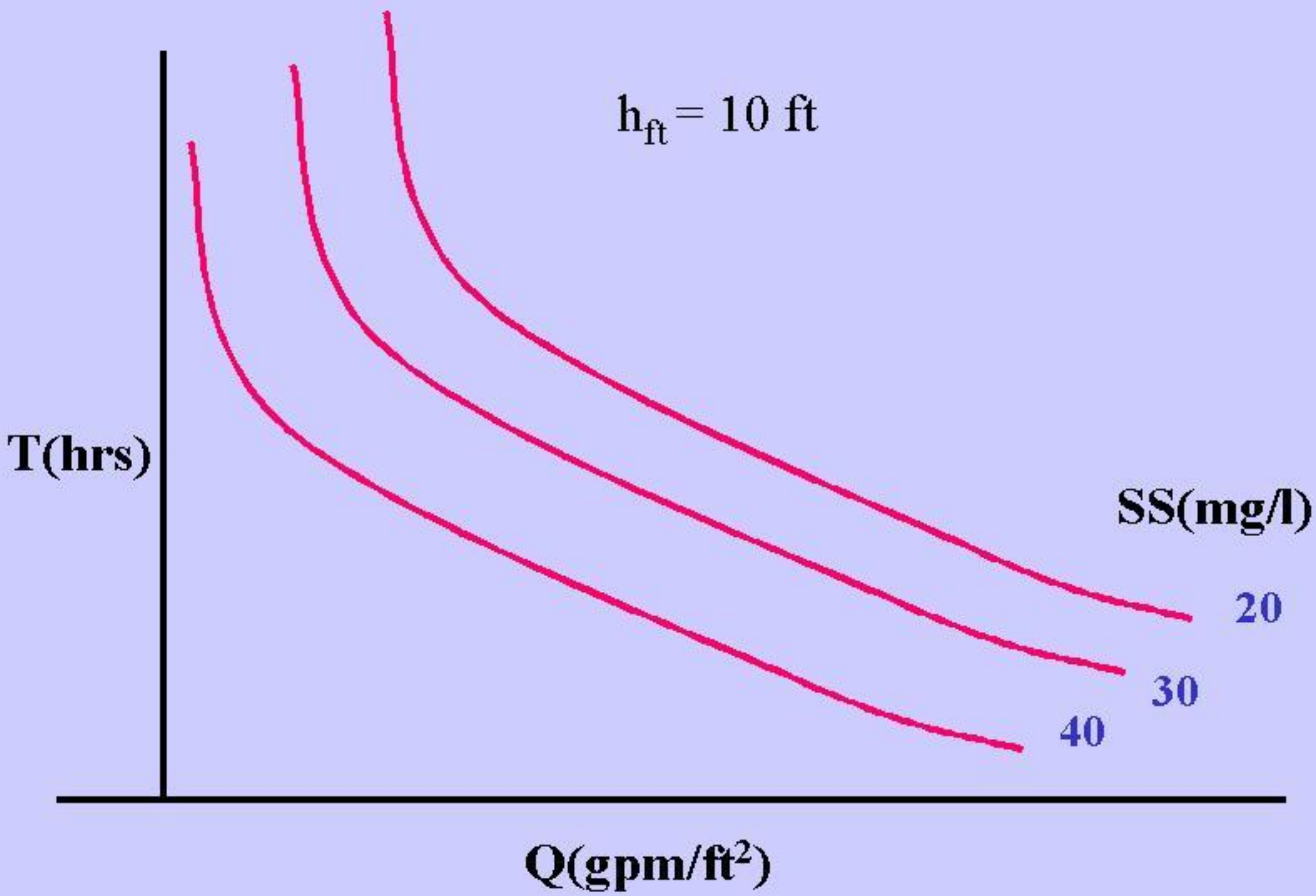
- 1-8 gpm/ft^2 = the acceptable range
- 2-3 gpm/ft^2 = average flow loading rates
- 4-5 gpm/ft^2 = peak flow loading rate

Terminal headloss

•Commonly 3 - 5 ft for water treatment

•Up to 10 ft for wastewater treatment (biological floc can tolerate more shear force than chemical floc without breaking up)

•Filter run = T = f(floc strength, Q and suspended solids concentration in influent).

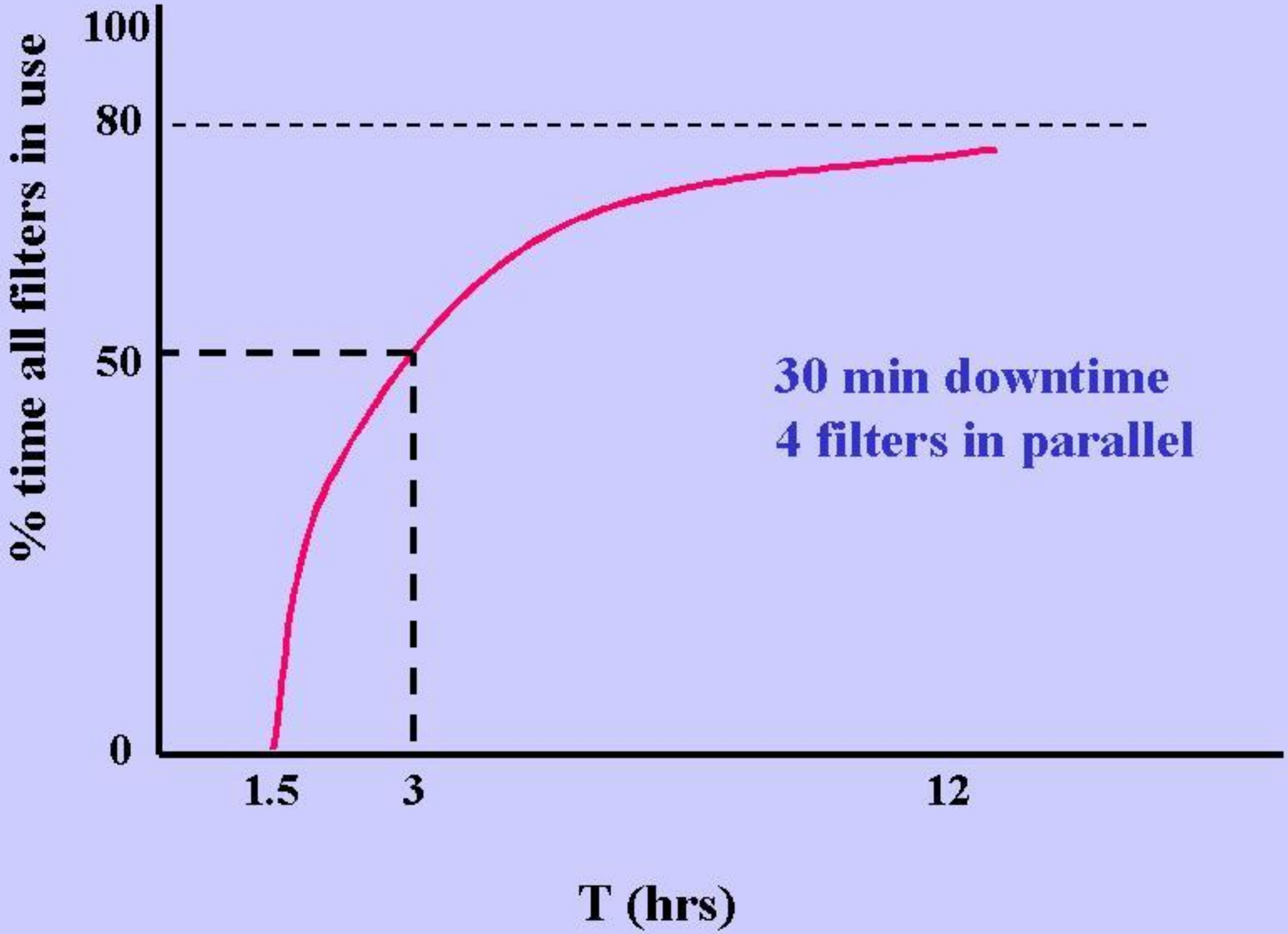


h_{ft} = 10 ft
T(hrs)
SS(mg/l)
20
30
40
Q(gpm/ft^2)

It turns out that the optimum T for filters is between 12 and 30 hrs at least for water treatment where the primary objective is water production. This is explained as follows.

If $T<12$ hours all of the filters will not be used simultaneously enough of the time. This results in overloading the on-line filters for a higher percentage of the time.

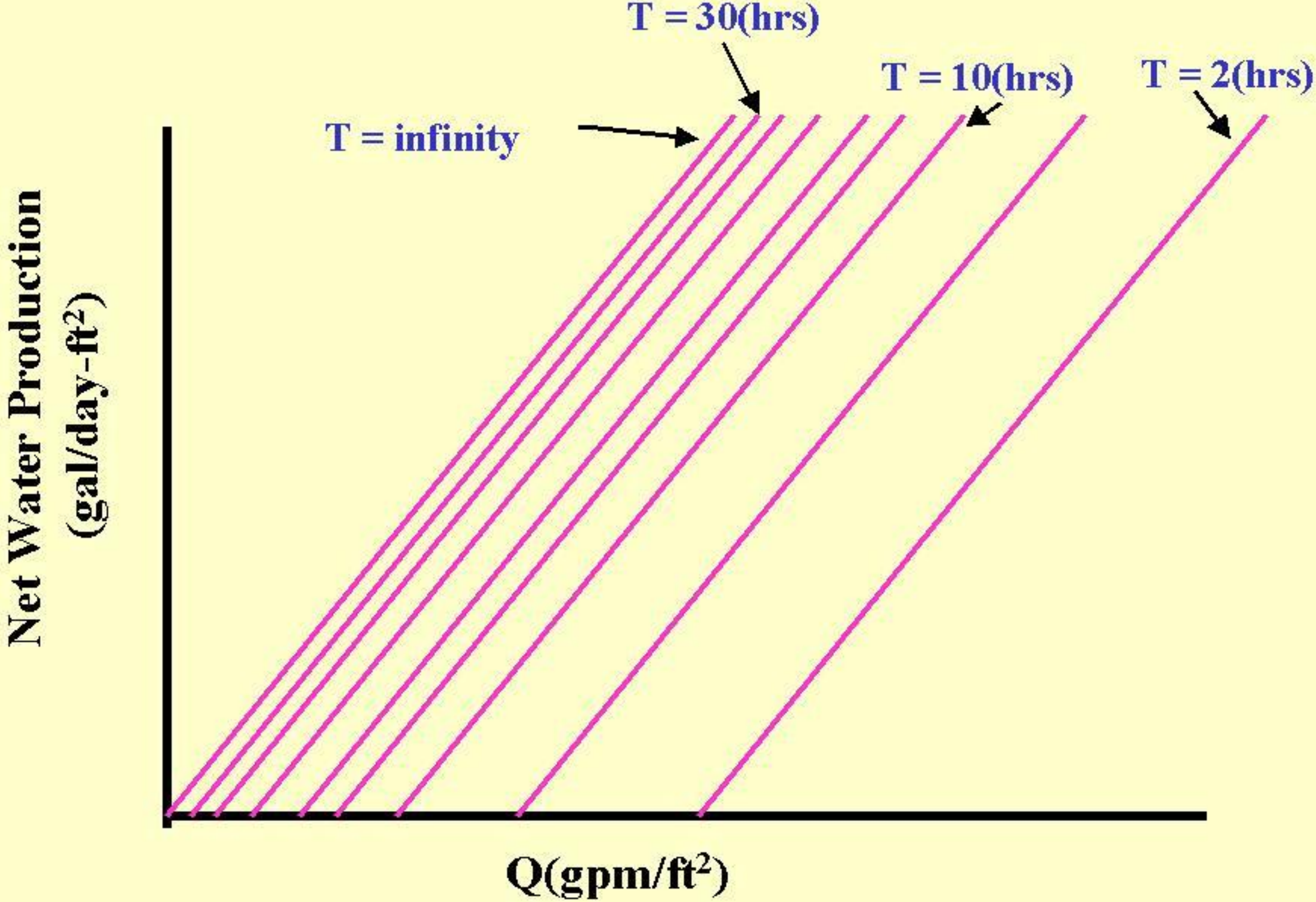
Filters can be "overloaded" for short periods of time but overloading for extended periods creates the requirement that the filter size should be increased. Where does the 12-hour lower limit come from? The following figure helps explain.


% time all filters in use
100
80
50
0
1.5
3
12
T (hrs)
30 min downtime
4 filters in parallel

A plot of % time all filters are in use versus T shows that the % that all filters are in use asymptotically approaches about 80% @ about T = 12 hrs.   This is based on 30 minute downtime for backwashing and 4 filters operated in parallel. Note that T cannot be less than 1.5 hours for 4 filters if only 1 filter is to be backwashed at a time. So anytime that backwashing is going on the on-line filters are carrying 1.33 of the design flow. In the case of T = 12 hrs, this overload is only going on for 20% of the time (an acceptable scenario).

Of course, if T > 12 hrs the filters will be carrying the overload for even less of the time. So why not operate at longer T? One of the reasons is that head loss from dirty filters has to be provided for. (Another consideration is the amount of clean backwash water required for each T.) ← minor

Net water production (per cycle) = (forward flow rate x T) - (backwash flow rate x backwash time). Assuming 30 min down time which includes air scour plus 5 mins. of 20 gpm/ft^2 backwash we can plot the following figure.


T = 30(hrs)
T = 10(hrs)
T = 2(hrs)
T = infinity
Net Water Production (gal/day-ft²)
Q(gpm/ft²)

In this figure @ T = infinity applied filtrate = produced filtrate. For T > 30 hours there is very little advantage in terms of net production of water. So there is no reason to go beyond this T and, in fact, we only encounter more headloss for little gain in productivity . So we may as well stop at this point.

Backwash requirements

When terminal head loss is reached the filters must be backwashed with clear water. Usually this clear water comes from the wet well that follows the filter. For downflow filters backwashing is done by fluidizing the bed in an upflow mode. Wash water is collected at the top of the filters in wash gutters and either sent back to the head end of the plant or to the sludge treatment train.

Backwash sequences

Bed expansion is between 15-30% accomplished by applying a backflow rate of about 15 gpm/ft^2 for 5 - 10 mins. Hydrodynamic shear cleans the media particles (attached, as well as strained). Optimum shearing occurs at about 50 % expansion but this tends to require excessive backwash velocities with coarser media particles and these high flow backwashs could fluidize the gravel underdrain.

Surface wash: Surface wash water is sometimes pumped at high velocity 1-2" above unexpanded bed. Surface wash can be used prior to and/or during expansion.

Air scour: air introduced just above gravel underdrain , 3-5 min prior to backwashing. (3- 5 scf/ft^2 of air).

Hydraulics of backwash bed expansion

We need to know velocity and flow rate necessary to fluidize bed so that pumps and wet wells can be designed appropriately.

Mathematical description of bed expansion:

$$\frac{D_e}{D} = \frac{(1-\varepsilon)}{(1-\bar{\varepsilon})}$$

$$\bar{\varepsilon} = 1 - \frac{D}{D_e}(1 - \varepsilon)$$

ε, $\bar{\varepsilon}$ are, respectively, the unexpanded and expanded porosity of a clean bed.

D, D_e are the unexpanded and expanded depth of the media.

An empirical observation relates the required approach velocity to the extent of fluidization:

$$V = K_e (\bar{\varepsilon})^{n_e}$$

Where V is the approach velocity required to attain a certain level of bed expansion.

n_e and K_e are constants that can be evaluated by a settling analysis of the media.

The following is an empirical expression that relates minimum fluidization velocity to media particle settling velocity.

$$V_s = 8.45 \cdot V_f$$

(units don't matter as long as they are consistent)

V_s = settling velocity of the media

V_f = <u>minimum</u> fluidization velocity of the media.

Another empirical observation:

$$n_e = 4.45\,\mathrm{Re}_0^{-0.1} \quad \text{(dimensionless)}$$

$$\mathrm{Re}_0 = \frac{\rho_l \cdot V_s \cdot d_{60}}{\mu}$$

or:

$$\mathrm{Re}_0 = \frac{\rho_l \cdot 8.45 \cdot V_f \cdot d_{60}}{\mu}$$

$$\mathrm{Re}_0 = 8.45 \cdot \mathrm{Re}_f$$

$$n_e = 4.45(8.45 \cdot \mathrm{Re}_f)^{-0.1}$$

$$= 3.59 \cdot \mathrm{Re}_f^{-0.1}$$

V_f can alternatively be computed using another empirical relationship:

$$V_f = \frac{0.00381(d_{60})^{1.82}\{\omega_s(\omega_m - \omega_s)\}^{0.94}}{\mu^{0.88}}$$

(be careful to use the proper units since this is an empirical relationship)

$\omega_{s,m}$ = specific weight of water, media (lb/ft^3)

μ = viscosity of water (centipoise)

d_{60} has units of millimeter.

V_f = min. fluidization velocity in gpm/ft^2

If $Re_f > 10$ then we need to apply a correction factor:

$$K_R = 1.775\, Re_f^{-0.272}$$

$$V_f' = K_R \cdot V_f$$

Then use: $$n_e = 3.59(Re_f')^{-0.1}$$

Use initial porosity and the empirical equation for fluidization velocity to compute $K_{e.}$

$$K_e = \frac{V_f}{\varepsilon^{n_e}}$$

Now the fluidization velocity for any bed expansion can be calculated using :

$$V = K_e (\bar{\varepsilon})^{n_e}$$

Pressure drop through the expanded bed is equal to the buoyant wt. of the bed (no expansion dependency):

$$\Delta p = \frac{D(1-\varepsilon)(\omega_m - \omega_s)}{62.4}$$

$$= D(1-\varepsilon)(sp.gr._{media} - sp.gr._{H_2O})$$

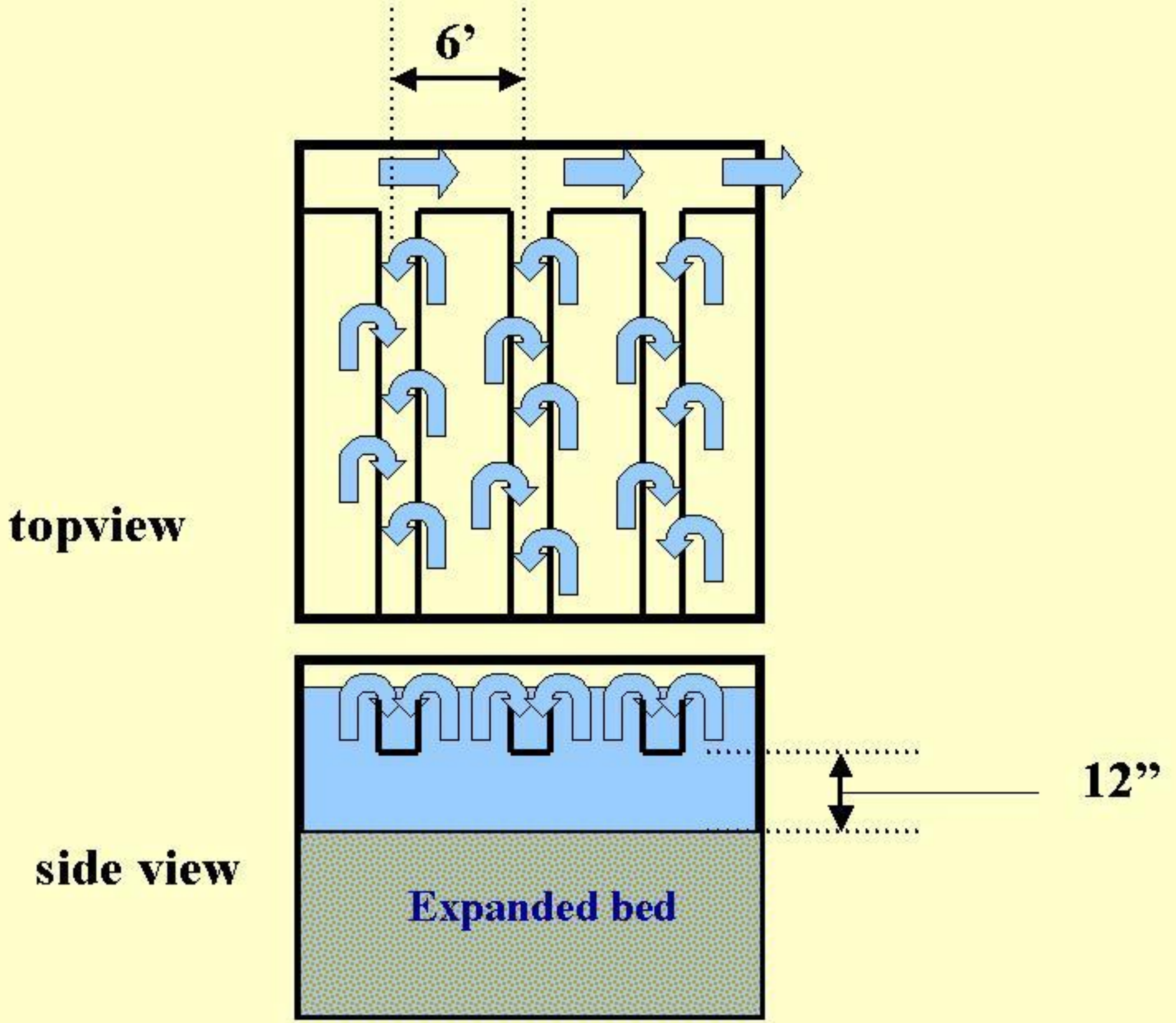
For multimedia beds apply expansion and pressure drop equations separately to each media layer.

Underdrain and washwater gutter design

Washwater gutter design

Washwater gutters carry away the backwash water that is laden with suspended solids. These gutters are located so that horizontal travel of suspended solids is less than 3 feet. This will assure capture of most of the released solids. This translates to maximum horizontal spacing of about 6' between gutters.

Gutters are located about 12" above top of expanded bed. This minimizes the amount of dirty water left in filter box and it also minimizes possible media loss.


6'
topview
side view
12"
Expanded bed

Sizing rectangular gutters

Flow capacity of rectangular gutter

$$Q = 2.5W \cdot (D_u)^{1.5}$$

W = width of gutter (ft)
D_u = depth of water in channel (ft)
Q in cfs

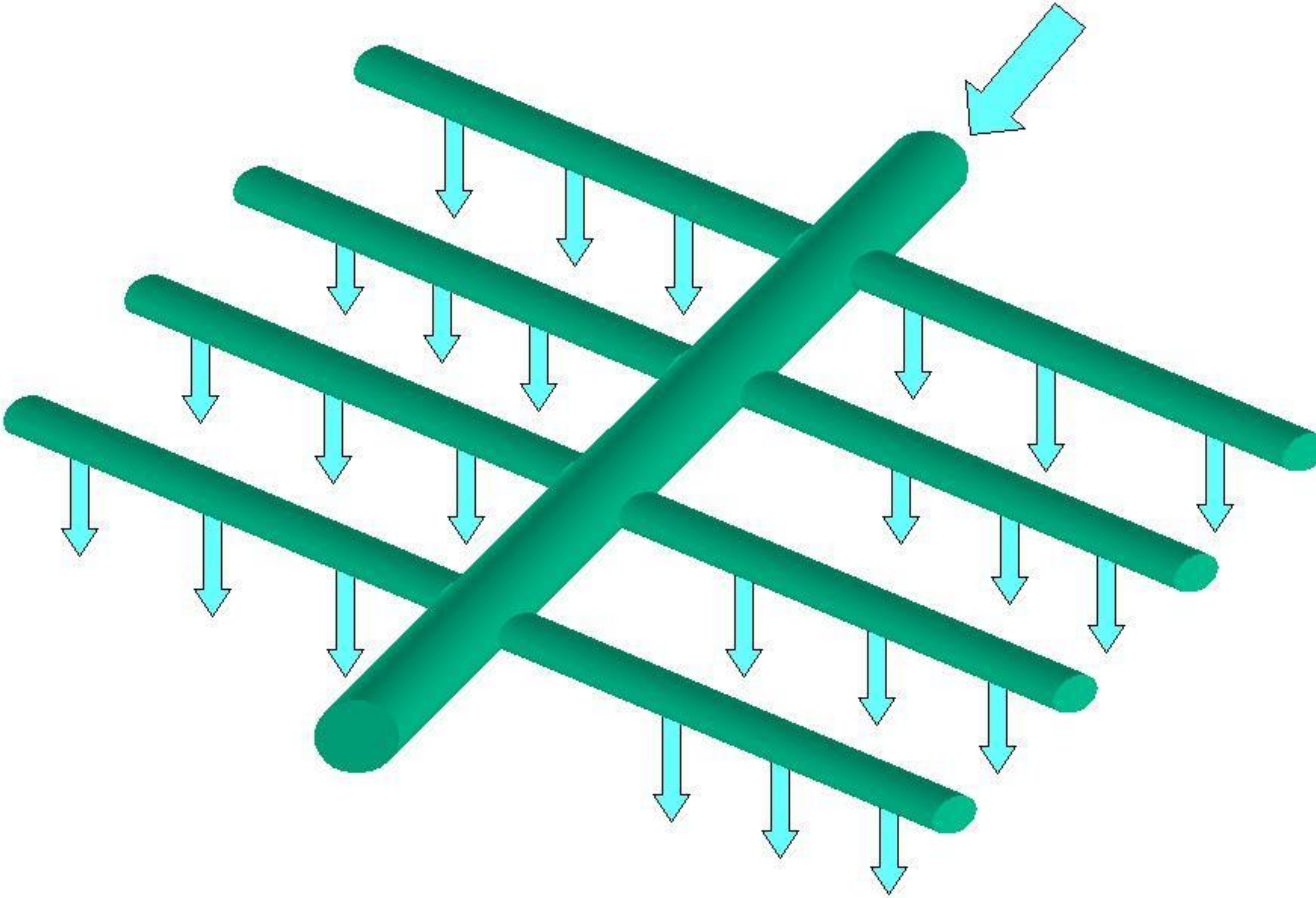
Actual design depth D = D_u + 2-3" freeboard.

Underdrain design

Purpose of the underdrain :

1) support media
2) evenly distribute backwash
3) collect filtrate

It is common to have a manifold-lateral system beneath the gravel as shown below.



Laterals are perforated on the bottom with holes located 3 - 12" apart. Hole size = 1/4 - 1/2". To get even distribution of flow orifice head loss made purposely high relative to head loss through the laterals. Common orifice head loss is about 15 ft.

Total head required for backwash is then sum of:

1) orifices

2) expanded bed

3) flow through gravel

4) manifold and lateral (minor)

5) elevation difference between backwash supply and wash gutters.

Backwash water is provided by an elevated tank or pump.

SLOW SAND FILTERS

Slow sand filters (as opposed to "rapid sand filters", the type discussed above) are operated at a much lower loading rate. Surface filtration is promoted in these filters because of the lower loading rates and because the effective size of the sand is smaller than that of the rapid sand filters. Effective size for these filters is 0.35 mm (uniformity coeff. = 1.75) as compared to 0.4 - 1.0 mm for rapid sand filters.

- Head applied above sand: 3-5 ft.
- Depth of sand is also about 3- 5 ft.
- Loading rates: 0.05 - 0.1 gpm/ft^2
- T: 1-6 months

No backwashing is employed with these filters, instead the upper 1 - 2" of sand is periodically scraped off and removed with periodic addition of new sand .

Removal mechanism primarily by filter cake on the surface of the sand. This layer is called a "schmutzedecke". The schmutzedecke is composed of inorganic and biological material therefore removal is by straining, adsorption and bioxidation.

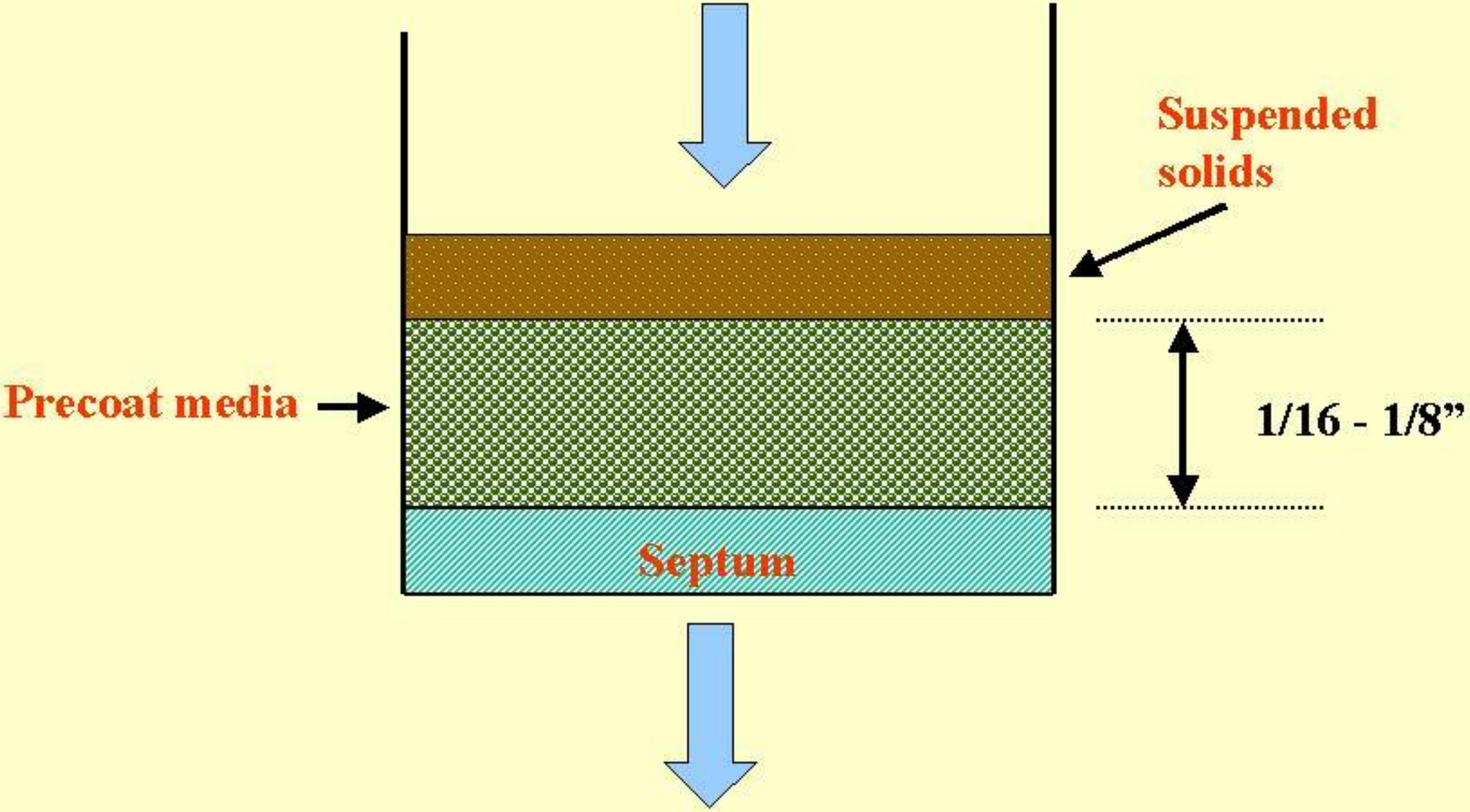
Advantage: No backwash requirements, good removal.

Disadvantage: Need large surface area because of the low hydraulic loading rate.

Precoat Filters

Description:

The filtration media is hydraulically deposited on a septum. The filtration media is usually perlite (siliceous volcanic rock), activated carbon or diatomaceous earth (siliceous exoskeletons of algae and diatoms).



Suspended
solids
Precoat media
1/16 - 1/8”
Septum

Mechanism of SS Removal

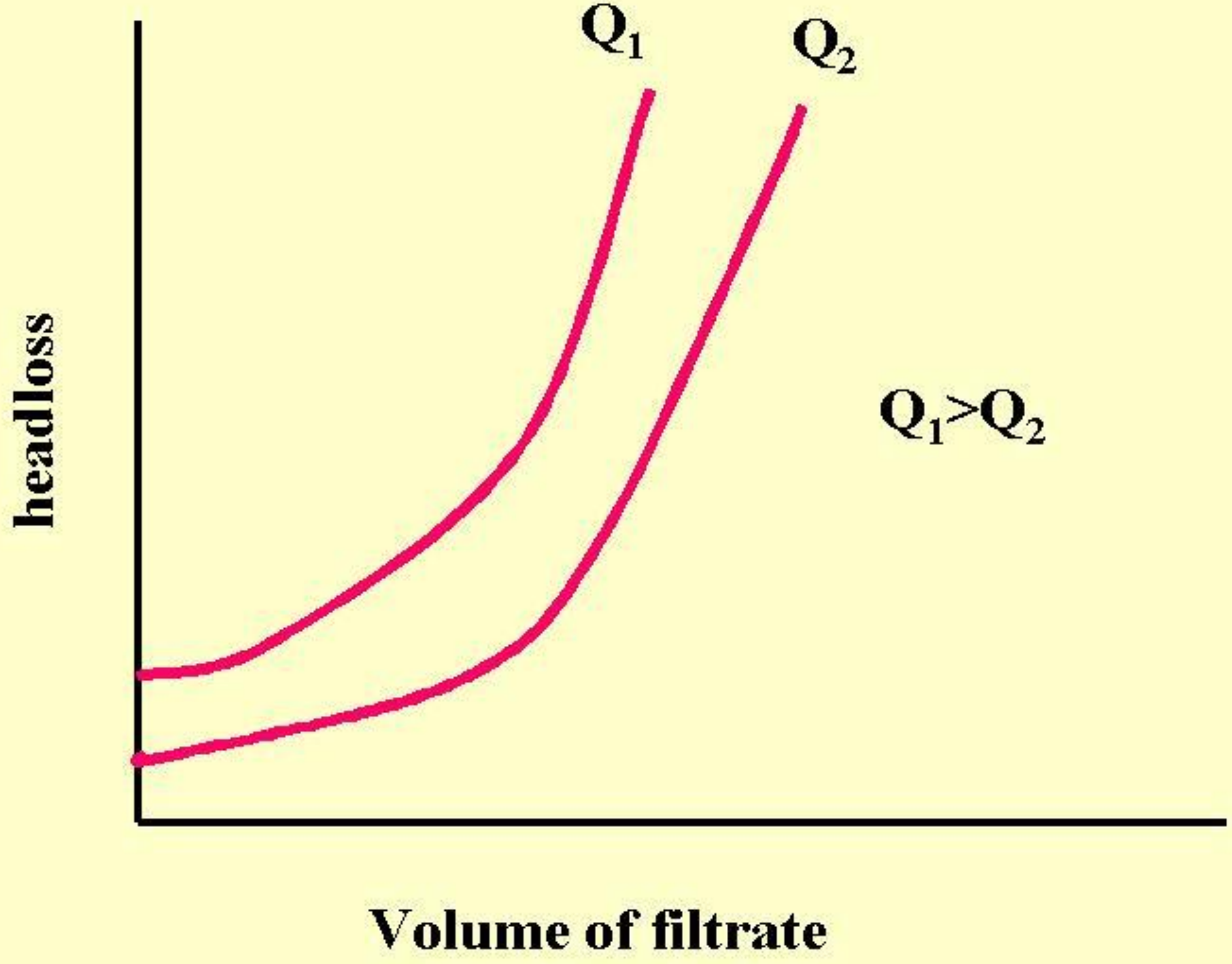
Removal is primarily by mechanical straining by the cake of suspended solids that builds up on the precoat. In other words "surface" filtration is the major type of activity.

Pretreatment

Pretreatment for precoat filtration usually involves adding a “body feed” to the feed stream. Body feed is simply material of the same composition as precoat material itself. The objective of the body feed is to minimize compressibility of the surface cake and to fill in any accidental holes where the precoat has not covered the septum. It is presumed that the body feed is incompressible. Body feed/conc. of SS (influent) = 3-6.

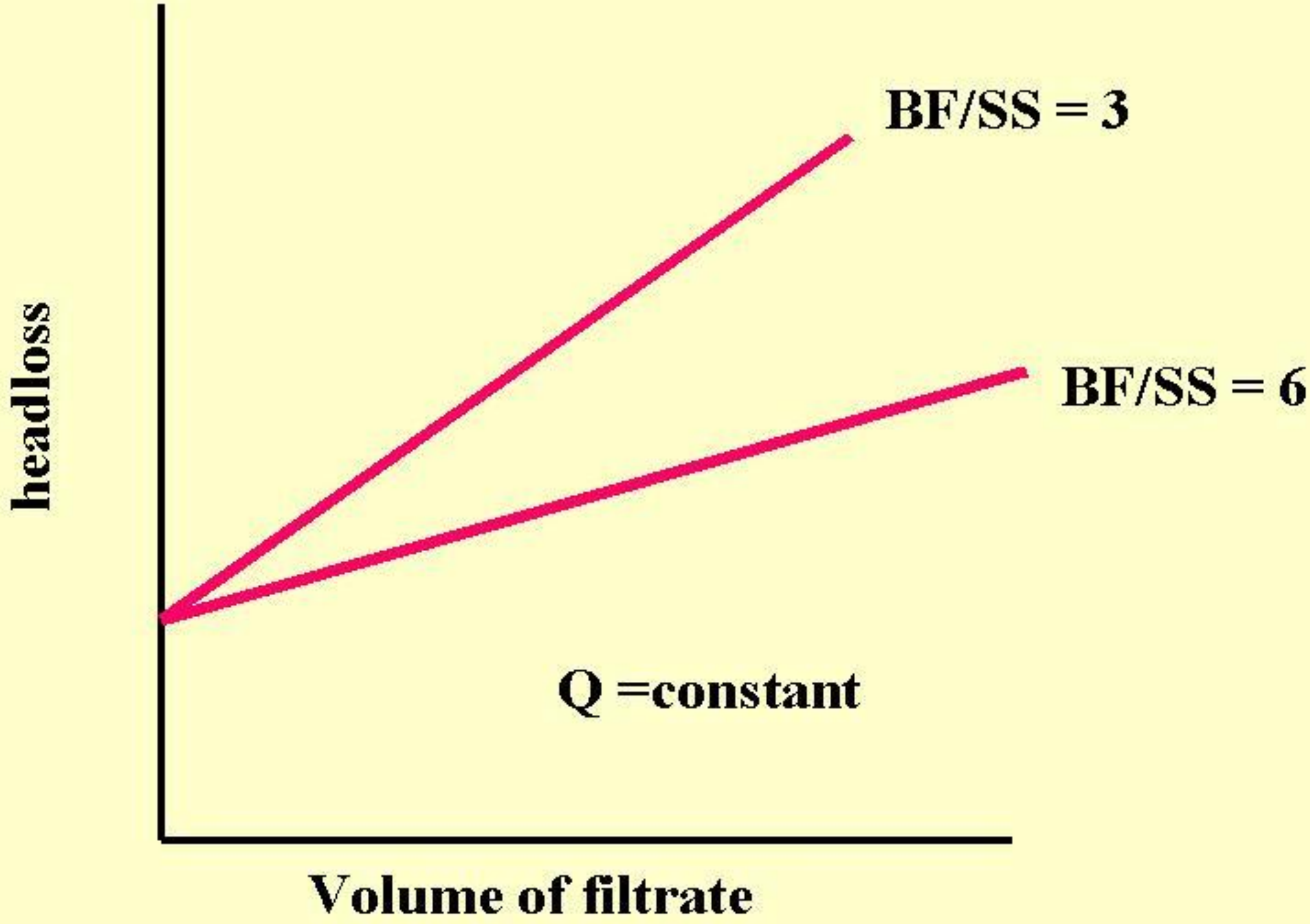
Headloss patterns

Head loss versus Q depends on whether body feed has been added. W/O body feed head-loss varies very non-linearly with Q since compressibility is a big factor. Body feed addition helps reduce this unfavorable non-linearity.



Q_1
Q_2
$Q_1 > Q_2$
headloss
Volume of filtrate

w/ body feed



DESIGN OF PRECOAT FILTERS

Filter cycle:

- application of precoat: 0.1 - 0.2 lbs/ft^2 (1/16-1/8" thickness). This is applied at a rate of about 1 gpm/ft^2. Application requires about 3-5 minutes. Initial head loss is about 0.5 to 1.5 ft. This high head loss gives some idea how tight the precoat porosity is.

- Filtration of water plus body feed:

Q = 0.5- 2.5 gpm/ft

T = 24 hrs.

Optimum body feed dosage equals that which gives linear headloss vs. filtrate volume

Optimum terminal headloss = 75-150 ft.