Chapter 12

Activated Sludge Processes

12.1 INTRODUCTION

The activated sludge process is generally considered to have had its origins in aeration experiments carried out by Ardern and Lockett in Manchester in 1914 (IWPC, 1987). They found that, when they retained and built up in an aerated vessel the biological floc formed in a series of aeration experiments on sewage, the time required for its purification progressively decreased as the concentration of floc increased. They referred to this floc as being 'activated' and the resulting process became known as the activated sludge process. In the intervening period this process has become one of the main methods, used worldwide for the purification of wastewaters containing biodegradable organic solids, its most important application being the treatment of domestic sewage.

12.2 PROCESS CONFIGURATION

The principal unit in all activated sludge processes is the aeration or biochemical reaction vessel. In it the waste organics are mixed with the active sludge and this so-called 'mixed liquor' is aerated for several hours, during which the microorganisms (mainly bacteria) in the sludge utilise the organic matter in the waste for energy and cell synthesis. In the conventional continuous flow process configuration, as illustrated in Fig 12.1, the activated sludge biomass is separated from the mixed liquor in a secondary settling tank (SST) and the separated sludge is recycled back to the aeration tank. A proportion of the settled sludge may be removed from the system on a continuous or intermittent basis to maintain the concentration of active biomass in the aeration unit within the desired value range. The activated sludge suspension in the aeration unit is commonly referred to as 'mixed liquor'.
The operating mixed liquor suspended solids (MLSS) concentration may vary in the range 1500-5000 mg l^{-1}. The active fraction of the MLSS is conveniently approximated for design purposes as its organic component or mixed liquor volatile suspended solids (MLVSS). The MLVSS:MLSS ratio is typically within the range 0.65-0.85.

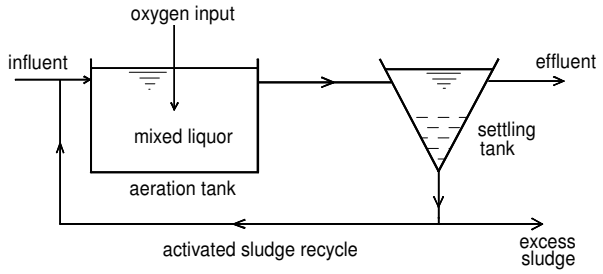

Fig 12.1 Schematic layout of the activated sludge process

The process can also be operated in batch mode, thereby eliminating the need for an adjunct sedimentation tank. Sequencing batch reactors (SBRs) are described in section 12.10.

12.3 PROCESS SELECTION

The category of AS process selected for a given application is determined by the process outcome required, as specified by effluent and waste sludge quality parameters. Activated sludge processes are generally designed to achieve one or more of the following process objectives:

(a) carbonaceous BOD removal only (non-nitrifying AS process)
(b) enhanced nitrogen removal
(c) enhanced phosphorus removal
(d) stabilised activated sludge biomass

Non-nitrifying or carbonaceous BOD removal processes are appropriate to circumstances where the required effluent quality is specified in terms of BOD_5 and suspended solids (SS) concentration only. For example, the EU Urban Wastewater Directive (2000/60/EC) specifies the following limit values for treated effluent discharge to non-sensitive waters:

Parameter	Limit value (mg l-1)
BOD_5	25
COD	125
SS	35

A limited degree of nitrogen removal is achieved in non-nitrifying AS processes through the incorporation of nitrogen in microbial cells and in non-biodegradable solid matter. This removal might constitute some 20-25% of the total nitrogen in typical municipal wastewater. Enhanced nitrogen removal is achieved by AS processes, which combine nitrification and denitrification stages with simultaneous BOD removal.

Similarly, a limited degree of phosphorus removal is effected by conventional AS processes. Enhanced biological phosphorus removal can be achieved by the incorporation of an anaerobic stage in the AS process, as discussed later in this chapter.

Where a stable waste sludge is a required process outcome, an extended aeration process in which the sludge biomass is retained in the reactor for a long period (<20d), is necessary. Such an extended aeration period will also reduce, through bio-oxidation, the quantity of surplus produced.

12.4 PROCESS DESIGN

As shown in Chapter 11, the microbial solids residence, θ_s, is a key performance parameter in biological reactors. The corresponding process parameter in an AS process is the solids residence time, SRT. SRT and θ_s have the same value where the active microbial fraction of the MLSS remains constant. The mean SRT for an AS process is given by the ratio of the sludge mass in the aeration basin to the sludge growth rate:

$$SRT = \frac{V.MLSS}{S_y.Q.BOD_5} \qquad (12.4)$$

where Q is the influent flow rate ($m^3\ d^{-1}$), BOD_5 is the average influent concentration (kg m^{-3}), S_y is the sludge yield (kg per kg BOD_5). In continuous flow processes the actual overall sludge residence time is somewhat greater than given by the foregoing measure of SRT by virtue of the fact that there is an additional mass of solids in transit between the aeration tank and the SST.

For practical design purposes, the AS yield can be empirically related (ATV, 2000) to the influent concentrations of BOD_5 and SS, the sludge SRT and the process temperature by the following empirical correlation:

$$S_y = 0.75 + 0.6\frac{SS}{BOD_5} - S_o \qquad (12.5)$$

where SS is the influent suspended solids concentration and S_o is the solids oxidation factor, which is a function of both sludge SRT and process temperature:

$$S_o = \frac{0.102\ SRT \cdot F_T}{1 + 0.17\ SRT \cdot F_T} \qquad (12.6)$$

The factor F_T accounts for the influence of the process temperature:

$$F_T = 1.072^{(T-15)} \qquad (12.7)$$

where T is the process temperature (^oC).

The heterotrophic microorganisms responsible for carbonaceous BOD removal dominate the microbial population in both non-nitrifying and nitrifying activated sludge processes, hence the foregoing expressions can be applied over the full AS process SRT range. The growth of the autotrophic nitrifying organisms makes a negligible mass contribution to the process MLSS in nitrifying AS processes.

The influences of influent suspended solids and sludge SRT and process temperature on sludge yield are shown graphically in Fig 12.2.

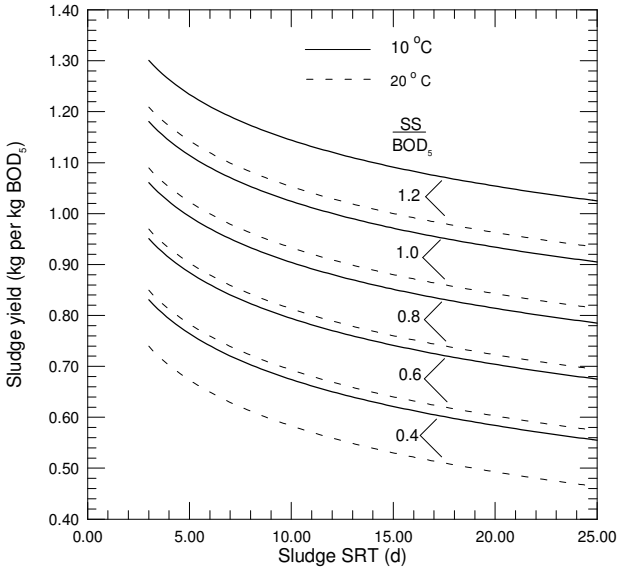

Fig 12.2 **Influence of process parameters on AS process sludge yield**
(equations 12.4, 12.5, 12.6 and 12.7)

While SRT is the most appropriate process design parameter, the sludge loading rate (SLR) is also widely used in design practice. These two parameters are correlated by equation (12.4):

$$SLR = \frac{Q \cdot BOD5}{V \cdot MLSS} = \frac{1}{S_y \cdot SRT} \qquad (12.8)$$

The correlation of SLR and SRT for a typical municipal wastewater ($SS/BOD_5 \cong 1.2$), with and without upstream primary sedimentation, is shown graphically in Fig 12.3. The plots show the impact of primary sedimentation through its removal of SS and BOD_5, both of which influence the sludge yield.

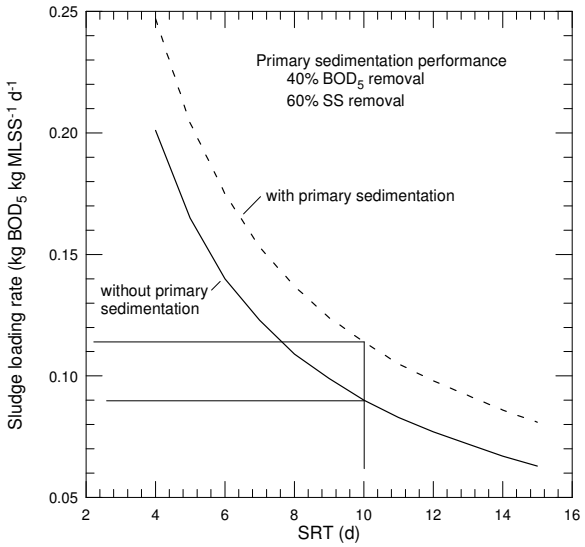

Fig 12.3 Correlation of SLR and SRT parameters

12.4.1 Carbonaceous BOD removal

Process design for carbonaceous BOD removal only requires a non-nitrifying process environment, which, in temperate climates (process operating temperature range 10-20 ^oC), is typically achieved by regulation of the process SRT at ≤ 5 days. Standard-rate non-nitrifying AS processes are conventionally designed to operate within the SRT range 3-5 days. The lower end of this SRT range is comfortably within the zone of stable process operation, where the rate of reduction in effluent substrate concentration with increasing SRT is relatively low. Standard-rate AS processes have been widely applied to achieve the so-called '20/30' effluent standard i.e. $BOD_5 \leq 20$ mg l^{-1} and SS ≤ 30 mg l^{-1}. It is worth noting that a better than 20/30 standard can be achieved by careful design of the SST to minimise the residual effluent SS. It has been estimated (ATV, 2000) that each mg l^{-1} of effluent SS increments effluent parameter values by:

0.3 – 1.0 mg l^{-1} BOD_5
0.8 – 1.4 mg l^{-1} COD
0.08 – 0.1 mg l^{-1} N
0.02 – 0.04 mg l^{-1} P

High-rate AS processes (SRT < 3d) can also be used for partial treatment or pre-treatment. Poor process stability and variable effluent quality are to be expected in this SRT range.

The hydraulic retention time (HRT) in the aeration tank is a function of the volumetric flow rate. AS processes in municipal WWTPs are typically hydraulically designed for flows up to 3 times average dry weather flow (DWF). Taking the variability of DWF into account, the nominal hydraulic retention time (V/Q) in carbonaceous AS processes in municipal WWTPs may vary in the range 2-12h. When the influence of sludge recycle from secondary clarification is taken into account, the actual hydraulic retention time at peak inflow may be as low as 1 hour.

12.4.2 Enhanced nitrogen removal

The enhanced biochemical removal of nitrogen from wastewaters is carried out in two distinct process stages. The first stage is the process of *nitrification*, or the conversion of ammonia to nitrate, and the second stage is *denitrification*, or the reduction of nitrate to gaseous nitrogen end-products.

12.4.2.1 Nitrification

Microbial nitrification is a two-step process, the first step being the conversion of ammonia to nitrite, which is accomplished by *Nitrosomonas* bacteria, while the second step involves the conversion of nitrite to nitrate by *Nitrobacter* bacteria. The overall chemical oxidation reaction is:

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O$$

Taking account of the incorporation of nutrients in the process of cell synthesis, using yields of 0.08 g VSS per g NH_4^+-N and 0.05 g VSS per g NO_2-N, for *Nitrosomonas* and *Nitrobacter*, respectively (USEPA, 1991), the overall reaction describing the complete nitrification process becomes:

$$1.00\,NH_4^+ + 1.89O_2 + 0.0805CO_2 \rightarrow 0.0161C_5H_7NO_2 + 0.952H_2O + 0.984NO_3 + 1.98H \quad (12.9)$$

where $C_5H_7NO_2$ is taken as representing the bacterial cell composition. Thus the conversion of 1mg of NH_4^+-N is estimated to result in the consumption of 4.32 mg oxygen, the production of 0.13 mg of nitrifying organisms and the destruction of 7.07 mg of alkalinity (as $CaCO_3$).

Nitrification kinetics

The nitrifying bacteria are chemoautotrophs; their growth energy is derived from the oxidation of inorganic nitrogen and their carbon source is carbon dioxide.

The growth rate of nitrifiers is estimated to be some 10-20 times slower than the growth rate of heterotrophs which are responsible for carbonaceous BOD removal. Of the two species responsible for nitrification, *Nitrobacter* has a higher growth rate than *Nitrosomonas*. The growth of the latter, which is responsible for the conversion of ammonia to nitrite, is thus normally rate-limiting for the nitrification process. It also follows from this that nitrite is not usually found in high concentrations in nitrifying processes operating under steady state conditions. The growth of *Nitrosomonas* can be expressed according to the Monod growth model (see chapter 11, equation (11.2)) as follows:

$$\mu_N = \hat{\mu}_N \frac{N}{K_N + N} \quad (12.10)$$

where μ_N = specific growth rate of *Nitrosomonas* (d^{-1})
$\hat{\mu}_N$ = maximum specific growth rate of *Nitrosomonas* (d^{-1})
K_N = half-saturation coefficient for *Nitrosomonas* (mg l^{-1} NH_4^+-N)
N = NH_4^+-N concentration (mg l^{-1})

For design purposes, the value of K_N may be taken as 1 mg $NH4^+$-N l^{-1}, while the value of the maximum specific growth rate constant is dependent on temperature and may be represented by the following empirical Arrhenius-type expression (USEPA, 1993):

$$\hat{\mu}_N = 0.47\theta^{(T-15)} \quad (12.11)$$

where θ is generally taken to have a value of about 1.1.

Because of the relatively low value of K_N, the nitrification process proceeds, under typical wastewater treatment conditions, at the maximum growth rate for the *Nitrosomonas* bacteria, i.e. it is a zero-order process, independent of the ammonia concentration. If, however, the ammonia nitrogen concentration drops close to the half saturation constant level of 1 mg l^{-1}, then the process becomes rate-limited by the reduced concentration according to equation (12.9).

As previously shown (Chapter 11, equation (11.14)), the effluent substrate (in this instance NH_4^+-N) in a completely mixed reactor, operating at steady state, can be expressed in terms of the sludge SRT and the Monod kinetic parameters, as follows:

$$NH_4^+ - N = \frac{K_N\left(1 + k_d \cdot SRT\right)}{SRT\left(\hat{\mu}_N - k_d\right) - 1} \qquad (12.12)$$

This relationship is plotted in Fig 12.4 for temperatures of 10 ^oC and 20 ^oC, illustrating the marked influence of temperature on nitrification performance. For example, Fig 12.4 indicates that a sludge SRT of about 10d is necessary to reduce the reactor ammonia nitrogen concentration to about 1 mg l^{-1} at an operating temperature of 10 ^oC, while at a temperature of 20 ^oC the ammonia concentration is reduced to 0.5 mg l^{-1} at an operating sludge SRT of about 5 days.

It has been observed (Stenstrom and Song, 1991) that the dissolved oxygen (DO) concentration has a significant influence on the nitrification process. It would appear that the growth rate of *Nitrosomonas* may be slowed down at DO concentrations less than 1 mg l^{-1}. The achievement of this limit value throughout the mixed liquor biomass, however, may require an operating DO level of at least 2 mg l^{-1}, depending on the intensity of mixing and the associated spatial DO gradients in the microbial floc.

As noted above, the nitrification process exerts a substantial alkalinity demand (7.1 mg alkalinity as $CaCO_3$ per mg NH_4^+-N). This inevitably reduces pH, particularly where there is insufficient buffering capacity in the wastewater being treated. The optimum pH range for nitrification would appear to be 7.0-8.5. For design purposes, this range may be extended to 6.5-9.0 (USEPA, 1993).

Nitrifying organisms are susceptible to inhibition by many organic substances, such as heavy metals (Hockenbury and Grady, 1977); Benmoussa et al., 1986). It has also been reported (Gujer, 1977) that the recycling of anaerobic digester supernatant may have an inhibitory effect on the *Nitrosomonas* growth rate.

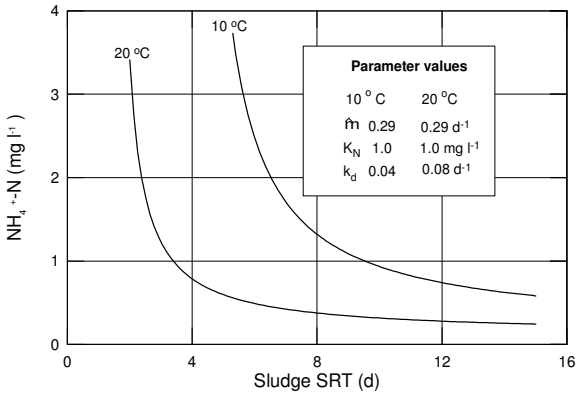

Fig 12.4 **Computed steady state ammonia nitrogen concentration in a completely mixed nitrification reactor** (equation 12.12)

12.4.2.2 Denitrification

The microbial reduction of nitrate is brought about by a variety of oxygen-utilising heterotrophic bacteria which, in the absence of oxygen, are capable of using nitrate in place of oxygen as a terminal electron acceptor. Research has shown that anoxic respiration of this kind also takes place at low DO concentration; The DO concentration at which denitrification stops has been reported to be 0.2 mg l^{-1} in pure cultures (Focht and Chang, 1975) and in activated sludge systems to be in the range 0.3-1.5 mg l-1 (Burdick at al., 1982). In the denitrification process, nitrate and its reduced forms act as electron acceptors, resulting in a stage-wise reduction of nitrate to gaseous nitrogen:

Redox state of nitrogen:	+5	+3	+2	+1	0
	NO_3^-	NO_2^-	NO	N_2O	N_2
	nitrate	nitrite	Nitric oxide*	Nitrous oxide*	Nitrogen*

*gaseous end products

A variety of heterotrophic bacteria, commonly present in activated sludges, participate in the denitrification process including *Alcaligenes, Achromobacter, Micrcoccus and Pseudomonas*. These organisms have the remarkable metabolic capability of using either oxygen or nitrate as an electron acceptor in their energy generation process. If oxygen is present, it is preferentially used over nitrate.

As heterotrophs, these bacteria use organic matter as their carbon source and hence remove BOD in conjunction with denitrification. The overall stoichiometry, neglecting cell synthesis, may be approximated as follows:

$$2NO_3^- + 2H^+ + 2.5C \rightarrow N_2 + 2.5CO_2 + H_2O \qquad (12.13)$$

Based on electron acceptor capacity, 1 g of nitrate nitrogen is equivalent to 2.86 g of oxygen. Thus, the combination of nitrification and denitrification processes can significantly reduce the overall oxygen demand relative to nitrification on its own.

The organic carbon requirement for the denitrification process, expressed in BOD terms, corresponds to a BOD_5 per NO_3^--N ratio ≥3. Thus, a typical fully nitrified effluent will not have sufficient residual biodegradable carbon to act as carbon source for the denitrification process. In practice, either the wastewater influent or a supplemental source such as methanol is used to provide the necessary organic carbon for the denitrification process.

The denitrification process increases the bicarbonate alkalinity, theoretically creating 3.57 mg alkalinity as $CaCO_3$ per mg of nitrate nitrogen reduced to nitrogen gas. This recovery of alkalinity partially reverses the drop in alkalinity and pH associated with the preceding nitrification process. This compensatory effect can be a significant process design consideration for wastewaters that are low in alkalinity and may be sufficient to prevent an inhibitory drop in pH in the nitrification step.

Denitrification process kinetics

The rate of growth of denitrifying organisms can be expressed in a Monod-type function, using nitrate as the rate-limiting nutrient:

$$\mu_D = \hat{\mu}_D \left(\frac{D}{K_D + D} \right) \left(\frac{S}{K_s + S} \right) \left(\frac{K_o}{K_o + S_o} \right) \qquad (12.14)$$

where

μ_D = specific growth rate (d^{-1})
$\hat{\mu}_D$ = maximum specific growth rate (d^{-1})
D = concentration of nitrate nitrogen (mg l^{-1})
K_D = half saturation coefficient (mg NO_3^--N l^{-1})

S and K_s refer to biodegradable substrate (mg BOD_5 l^{-1}); the Ks value may be taken as that which applies to heterotrophic growth under aerobic conditions.

So is the DO concentration (mg l^{-1}) and K_o is the half saturation coefficient for DO. The term $K_o/(K_o+S_o)$ acts as a switching function, turning the denitrification process on and off. A value of K_o of 0.1 mg l^{-1} has been suggested (IAWPRC model, 1994) for denitrification, implying a halving of the growth rate at a DO concentration of 0.1 mg l^{-1} relative to its value at zero DO.

As with all biological processes, denitrification is significantly influenced by temperature. The magnitude of this influence can be expressed by an Arrhenius-type function of the form:

$$\mu = \mu_{20}\theta^{(T-20)} \qquad (12.15)$$

where the value of θ lies within the range 1.02-1.08 (USEPA, 1993). At a θ value of 1.05, the mean of the range, the growth rate at 10 °C is calculated to be 61% of its rate at 20 °C.

In general, the denitrification process is much less sensitive to inhibitory substances than is the nitrification process. Experimental findings indicate that denitrification rates may be depressed below pH 6 and above pH 8.

12.4.2.3 Combined nitrification and denitrification processes

It is clear from the foregoing discussion that biological nitrification and denitrification processes have conflicting environmental requirements for optimal operation. Nitrification requires a highly aerobic environment with a sufficiently long microbial residence time to allow the development of a sufficiently high concentration of the slow-growing nitrifying bacteria, Nitrosomonas and Nitrobacter. These conditions result in a very low biodegradable carbon substrate level in nitrifying reactors. Denitrification, on the other hand, requires an anoxic environment and the availability of an ample biodegradable carbon substrate concentration (BOD_5:NO_3^--N ≥ 3).

The minimum sludge SRT for nitrification corresponds to the inverse maximum growth rate:

$$SRT_{(min)} = \frac{1}{\hat{\mu}_N}$$

where $\hat{\mu}_N$ is a function of temperature, as defined by equation (12.14). One approach to selecting the design sludge SRT value, SRT_d, is to apply a design service factor (DSF) to $SRT_{(min)}$ to take into account fluctuations in load and the desirability of operating in a stable nitrification zone (see Fig 12.8):

$$SRT_d = \frac{DSF}{\hat{\mu}} \qquad (12.16)$$

For example, at an operating temperature of 10 °C and using a DSF of 3.0, the value of θ_{sd} is found from equation (12.13) and (12.19) to be 10.2 days. A sludge SRT of about 10 days is commonly used in the design of nitrifying AS processes treating municipal wastewaters (TKN:BOD_5 = 0.1-0.2) in temperate climatic conditions, where the winter wastewater temperature is unlikely to drop below 10 °C.

In situations where the BOD_5:N ratio is high, it may be economically worthwhile to consider a two-stage biological process, the first stage being operated at a non-nitrifying sludge SRT and hence removing most of the carbonaceous BOD at an efficiently high specific conversion rate in a correspondingly low process volume. The second stage can then be designed as a nitrification stage, resulting in a significant reduction in the overall process volume relative to the volume required for a single stage process.

Denitrification requires an adequate carbon source to sustain the growth of the denitrifiers and a sufficient residence time under anoxic conditions to allow microbial utilisation of the nitrate. The rate of growth of denitrifiers (equation (12.17)) is inhibited by dissolved oxygen and positively influenced by the organic substrate and nitrate concentrations. Empirical evidence from municipal wastewater treatment practice (Schlegel, 1987) would indicate that the maximum denitrification may not be expected to exceed 0.15 kg NO_3^--N per kg influent BOD_5. The following empirical expression is proposed for the practical design of denitrifying reactors at 10 ^oC:

$$\frac{SRT_{DN}}{SRT} = 130\, R_{DN}{}^3 \qquad (12.17)$$

where SRT_{DN} is the anoxic zone solids residence time, SRT is the total solids residence time (aerobic + anoxic), R_{DN} is the nitrate nitrogen required to be denitrified per kg of influent BOD_5, through anoxic respiration. It is recommended (ATV, 2000) that the SRT_{DN}/SRT ratio should not be less than 0.2 or greater than 0.5.

Biological nitrogen removal requires the combination of nitrification and denitrification processes in an integrated process configuration. The most commonly used process configurations are (a) pre-denitrification, (b) step-feed denitrification, and (c) simultaneous nitrification/denitrification.

A pre-denitrification process layout is illustrated in Fig 12.5 (a). Assuming that full nitrification is achieved in the aerobic reactor and the anoxic reactor is designed for the required level of denitrification, the effluent NO_3^--N produced by such a process can be written as:

$$NO_3{}^-N = \frac{TKN^*}{(1 + R_{as} + R_i)} \qquad (12.18)$$

where TKN* is the nitrogen available for nitrification, that is the influent TKN less that contained in the process excess sludge. The latter can be approximated for design purposes as 0.05 times the influent BOD_5. Thus, for example, where the influent TKN:BOD_5 ratio is 0.2, TKN* ≅ 0.75 TKN_i, where TKN_i is the influent TKN.

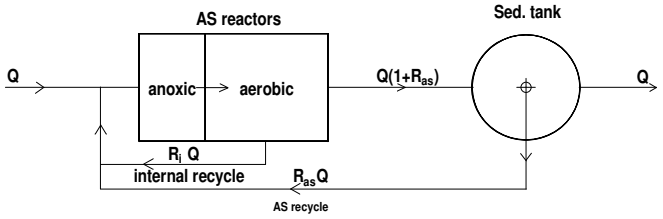

Fig 12.5 (a) **Pre-denitrification nitrogen removal process layout**

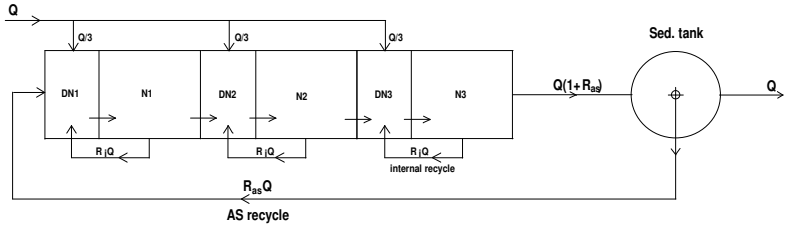

Fig 12.5 (b) **Step-feed nitrogen removal process layout**

The step-feed nitrogen removal process, which is illustrated in Fig 12.5 (b), consists of a number of anoxic and aerobic reactor pairs in series, the combined reactor volumes providing the appropriate sludge SRTs required for the processes of nitrification and denitrification. As illustrated, the influent

flow is step-fed into the anoxic reactors, requiring its subdivision into n equal streams, where n is the number of reactor pairs. Application of equation (12.18) to the downstream reactor pair of a step-feed process gives the expected effluent NO_3^--N:

$$NO_3^-N = \frac{TKN^*}{n(1+R_{as}+R_i)} \qquad (12.19)$$

Thus, the step-feed process enables the achievement of a low effluent nitrate concentration without the need for high recirculation rates, which may inhibit the denitrification process by the associated recycling of dissolved oxygen.

Simultaneous nitrification/denitrification is found to take place in processes with a large sludge SRT, such as oxidation ditches, where the reactor may have local zones that are well aerated and support nitrification and others that are poorly mixed/aerated providing an environment that supports denitrification. Denitrification can be enhanced in such reactors by directing the influent to the latter zone.

12.5 BIOLOGICAL PHOSPHORUS REMOVAL

The activated sludge process can be manipulated to enhance phosphorus-removal through creating process environmental conditions that produce so-called 'luxury' uptake of phosphorus, i.e. an uptake in excess of the normal metabolic fixation of phosphorus by bacterial cells. This enhanced uptake results in the phosphorus content of the biomass being increased from a typical 1.5-2.0% on a dry weight basis in a conventional AS process to the region of 3-6% in a biological phosphorus removal process (Stensel, 1991).

This enhanced uptake of phosphorus is achieved by subjecting the mixed liquor to an anaerobic/aerobic cycle. In the anaerobic phase there is an uptake of fermentation products (volatile fatty acids (VFAs) such as acetic acid and propionic acid), which accumulate as storage products within the microbial cells. This uptake of VFAs is accompanied by a corresponding release of cell phosphorus into solution. In the following aerobic stage the stored products are oxidised, resulting in a simultaneous enhanced uptake of phosphorus which is stored as polyphosphate within the cell.

The enhanced removal of phosphorus in the activated sludge process is considered to be due to a specific genus of bacteria, *Acetinobacter calcoaceticus*. The amount of phosphorus incorporated into the microbial biomass would appear to be not greatly influenced by temperature in the range 5-20 ^oC, with some research evidence to indicate a higher incorporation at lower temperatures (Stensel, 1991).

The two main factors that are known to influence biological phosphorus removal are the VFA:P ratio in the anaerobic reactor and the sludge SRT in the aerobic reactor. The VFA availability is dependent on the prior fermentation conditions and carbon substrate availability (BOD). Fukase et al. (1982) found that the BOD:P removal ratio increased from 19 to 26 as the sludge SRT was increased from 4.3 to 8.0 d. At the same time the phosphorus content of the activated sludge decreased from 5.4% to 3.7%. In practice, it may be difficult to achieve an effluent total phosphorus (TP) level in the range 1-2 mg l^{-1}, where the BOD_5:P ratio in the influent is less than 20.

The anaerobic reactor is usually sized to provide an hydraulic residence time of 0.5-1h, based on maximum dry weather flow and return sludge flow (USEPA, 1987; ATV, 2000). The shorter residence time is adequate for septic wastewaters having a relatively high BOD:P ratio, while the longer residence time may be necessary to allow some breakdown of particulate BOD in wastewater with a low soluble BOD content. As far as possible nitrate should be excluded from the anaerobic reactor, as its presence inhibits the fermentation process that produces VFAs.

The aerobic reactor is designed as a conventional activated sludge process, the selected sludge SRT depending on whether nitrification is required or not. As a general rule, where the aerobic reactor is designed for nitrification, a denitrifying anoxic zone should also be provided to avoid carryover of nitrate into the anaerobic reactor for the reasons already outlined. As noted above, the incorporated

phosphorus content of AS decreases with increasing sludge SRT as does the process sludge yield. Hence, the required BOD:P ratio for a given residual phosphorus concentration increases with the aerobic reactor sludge SRT.

In addition to the foregoing process design considerations, two operational considerations, in particular, must be taken into account to achieve a low residual effluent phosphorus concentration: (a) the effluent suspended solids must be maintained at a low level, since they have a high phosphorus concentration (2-4% by dry weight) and (b) the excess sludge must be maintained in an aerobic state to avoid phosphorus loss.

The TP:BOD_5 ratio in normal municipal wastewater is typically within the range 0.025-0.035. In the absence of enhanced P-uptake, the phosphorus requirement for heterotrophic growth may be approximated as a TP:BOD_5 ratio of 0.01. Under optimal operating conditions for enhanced biological P-removal, the uptake of phosphorus can be doubled relative to a TP:BOD_5 ratio of 0.02. However, enhanced biological P-removal may require to be backed up by chemical precipitation, where a very low effluent P is required and also where a high degree of reliability in process performance is essential.

12.6 DISSOLVED OXYGEN REQUIREMENTS

A concentration of dissolved oxygen (DO) must be maintained in the mixed liquor to satisfy microbial respiration requirements. For carbonaceous oxidation the limiting DO concentration is considered to be about 0.5 mg l^{-1}. Nitrifying processes require a somewhat higher DO, as discussed earlier in section 12.4.2. For process energy efficiency it is desirable that the operating concentration of dissolved oxygen should be as low as possible since the rate of oxygen transfer by aeration systems is proportional to the oxygen saturation deficit and hence energy requirements for aeration are minimised by operating at the lowest satisfactory oxygen concentration.

The carbon-related AS process daily oxygen demand (POD_C) is obviously directly related to the daily influent BOD_5 load and is also influenced by the sludge SRT and the process temperature. These parameters have been empirically correlated as follows: (ATV, 2000):

$$POD_c = 0.56 + \frac{0.15 \cdot SRT \cdot F_T}{1 + 0.17 \cdot SRT \cdot F_T} \text{ kg } O_2 \text{ per kg } BOD_5 \qquad (12.20)$$

where F_T is a function of temperature, as already defined (eqn. 12.7). The relationship of equation (12.20) is plotted in Fig 12.6 for temperatures of 10 and 20 ^oC. The heterotrophic biomass respiration rate, computed from equations (12.5) and (12.20), is plotted in Fig 12.7. These plots illustrate the strong influence of both temperature and SRT on respiration rate.

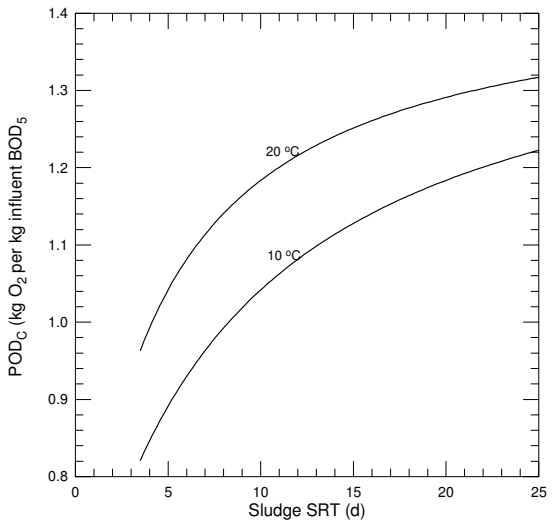

Fig 12.6 **Carbonaceous process oxygen demand (POD_C)** (equation (12.19))

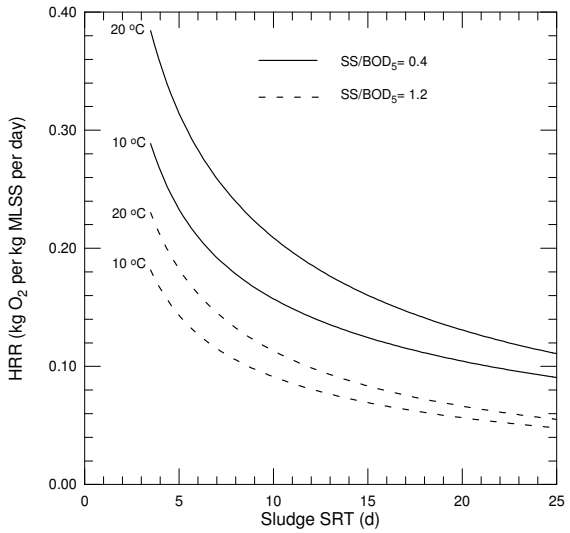

Fig 127 **Heterotrophic respiration rate (HRR) as a function of sludge SRT**

As discussed in Section 12.4.2, the oxidation of 1mg ammonia nitrogen to nitrate requires 4.32 mg O_2, while the denitrification of 1 mg nitrate nitrogen to N_2 produces 2.86 g O_2. Hence the AS process nitrogen-related oxygen demand POD_N can be expressed as follows:

$$POD_N = 4.32\ N_{OX} - 2.86\ N_{DN} \qquad (12.21)$$

where N_{OX} is the nitrate nitrogen produced by the nitrification process and N_{DN} is the nitrate nitrogen reduced to nitrogen gas by the denitrification process.

Equations (12.19) and (12.20) quantify the carbon-related and nitrogen-related oxygen demands, respectively, of the AS process on an influent load basis. In municipal wastewater treatment, in particular, there is a significant fluctuation in loading in any 24h period, as illustrated in Fig 12.8. This fluctuation must be taken into account in calculated the required oxygen transfer capacity of the installed aeration system under field conditions (OTC_F):

$$OTC_F = PF_C * POD_C + PF_N * POD_N \text{ kg } O_2 \text{ d}^{-1} \quad (12.22)$$

Where PF_C and PF_N are peak oxygen demand factors related to carbon and nitrogen oxidation, respectively; POD_C is the average daily carbon-related oxygen demand and POD_N is the average daily nitrogen-related oxygen demand. The design value of PF_C for municipal treatment plants is typically in the range 1.1 –1.3, the higher value relating to low sludge SRT plants. The design value of PF_N for municipal nitrifying AS processes is typically in the range 1.5-2.5. It should also be noted that the peak carbon-related and nitrogen-related demands may not occur simultaneously (ATV, 2000). The recorded diurnal fluctuation in influent BOD_5 and TKN loads, recorded at a wastewater treatment works (WwTW) serving a population of about 30 000PE, is illustrated in Fig 12.8.

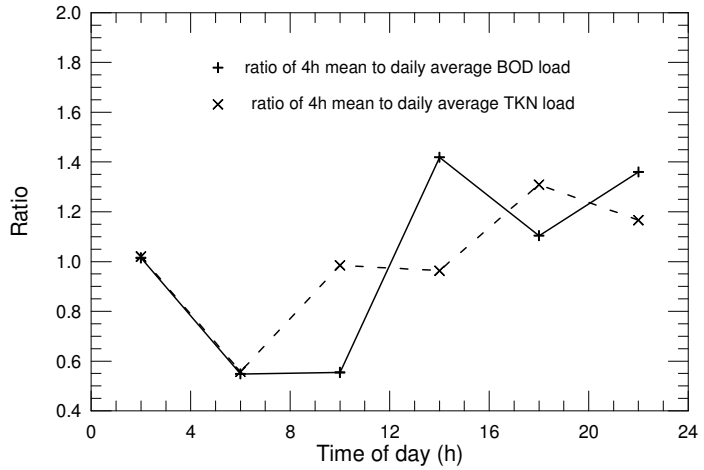

Fig 12.8 Example of diurnal load variation at a municipal WwTW (Works loading 30 000 PE)

As discussed in Chapter 14, the standard oxygen transfer capacity (SOTC) of aeration systems is conventionally related to clean water at zero oxygen concentration in solution and a water temperature of 20 ^oC. The reader is referred to Chapter 14 for a discussion of the relation between the SOTC and the transfer capacity under field conditions (OTCF).

12.7 SEPARATION AND RECYCLING OF ACTIVATED SLUDGE

12.7.1 Settleability parameters

The most important physical characteristic of an activated sludge is its separability from the liquid in which it is dispersed. Activated sludges with good separation characteristics are flocculent and settle as shown in Fig 2.3, leaving a clear supernatant free of visible suspended matter. The settleability of an activated sludge is measured by the volume occupied by unit mass of sludge solids (ml g^{-1} or l kg^{-1}) following a specified settling period. Three standardised test procedures have been developed (i) the sludge volume index or *SVI*, (ii) the stirred sludge volume index or *SSVI*, (iii) the diluted sludge volume index or *DSVI*.

The SVI is measured the volume in ml occupied by 1g of sludge when 1 litre of mixed liquor has been allowed to settle without stirring for 30 min in a 1-litre cylinder. Normal flocculent sludges have SVI

values within the range 80-120 ml g^{-1}, while non-flocculent sludges or *bulking* sludges may have SVI values in excess of 200. The SVI test has been found to be sensitive to solids concentration, having poor reproducibility for the same sludge at different concentrations.

The *DSVI* or diluted sludge volume index is a modification of the SVI test that has been shown (IAWQ, 1997) to give improved reproducibility, particularly in respect of insensitivity to sludge concentration. The DSVI is carried out in the same manner as the SVI with the additional requirement that the sludge is diluted to ensure that the settled volume after 30 min is within the range 150-250 ml. The dilution should be made with water derived from the mixed liquor. DSVI is the preferred settleability parameter for use in the design of activated sludge clarifiers, commonly referred to as secondary sediment tanks (SSTs).

The *SSVI* or stirred sludge volume index (White, 1976) is a further measure of sludge settleability. It is the volume as ml g^{-1} occupied by an AS sample after 30 min settling in a cylindrical vessel, 0.5m high, stirred at a speed of 1 rpm. It has been found that the SSVI test procedure gives better reproducibility than the SVI test procedure (IAWQ, 1997).

12.7.2 Sludge bulking and its control

It is generally considered that the floc-forming properties of an activated sludge are related to the morphology of the dominant organism type. Where cocoid and rod-shaped bacteria predominate, activated sludges invariably exhibit good floc-forming properties and hence good settling behaviour. It is considered that the presence of some filamentous bacteria within the floc is beneficial in strengthening the floc structure. However, the predominance of filamentous bacteria is invariably detrimental to floc structure and settling behaviour. The most common detrimental effect of excessive filamentous growth is the creation of so-called bulking sludge, which, as the name implies, has a low aggregative capacity and exhibits poor settling behaviour. About 25 different filamentous organisms are known to cause sludge bulking (Gray, 1990). The environmental conditions that give rise to the development of a bulking sludge and the commonly associated filamentous organisms are summarised in Table 12.3.

Table 12.3 Bulking activated sludge

Suggested cause	Indicative filament type
Low dissolved oxygen	Type 1701, *Spherotilus natans, Haliscomenobacter hydrossis, Microthrix parvicella*
Low loading rate (high sludge SRT)	*M. Parvicella, H. hydrossis, Nocardia amarae*, Types 021N, 0041, 0675, 0092,0581, 0961, 0803
Septic wastewater/sulphide	*Thiotrix,Beggiatoa*, Type 021N
Nutrient deficiency	*Thiothrix, S. natans*, Type 021N
Low pH	Fungi

While plant operating experience and research (Pipes, 1967; Chambers and Tomlinson, 1982; Strom and Jenkins, 1984) have shown that the environmental conditions listed in Table 12.3 are favourable to the growth of filamentous organisms, they are not necessarily the only causative factors and may not always give rise to bulking sludge, the precise cause of which can sometimes be difficult to identify. Bulking occasionally occurs in the absence of filamentous bacteria. This type of bulking is known as zoogleal or viscous bulking, resulting in a mixed liquor having a slimy non-flocculent consistency (Pipes, 1979).

A second detrimental operational characteristic which may result from excessive filamentous growth is the problem of foaming in the aeration tank, which is particularly associated with the Nocardia and Microthrox species (Foot, 1992). Biological foaming takes the form of a floating layer of activated

sludge of mousse-like consistency, containing entrapped air bubbles in a mainly filamentous floc structure. It would appear that foam-forming filamentous organisms possess hydrophobic properties and may also be capable of generating extracellular surfactants, both of which are prime ingredients in the production of a floating activated sludge foam layer. Foaming problems appear to be mainly associated with low loading conditions (high sludge SRT). A variety of operational measures have been applied in the control of foaming (Tipping, 1995; Pitt and Jenkins, 1990), the most common being the reduction of the microbial solids residence time or sludge SRT, which can be achieved by reducing the operating MLSS concentration. Other methods include the use of water sprays, anti-foaming agents, biocide addition and physical removal of the float layer.

As noted in Table 12.3, nutrient deficiency is a potential cause of excessive filamentous growth. The incoming wastewater must provide all the essential nutrients for microbial growth, including nitrogen, phosphorus and trace elements (see Table 11.1). Municipal wastewaters generally satisfy the nutrient requirements bur some industrial wastewaters may not. Pipes (1979) recommends a BOD_5:N:P ratio of 100:5:1 to prevent bulking. The BOD_5:N:P) in domestic sewage is typically in the region 100:20:3; hence, in most activated sludge applications, there is an excess of nitrogen and phosphorus over that required for microbial growth.

12.7.3 Selector tanks

Bulking in high sludge SRT AS processes is most commonly associated with completely mixed aeration basins. In such situations there is a uniformly low substrate concentration throughout the reactor volume, which favours the growth of filamentous organisms over floc-formers. Under conditions of high soluble substrate concentration, however, it is believed that the floc-forming organisms can absorb substrate more rapidly than the filamentous species and hence can outgrow the latter. This negative operational characteristic of low-rate completely mixed processes can be neutralised by the use of an upstream aerobic selector tank (Chudoba et al., 1973), which is operated at a sufficiently high loading rate to favour the growth of floc-forming organisms over filamentous organisms. Upstream anaerobic tanks designed to promote enhanced phosphorus removal can also act in a selector capacity.

The German ATV-DVWK Design Standard (2000) gives guidance for sizing aerobic selectors recommending a volumetric loading of 10 kg BOD_5 per m^3 per day, coupled with an oxygen input of 4 kg O_2 per m^3 per day.

While selectors have been widely used in conjunction with enhanced nitrogen removal plants their success in eliminating sludge bulking has been found to be somewhat unpredictable (Eikelboom, 1994). The occurrence of bulking problems has been reported (IAWQ, 1997) to be seasonal and to be more likely with settled sewage.

It is worthy of note that the foregoing environmental requirement in respect of substrate concentration is automatically satisfied in reactors of plug-flow configuration, in which the substrate concentration decreases from a maximum at the inlet end to a minimum at the outlet end.

12.8 SECONDARY EDIMENTATION TANKS

The maintenance of a steady-state mixed liquor biomass concentration in AS reactors is conventionally achieved by recycling the settled sludge from a downstream sedimentation process (secondary sedimentation), as illustrated in Fig 12.1. The secondary sedimentation tank (SST) has to perform two functions: (a) it must produce a well-clarified effluent and (b) it must have a sufficient solids flux capacity to allow the sludge to be recycled. The latter is frequently the governing performance criterion.

The basic process variables that influence the design of SSTs are:

- the settleability of the MLSS, as measured by either the SVI, DSVI or SSVI parametric values

- the MLSS concentration in the AS reactor
- the forward flow rate Q from the AS reactor to the SST
- the underflow recycle rate Q_u from the SST back to the AS reactor.

At the present time there is no universally accepted analytical basis for SST design. A comprehensive review of SST theory, modelling, design and operation was carried out by the IAWQ and its findings are contained in Scientific and Technical Report No. 6 (1997).

Two empirically-based computational procedures for the design of SSTs are outlined in the following sections. The first of these is based on work carried out in the UK by White (1975, 1976),the application of solids flux capacity considerations and is based on the SSVI settleability parameter. The second is an empirical design procedure used in Germany, based on the DSVI settleability parameter, the details of which are to be found in the ATV-DVWK Standard 131E (2000).

12.8.1 SST Design

White (1975) suggested a design procedure for SSTs, which is based on an empirical correlation of the limiting solids flux capacity, F_L, the sludge underflow rate, Q_u and the sludge SSVI:

$$F_L = 307(SSVI)^{-0.77}(Q_u / A)^{0.68} \text{ kg m}^{-2}\text{h}^{-1} \qquad (12.23)$$

where A is the plan area of the sedimentation tank (m^2) and Q_u is the underflow rate (m^3h^{-1}).

The specific applied solids loading rate, F_a, is:

$$F_a = (Q_u + Q)MLSS/A \qquad (12.24)$$

where Q is the wastewater flow ($m^3\ h^{-1}$).

Neglecting the leakage of solids in the clarified effluent, it is clear that the solids flux capacity of the tank is reached when $F_a = F_L$. Thus, by combining equations (12.7) and (12.8), the following parametric relationship applies at the limiting solids loading condition:

$$MLSS = 307(SSVI)^{-0.77}\left(\frac{Q}{A}\right)^{-0.32}\left(\frac{R^{0.68}}{1+R}\right) \qquad (12.25)$$

where R is the sludge recycle ratio (Q_u/Q) and Q/A is the SST surface loading rate (m h^{-1}).

Equation (12.25) correlates four parameters, MLSS, SSVI, Q/A and R, at the limiting solids flux condition. Ekama and Marais (1986) noted that that White's empirical function is valid up to the critical limiting recycle rate, which they quantified from solids flux considerations as follows:

$$\left(\frac{Q_u}{A}\right)_c = 1.6.12 - 0.00793\,SSVI \quad \text{m h}^{-1} \qquad (12.26)$$

for SSVI < 125 l kg^{-1}.

The maximum SST surface loading rate (Q/A), as derived from equations (12.25) and (12.26), is plotted as a function of SSVI in Fig 12.9 for MLSS values in the normal operating range of 2 to 4 kg m^{-3}.

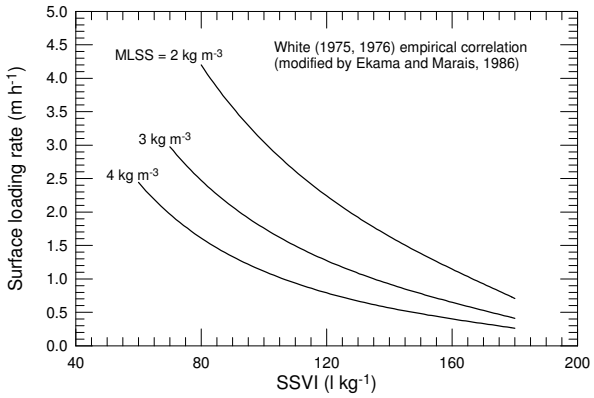

Fig 12.9 **Maximum SST surface loading based on equations (12.25) and (12.26)**

The second empirical design procedure is set out in detail in the German ATV-DVWK Standard 131E of May 2000. It is applicable to SSTs up to 60m in length or diameter. DSVI is the settleability parameter used and its product with MLSS, known as the diluted sludge volume or DSV (DSV = MLSS x DSVI), is a key parameter of the ATV design procedure. The following operational limit values apply:

- MLSS > 1 kg m^{-3}
- diluted sludge volume index $50 \leq$ DSVI ≤ 200 l kg^{-1}
- DSV ≤ 600 l m^{-3}
- recycle rates: $R \leq 0.75$ for horizontal flow tanks
  $R \leq 1.00$ for vertical flow tank, both R-values relating to peak wet weather flow

The maximum permissible *surface loading rate* q_a (m h^{-1}) is specified by a sludge volume loading rate parameter, q_{sv}, defined as follows:

$$q_{sv} = q_a \cdot \text{DSV} \quad \text{l m}^{-2}\,\text{h}^{-1} \qquad (12.27)$$

To achieve an effluent suspended solids ≤ 20 mg l^{-1}, the following limit values are specified:

horizontal-flow tanks	$q_{sv} \leq 500$ l m^{-2}h^{-1}; $q_a \leq 1.6$ m h^{-1}
vertical-flow tanks	$q_{sv} \leq 650$ l m^{-2}h^{-1}; $q_a \leq 2.0$ m h^{-1}

For design purposes horizontal-flow tanks are taken to be those where the ratio of the inflow depth below the water surface to the length of the horizontal component of the flow path to the outlet is less than 1:3. SSTs where this ratio is greater than 1:2 are classified as vertical-flow SSTs. For intermediate ratios, the permitted q_{sv} value may be linearly interpolated.

The computed solids recycle capacity ultimately determines the maximum feasible operating MLSS concentration in the aeration tank. The recycle capacity is a function of the recycle rate and the underflow solids concentration. The following empirical expression is used to estimate the sludge concentration in the bottom sludge layer, S_{BL}, in the SST:

$$S_{BL} = \frac{1000}{\text{DSVI}}\, t_{th}^{\,1/3} \ \text{kg m}^{-3} \qquad (12.28)$$

where t_{th} is the thickening time in hours. The ATV recommended thickening times are set out in Table 12.4.

The solids concentration in the returned sludge S_U is less than S_{BL} due to dilution resulting from short-circuit sludge flow. This is taken into account as follows:

SSTs fitted with scraper systems	$S_U \cong 0.7\ S_{BL}$
SSTs fitted with suction pipes	$S_U \cong 0.5 - 0.7\ S_{BL}$

With vertical-flow SSTs, it is permitted to assume that $S_U \cong S_{BL}$.

Table 12.4 Recommended sludge thickening times

Type of wastewater treatment	Thickening time t_{th} (h)
Activated sludge plants without nitrification	1.5-2.0
Activated sludge plants with nitrification	1.0-1.5
Activated sludge plants with denitrification	2.0-2.5*

*t_{th} > 2h requires very advanced denitrification

Under steady state operating conditions the forward flux of sludge to the SST is equal to the recycle rate:

$$MLSS \cdot Q(1+R) = R \cdot Q \cdot S_U \qquad (12.29)$$

Based on equation (12.28), assuming a sludge thickening time of 2h, a maximum R-value of 0.75 for horizontal-flow SSTs and a maximum R-value of 1 for vertical-flow SSTs, the maximum feasible operating MLSS values are found to be related to DSVI as follows:

horizontal-flow SSTs	$MLSS = \frac{378}{DSVI}$
vertical-flow SSTs	$MLSS = \frac{600*}{DSVI}$

(* limit value for the product MLSS x DSVI is 600 l m^{-3}). These functional relations are illustrated graphically in Fig 12.10.

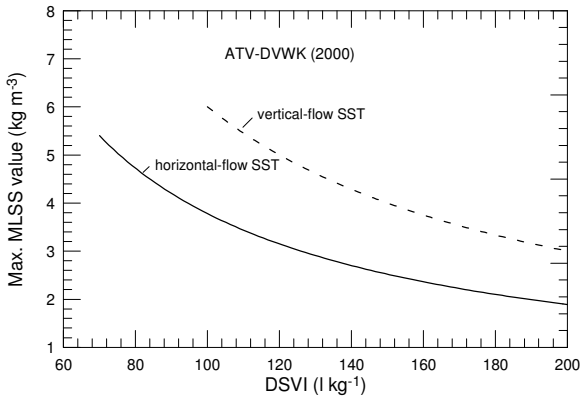

Fig 12.10 **Maximum permissible MLSS as function of sludge DSVI**
(R = 0.75 at peak flow; t_{th} = 2h)

Surface loading rate envelopes for horizontal-flow and vertical-flow SSTs, based on equation (12.28), subject to the constraints imposed by the sludge recycle requirements, are plotted in Fig 12.11. The plotted envelopes illustrate the potentially greater capacity of vertical-flow SSTs relative to horizontal-flow SSTs, in respect of DSV range and surface loading rate.

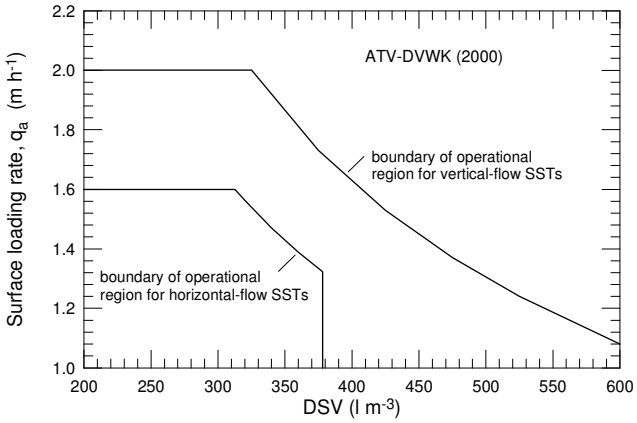

Fig 12.11 **Surface loading design envelopes for SSTs**
(R = 0.75 at peak flow; t_{th} = 2h)

The ATV Standard also specifies a procedure for the computation of the required tank depth, which is considered to comprise four zones:

h_1: clean water zone
h_2: separation zone/return flow zone
h_3: density flow and storage zone
h_4: thickening and sludge removal zone

The clean water zone should have a depth of at least 0.5m:

$$h_1 = 0.5 \text{ m}$$

The separation/return flow zone is sized to provide a retention time of 0.5h, referred to the free water volume, for the incoming flow, including recycle:

$$h_2 = \frac{0.5 q_a (1+R)}{\left(1 - \frac{DSV}{1000}\right)} \quad \text{m}$$

The density flow and storage zone depth is empirically determined as follows:

$$h_3 = \frac{0.45 \cdot DSV \cdot q_a (1+R)}{500} \quad \text{m}$$

The volume of the thickening and sludge removal zone is taken as that required to store the thickened sludge volume produced in the selected thickening time, assuming a uniform thickened concentration S_{BL}:

$$h_4 = \frac{MLSS \cdot q_a (1+R)}{S_{BL}} = \frac{DSV \cdot q_a (1+R) t_{th}^{2/3}}{1000} \quad \text{m}$$

The required overall liquid depth H_L ($H_L = h_1 + h_2 + h_3 + h_4$) is:

$$H_L = 0.5 + q_a (1+R) \left[\frac{0.5}{1 - \frac{DSV}{1000}} + DSV \left(0.0009 + \frac{t_{th}^{2/3}}{1000} \right) \right] \qquad (12.30)$$

The following numerical example illustrates the ATV calculation sequence for SST design:

Activated sludge DSVI = 120 l kg^{-1}; SST thickening time, t_{th} = 2h

	Horizontal-flow SST	Vertical-flow SST
Recycle ratio, R (peak flow)	0.75	1.00
Max. permissible DSV (l m^{-3})	378	600
Max, permissible MLSS (kg m^{-3})	3.15	5.00
Selected MLSS (kg m^{-3})	3.00	4.00
Resultant DSV (l m^{-3})	360	480
Max. permissible q_a (m h^{-1})	1.39	1.25
Selected q_a (m h^{-1})	1.25	1.25
Required liquid depth (m)	4.15	5.38

The thickened sludge in circular SSTs can be removed by either a vertical suction pipe system or can be moved to a central hopper by a scraper blade system, both of which are suspended from a rotating bridge.

ATV recommends that the design velocity in suction pipes should be in the range 0.6 – 0.8 m s^{-1} and the suction pipe spacing should not exceed 3 to 4m.

ATV guidance values for the design of sludge scraper systems are given in table 12.5

Table 12.5 Guidance values for design of sludge scrapers (ATV, 2000)

	Circular SST	Rectangular SST	
		Bridge	Flight
Scraper blade height (m)	0.4-0.6	0.4-0.9	0.15-0.3
Bridge peripheral vel. (m h^{-1})	72-144	<108	36-108

Effluent is commonly discharged over peripheral weirs in circular SSTs. To avoid carryover of solids, the weir loading should not exceed 10 m^3 $m^{-1}h^{-1}$ for peripheral weirs and 6 $m^3m^{-1}h^{-1}$ for inboard double-sided decanting channels (ATV, 2000).

12.9 BATCH PROCESSES

As noted in the introduction to this chapter, the AS process had its origins in batch process experimental studies, but was subsequently developed for operation in continuous flow mode, as illustrated in Fig 12.1. Advances in process automation and the development of aeration systems that allow non-continuous aeration without the risk of clogging, have facilitated the development of batch-operated AS processes, commonly known as sequencing batch reactors or SBRs. The SBR reactor is designed to accommodate both the AS process functions and solids separation through a sequence of operational modes, as illustrated in Fig 12.12.

It is clear that a single SBR cell must be supplemented by external storage to hold the influent volume generated in the non-filling period of the cycle duration. Alternatively, by using two or more cells, the need for external storage can be eliminated.

SBR processes can be designed for carbonaceous BOD removal only or for enhanced nitrogen and phosphorus removal by providing the appropriate environmental conditions, as already outlined for conventional continuous flow processes. For example, the fill phase can be used for denitrification if accompanied by an adequate input of mixing. Anaerobic selectors can be incorporated to promote enhanced biological removal of phosphorus. Where nitrification is required, the aerobic sludge SRT (=sludge residence time x fraction of time aerators are operating) must be sufficient to sustain the process (refer section 12.2.2).

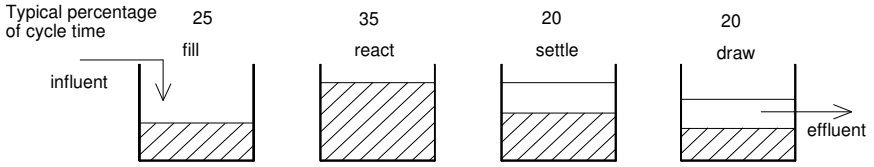

Fig 12.12 **Typical operating cycle for an SBR Activated sludge process**

The aeration system is switched on during the react phase and in some cases also during the filling phase of the cycle. It is switched off during the settle and decant phases. The system must be designed for on/off operation.

There is a more favourable solids separation environment in SBRs than in SSTs due to (a) settling is quiescent, and (b) the absence of a recycle stream reduces the solids flux. This enables up to 50% higher surface loading rates, relative to SSTs, to be used in SBRs.

The draw phase of the SBR cycle requires the use of floating decanter systems or other automatically controlled variable-level liquid withdrawal systems. As a consequence, weir loadings tend to be significantly higher than in SSTs. This may lead to the entrainment of solids from the sludge blanket if there is not a sufficient depth of clear water between the weir level and the top of the sludge blanket.

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