

Aerobic and Anoxic Suspended-Growth Biotechnologies

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Abstract Among the suspended-growth biological treatment processes covered in this chapter are conventional activated sludge, high rate activated sludge, pure oxygen activated sludge, contact stabilization, activated sludge with nitrification, separate stage nitrification, separate stage denitrification, extended aeration, oxidation ditch, Powdered Activated Carbon Treatment (PACT) process, carrier-activated sludge processes (CAPTOR and CAST systems), activated biofilter (ABF), vertical loop reactor (VLR), and phostrip process. This chapter describes the above processes and explains their practice, limitations, design criteria, performance, energy requirements, process equipment, performance, and costs.

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Key Words Suspended-growth • conventional activated sludge • high rate activated sludge • pure oxygen activated sludge • contact stabilization • nitrification • denitrification • carrier-activated sludge • ABF • VLR • CAPTOR • CAST • PACT.

1. CONVENTIONAL ACTIVATED SLUDGE

1.1. Description

Activated sludge is a continuous flow, biological treatment process characterized by a suspension of aerobic microorganisms, maintained in a relatively homogeneous state by the mixing and turbulence induced by aeration. The microorganisms are used to oxidize soluble and colloidal organics to CO_2 and H_2O in the presence of molecular oxygen (1). The process is generally preceded by primary sedimentation. The mixture of microorganisms and wastewater formed in the aeration basins, called mixed liquor, is transferred to gravity clarifiers for liquid–solids separation (Fig. 13.1). The major portion of the microorganisms settling out in the clarifiers is recycled to the aeration basins to be mixed with incoming wastewater, while the excess, which constitutes the waste biosolids (sludge), is sent to the biosolids handling facilities (2). The rate and concentration of activated sludge returned to the aeration basins determines the mixed liquor suspended solids (MLSS) level developed and maintained in the basins. During the oxidation process, a certain amount of the organic material is synthesized into new cells, some of which then undergo auto-oxidation (self-oxidation, or endogenous respiration) in the aeration basins, the remainder forming net growth or excess biosolids (3). Oxygen is required in the process to support the oxidation and synthesis reactions. Volatile compounds are driven off to a certain extent in the aeration process. Metals will also be partially removed, with accumulation in the sludge.

Activated sludge systems are classified as high rate, conventional, or extended aeration (low rate) based on the organic loading. In the conventional activated sludge plant, the wastewater is commonly aerated for a period of 4–8 h (based on average daily flow) in a plug flow hydraulic mode (Fig. 13.1). Either surface or submerged aeration systems can be employed to transfer oxygen from air to wastewater. Compressors are used to supply air to the submerged systems, normally through a network of diffusers, although newer submerged devices, which do not come under the general category of diffusers (e.g., static aerators and jet aerators) are applied. Diffused air systems may be classified as fine bubble or coarse bubble. Diffusers

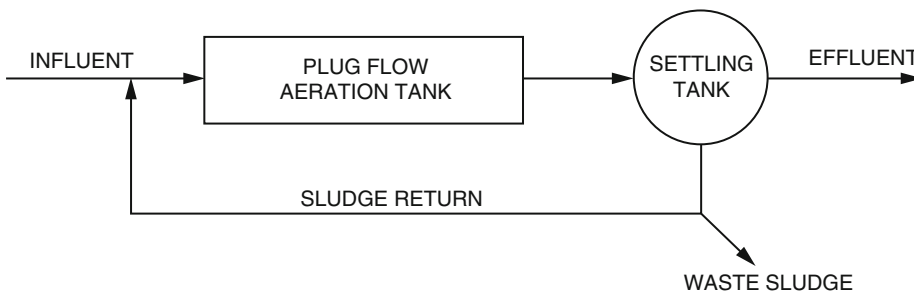


Fig. 13.1. Conventional activated sludge flow diagram (Source: U.S. EPA).

commonly used in activated sludge service include the following: porous ceramic plates laid in the basin bottom (fine bubble), porous ceramic domes or ceramic or plastic tubes connected to a pipe header and lateral system (fine bubble), tubes covered with synthetic fabric or wound filaments (fine or coarse bubble), and specially designed spargers with multiple openings (coarse bubble) (1).

Activated sludge is the most versatile and widely used biological process in wastewater treatment (4). Common process modifications include step aeration (Fig. 13.2); contact stabilization (Fig. 13.3); and complete mix flow regimes (Fig. 13.4). Alum or ferric chloride is sometimes added to the aeration tank for phosphorus removal (1).

The activated sludge process is used for the treatment of both domestic wastewater and biodegradable industrial wastewater. The main advantage of the conventional activated sludge system is the lower initial cost of the system, particularly when a high quality effluent is required. Industrial wastewater, which is amenable to biological treatment and degradation, may be jointly treated with domestic wastewater in a conventional activated sludge system.

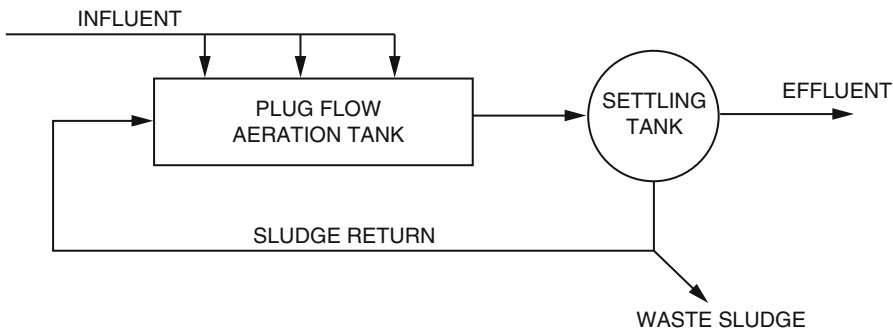


Fig. 13.2. Step aeration flow diagram (Source: U.S. EPA).

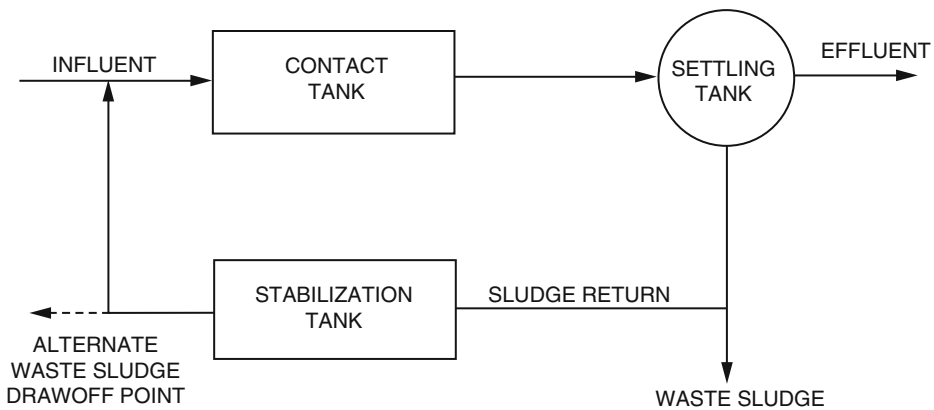


Fig. 13.3. Contact stabilization flow diagram (Source: U.S. EPA).

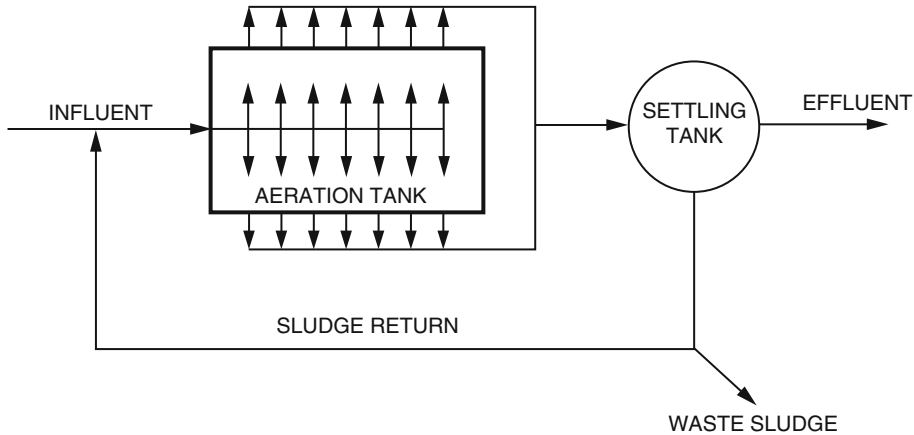


Fig. 13.4. Complete mix activated sludge flow diagram (Source: U.S. EPA).

Some of the process disadvantages are (5, 6):

1. Limited BOD₅ (5-day biochemical oxygen demand) loading capacity
2. Poor organic load distribution
3. Plant upset with extreme variations in hydraulic, organic, and toxic loadings
4. Operational complexity
5. High operating costs and energy consuming mechanical compressors
6. Diffuser maintenance

1.2. Performance and Design Criteria

BOD and ammonia-N removals are as follows (1, 7):

- BOD₅ removal (conventional activated sludge) 85–95%
- NH₄-N removal (nonnitritified systems) 10–20%

The residuals or biosolids increase (as measured by volatile suspended solids, VSS, production from the conventional activated sludge process) as food-to-microorganism (F/M) loadings increase (8, 9). For an F/M value of 0.3 lb BOD₅/d/lb MLVSS, the excess in VSS is 0.5 lb/lb BOD₅ removed; while for an F/M ratio of 0.5 lb BOD₅/d/lb MLVSS, the excess in VSS increases to 0.7 lb/lb BOD₅ removed. Here 1 lb/d/lb = 1 kg/d/kg; 1 lb/lb = 1 kg/kg.

Design criteria for the conventional activated sludge process are summarized as follows (1, 7):

- Volumetric loading = 25–50 lb BOD₅/d/1,000 ft³ = 400.7–801.4 g BOD₅/d/m³
- Aeration detention time (based on average daily flow) = 4–8 h
- MLSS = 1,500–3,000 mg/L
- F/M = 0.25–0.5 lb BOD₅/d/lb MLVSS = 0.25–0.5 kg BOD₅/d/kg MLVSS
- Air required = 800–1,500 std. ft³/lb BOD₅ = 50–94 m³/kg removed
- Biosolids retention time = 5–10 d

1.3. Mechanical Aeration

Mechanical aeration methods include the submerged turbine with compressed air spargers (agitator/sparger system) and the surface-type mechanical entrainment aerators (Fig. 13.5). The surface-type aerators entrain atmospheric air by producing a region of intense turbulence at the surface around their periphery. They are designed to pump large quantities of liquid, thus dispersing the entrained air and agitating and mixing the basin contents. The agitator/sparger system consists of a radial-flow turbine located below the mid-depth of the basin, with compressed air supplied to the turbine through a sparger (1).

The submerged turbine aeration system affords a convenient and relatively economical method for upgrading overloaded activated sludge plants. To attain optimum flexibility of oxygen input, the surface aerator can be combined with the submerged turbine aerator. Several manufacturers supply such equipment, with both aerators mounted on the same vertical shaft. Such an arrangement might be advantageous if space limitations require the use of deep

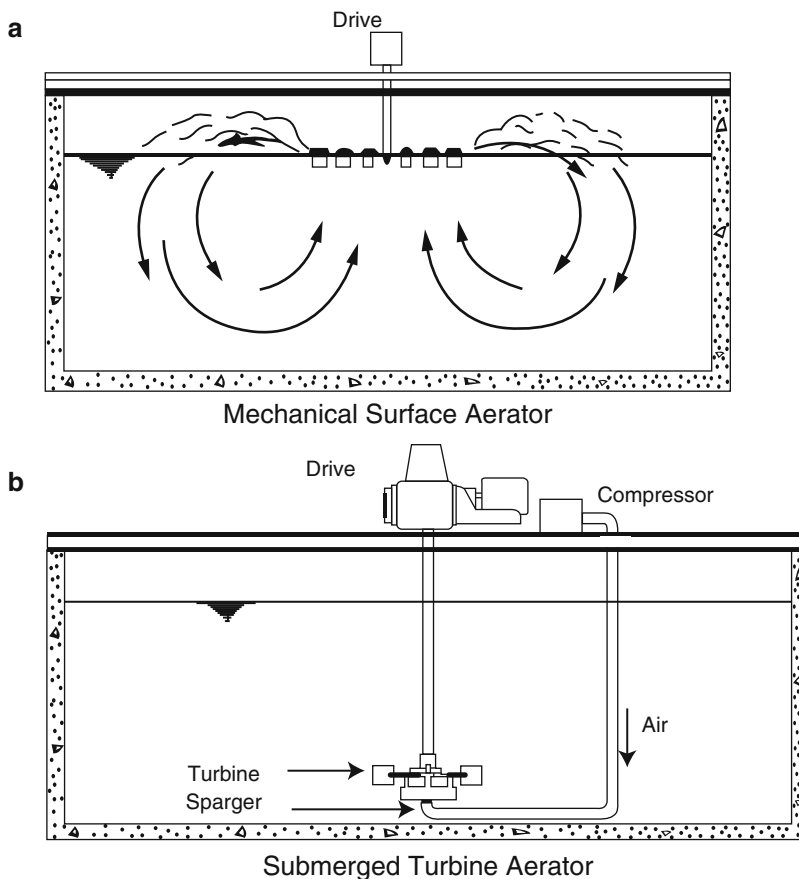


Fig. 13.5. Mechanical aeration (Source: U.S. EPA).

aeration basins. In addition, mechanical aerators may be either the floating or fixed installation type.

Mechanical aerators have been used primarily in industrial waste activated sludge treatment plants and are considered an attractive aeration system for very deep basins (with bottom mixers or spargers plus surface aerators), for activated sludges having high oxygen uptake rates, and for high concentrations of MLSS, as in aerobic digesters.

The mixing equipment for aeration or oxygen transfer must be sized to keep the solids in uniform suspension at all times. Depending on basin shape and depth, 4,000 mg/L of MLSS require about 0.75–1.0 HP/1,000 ft³ (0.02–0.03 kw/m³) of basin volume to prevent settling if mechanical aerators are employed. However, the power required to transfer the necessary oxygen will usually equal or exceed this value (1).

2. HIGH RATE ACTIVATED SLUDGE

2.1. Description

Activated sludge systems have traditionally been classified as high rate, conventional, or extended aeration (low rate), based on organic loading. The term modified aeration has been adopted to apply to those high rate air activated sludge systems with design F/M loadings in the range of 0.75–1.5 lb BOD₅/d/lb MLVSS (0.75–1.5 kg BOD₅/d/kg MLVSS). Modified aeration systems are characterized by low MLSS concentrations, short aeration detention times, high volumetric loadings, low air usage rates, and intermediate levels of BOD₅ and suspended solids removal efficiencies (1). Before the enactment of nationwide secondary treatment regulations, modified aeration was utilized as an independent treatment system for plants where BOD₅ removals of 50–70% would suffice. With present-day treatment requirements, modified aeration no longer qualifies as a “stand-alone” activated sludge option.

Modified aeration basins are normally designed to operate in either complete mix (Fig. 13.4) or plug flow (Fig. 13.1) hydraulic configurations. Either surface or submerged aeration systems can be employed to transfer oxygen from air to wastewater, although submerged equipment is specified more frequently for this process. Compressors are used to supply air to submerged aeration systems.

Primarily due to rapidly escalating power costs, interest has been expressed in the development of high rate, diffused aeration systems, which would produce a high quality secondary effluent. As with modified aeration, aeration detention times would remain low and volumetric loadings high. In contrast to modified aeration systems, high MLSS concentrations would have to be utilized to permit F/M loadings to be maintained at reasonable levels. The key to development of efficient high rate air systems is the availability of submerged aeration equipment that could satisfy the high oxygen demand rates that accompany high MLSS levels and short aeration times. New innovations in fine bubble diffuser and jet aeration technology offered the technology for uniting high efficiency oxygen transfer with high rate air activated sludge flow regimes to achieve acceptable secondary treatment as independent “stand-alone” processes (10).

Since the early 1970s, modified aeration was employed generally as a pretreatment or roughing process in a two-stage activated sludge system, where the second stage is used for

biological nitrification (10, 11). Alum or one of the iron salts is sometimes added to modified aeration basins preceding second-stage nitrification units for phosphorus removal.

2.2. Performance and Design Criteria

BOD₅ removal for modified aeration is in the range of 50–70%; for high solids, high rate air system, a removal of 85–95% is obtainable. Ammonia-N removal is only 5–10% (1, 7).

A modified air aeration system produces, on the average, 1.1 lb excess VSS (secondary effluent plus waste sludge)/lb BOD₅ removed at an average F/M ratio loading of 1.2 lb BOD₅/d/lb MLVSS. Here 1 lb/lb = 1 kg/kg; 1 lb/d/lb = 1 kg/d/kg.

Design criteria for the two high rate air activated sludge process options are summarized as follows (1, 7):

- (a) Modified aeration
 - Volumetric loading = 50–100 lb BOD₅/d/1,000 ft³ = 801.4–1602.8 kg BOD₅/d/m³
 - MLSS = 800–2,000 mg/L
 - Aeration detention time (based on influent flow) = 2–3 h
 - F/M = 0.75–1.5 lb BOD₅/d/lb MLVSS = 0.75–1.5 kg BOD₅/d/kg MLVSS
 - Air required = 400–800 Std. ft³ air/lb BOD₅ removed = 25–50 m³/kg BOD₅ removed
 - Oxygen required = 0.4–0.7 lb O₂/lb BOD₅ removed = 0.4–0.7 kg O₂/kg BOD₅ removed
 - Sludge retention time = 0.75–2 d
 - Recycle ratio (R) = 0.25–1.0
 - Volatile fraction of MLSS = 0.7–0.85
- (b) High solids, high rate aeration
 - Volumetric loading = 50–125 lb BOD₅/d/1,000 ft³ = 801.4–2003.5 kg BOD₅/d/m³
 - MLSS = 3,000–5,000 mg/L
 - Aeration detention time (based on influent flow) = 2–4 h
 - F/M = 0.4–0.8 lb BOD₅/d/lb MLVSS = 0.4–0.8 kg BOD₅/d/kg MLVSS
 - Air required = 800–1,200 Std. ft³ air/lb BOD₅ removed = 50–75 m³/kg BOD₅ removed
 - Oxygen required = 0.9–1.2 lb O₂/lb BOD₅ removed = 0.9–1.2 kg O₂/kg BOD₅ removed
 - Sludge retention time = 2–5 d
 - Recycle ratio (R) = 0.25–0.5
 - Volatile fraction of MLSS = 0.7–0.8

3. PURE OXYGEN ACTIVATED SLUDGE, COVERED

3.1. Description

The use of pure oxygen for activated sludge treatment has become competitive with the use of air due to the development of efficient oxygen dissolution systems (10). The covered oxygen system is a high rate activated sludge system. The main benefits cited for the process include reduced power requirements for dissolving oxygen in the wastewater, reduced aeration tank volume requirements, and improved biokinetics of the activated sludge system (11). In the covered system, oxygenation is performed in a staged, covered reactor, in which

oxygen gas is recirculated within the system until it reaches a reduced level of purity and a decreased undissolved mass at which it can no longer be used and is vented to the atmosphere. High-purity oxygen gas (90–100% volume) enters the first stage of the system and flows concurrently with the wastewater being treated through the oxygenation basin. Pressure under the tank covers is essentially atmospheric, being held at 2–4 in. (5.1–10.2 cm) water column, sufficient to maintain oxygen gas feed control and prevent backmixing from stage to stage. Effluent mixed liquor is separated in conventional gravity clarifiers, and the thickened sludge is recycled to the first stage for contact with influent wastewater (Fig. 13.6).

Mass transfer and mixing within each stage are accomplished either with surface aerators or with a submerged turbine rotating-sparg system. In the first case, mass transfer occurs in the gas space; in the latter, oxygen is sparged into the mixed liquor where mass transfer occurs from the oxygen bubbles to the bulk liquid. In both cases, the mass-transfer process is enhanced by the high oxygen-partial pressure maintained under the tank covers in each stage (12).

Volatile compounds are driven off to a certain extent in the oxygenation process and removed in the vent gas. Metals may also be expected to be partially removed, with accumulation in the sludge. High purity oxygen may be produced on-site by cryogenic or PSA (Pressure Swing Adsorption) generators, or purchased as liquid oxygen produced off-site and stored at the treatment plant. Cost effectiveness of oxygen source depends upon plant size and process train.

Although flexibility is claimed to permit operation in any of the normally used flow regimes, i.e., plug flow, complete mix, step aeration, and contact stabilization, the method of oxygen contact employed favors the plug flow mode. Process may be designed to achieve: optimum carbonaceous oxidation only, combined carbonaceous and nitrogenous oxidation or optimum nitrogenous oxidation as a separate stage after secondary treatment (11).

The pure oxygen process can be applied to both domestic and biologically degradable industrial wastewaters; for upgrading existing air activated sludge plants; for new facilities – to reduce construction cost where effective odor control is required, where high effluent dissolved oxygen is required, where reduced quantity and higher concentration of waste sludge is required and where reduced aeration detention time is required.

3.2. Performance and Design Criteria

Performance data for pure oxygen are summarized below (11, 12):

- (a) Carbonaceous Oxidation:
 - COD removal = 75–80%
 - BOD₅ removal = 90–95%
 - Suspended solids removal = 75–90%
- (b) Nitrogenous Oxidation – NH₄-N removals:
 - Single stage with carbonaceous oxidation = 20–90%
 - Separate stage nitrification after carbonaceous oxidation = 80–98%
- (c) Generated residuals 0.42–1.0lb VSS/lb BOD₅ removed. = 0.42–1.01 kgVSS/kgBOD₅ removed

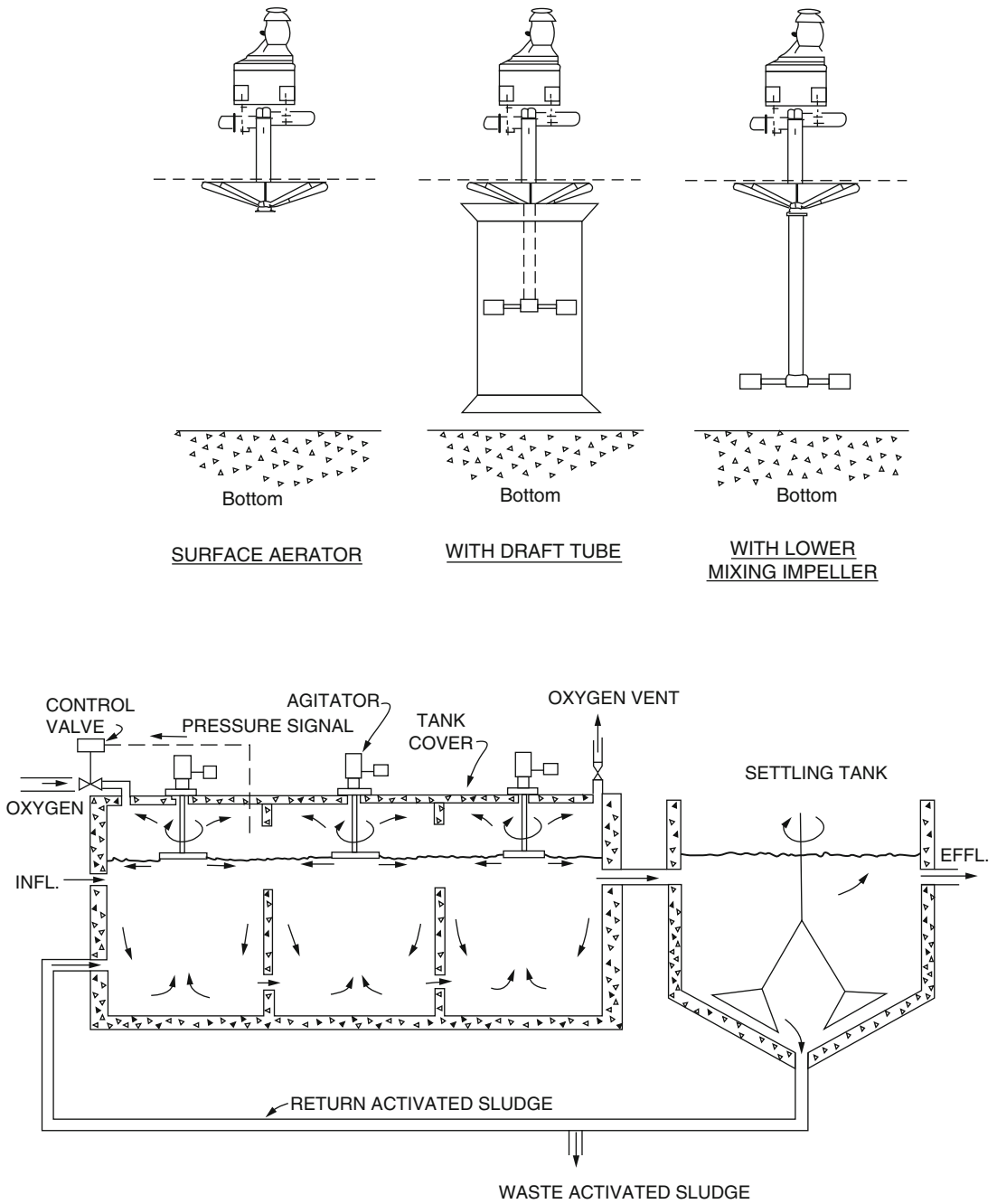


Fig. 13.6. Types of mechanical aerators and pure oxygen activated sludge (Source: U.S. EPA).

Design Criteria (Carbonaceous BOD Oxidation) (7, 11, 12):

- Volumetric loading = 100–200 lb BOD₅/d/1,000 ft³ = 1601.6–3203.6 g BOD₅/d/m³
- F/M = 0.5–1.0 lb BOD₅/d/lb MLVSS = 0.5–1.0 kg BOD₅/d/kg MLVSS
- Oxygen required = 0.6–0.8 lb O₂/lb COD removed = 0.6–0.8 kg O₂/kg COD removed
- MLSS = 3,000–6,000 mg/L
- Aeration detention time = 1–3 h
- Mixed liquor dissolved oxygen = 4–8 mg/L
- Oxygen required = 0.9–1.3 lb O₂/lb BOD₅ removed = 0.9–1.3 kg O₂/kg BOD₅ removed

4. CONTACT STABILIZATION

4.1. Description

Contact stabilization is a modification of the activated sludge process (see Fig. 13.3). In this modification, the adsorptive capacity of the floc is utilized in the contact tank to adsorb suspended, colloidal, and some dissolved organics. The hydraulic detention time in the contact tank is only 30–60 min (based on average daily flow). After the biological solids are separated from the wastewater in the secondary clarifier, the concentrated biosolids are separately aerated in the stabilization tank with a detention time of 2–6 h (based on solids recycle flow). The adsorbed organics undergo oxidation in the stabilization tank and are synthesized into microbial cells. If the detention time is long enough in the stabilization tank, endogenous respiration will occur, along with a concomitant decrease in excess biosolids production. Following stabilization, the reaerated biosolids are mixed with incoming wastewater in the contact tank and the cycle starts anew (1, 13). Volatile compounds are driven off to a certain extent by aeration in the contact and stabilization tanks. Metals will also be partially removed, with accumulation in the sludge.

This process requires smaller total aeration volume than the conventional activated sludge process. It can also handle greater organic shock and toxic loadings because of the biological buffering capacity of the stabilization tank and the fact that, at any given time, the majority of the activated sludge is isolated from the main stream of the plant flow. Generally, the total aeration basin volume (contact plus stabilization basins) is only 50–75% of that required in the conventional activated sludge system.

4.2. Applications

Contact stabilization has evolved as an outgrowth of activated sludge technology since 1950 and seen common usage in package plants and some usage for on-site constructed plants.

Contact stabilization can be most advantageously applied in the following cases (1, 13):

1. Wastewaters that have an appreciable amount of BOD₅ in the form of suspended and colloidal solids
2. Upgrading of an existing, hydraulically overloaded conventional activated sludge plant
3. New installations, to take advantage of low aeration volume requirements
4. Where the plant might be subject to shock organic or toxic loadings
5. Where larger, more uniform flow conditions are anticipated (or if the flows to the plant have been equalized)

Some of the limitations or disadvantages associated with contact stabilization include the following:

1. It is unlikely that effluent standards can be met using contact stabilization in plants with flow rates $< 50,000$ gal/d (189, 250 L/d) without some prior flow equalization
2. Operational complexity
3. High operating costs
4. High energy consumption and high diffuser maintenance
5. As the fraction of soluble BOD₅ in the influent wastewater increases, the required total aeration volume of the contact stabilization process approaches that of the conventional process

4.3. Performance and Design Criteria

Contact stabilization can achieve the following BOD₅ and NN₄-N removals (1, 14, 15):

- BOD₅ removal = 80–95%
- NN₄-N removal = 10–20%

Design criteria for the contact stabilization process are summarized as follows (1, 7, 14, 15):

- F/M = 0.2–0.6 lb BOD₅/d/lb MLVSS = 0.2–0.6 kg BOD₅/d/kg MLVSS
- Volumetric loading = 30–50 lb BOD₅/d/1,000 ft³ (based on contact and stabilization volume) = 481–801 g/d/m³
- MLSS = 1,000–2,500 mg/L, contact tank; 4,000–10,000 mg/L, stabilization tank
- Aeration time = 0.5–1.0 h, contact tank (based on average daily flow) 2–6 h, stabilization basin (based on sludge recycle flow)
- Sludge retention time = 5–10 d
- Recycle ratio (R) = 0.25–1.0
- Air supplied = 800–2,100 Std. ft³ air/lb BOD₅ removed = 50–131 m³ air/kg BOD₅ removed
- Oxygen required = 0.7–1.0 lb O₂/lb BOD₅ removed = 0.7–1.0 kg O₂/kg BOD₅ removed
- Volatile fraction of MLSS = 0.6–0.8

5. ACTIVATED SLUDGE WITH NITRIFICATION

5.1. Description

This process is also referred to as single-stage nitrification, because ammonia and carbonaceous materials are oxidized in the same aeration unit (the flow diagram is similar to Fig. 13.1). As in any aerobic biological process, carbonaceous materials are oxidized by heterotrophic aerobes. In addition, a special group of autotrophic aerobic organisms called nitrifiers oxidize ammonia in two stages: *Nitrosomonas* bacteria convert ammonia to nitrite and *Nitrobacter* convert nitrite to nitrate (1, 11).

The optimal conditions for nitrification, in genera are (1, 11, 16):

1. Temperature of about 30°C
2. pH of about 7.2–8.5
3. F/M of about 0.05–0.15
4. Relatively long aeration detention time as nitrifiers have a lower growth rate than other aerobe
5. Sludge retention time of about 20–30 d, depending upon temperature.

The degree of nitrification depends mainly on three factors (17):

1. SRT (sludge retention time), d
2. Mixed liquor DO concentration, mg/L
3. Wastewater temperature, °C

Of the above three factors, SRT is of primary importance because of the slow growth rate of nitrifiers. If the sludge is wasted at too high a rate, the nitrifiers will be eliminated from the system. Generally, nitrification begins at an SRT of about 5 d, but does not become appreciable until the SRT reaches about 15 d, depending upon temperature. The aeration system is designed to provide the additional oxygen needed to oxidize the ammonia nitrogen. Biological nitrification is very sensitive to temperature, resulting in poor reduction in colder months. In addition, heavy metals such as Cd, Cr, Cu, Ni, Pb and Zn, phenolic compounds, cyanide and halogenated compounds can inhibit nitrification reactions.

The conventional and high rate modifications of the activated sludge process do not provide the necessary hydraulic and sludge detention time. Besides, the F/M ratio is higher. As a result, single stage nitrification cannot be achieved in these configurations, although they effect a small reduction, about 20% in ammonia-N. Any low rate modification of the activated sludge process, such as the extended aeration and the oxidation ditch, can be used. In addition, the use of powdered activated carbon has the potential to enhance ammonia removal.

5.2. Performance and Design Criteria

A well-established extended aeration process will decrease ammonia-nitrogen to around 1 mg/L if the aerator temperature is about 55°F (1, 17).

This process produces no primary sludge. The secondary sludge is lesser in quantity and better stabilized than the high rate and conventional activated sludge process, which minimizes the magnitude of the disposal problem considerably.

The design criteria when using extended aeration modification are (5–7, 17):

- Volumetric loading = 5–10 lb BOD₅/d/1,000 ft³ = 80–160 gBOD₅/d/m³
- MLSS = 3,000–6,000 mg/L
- F/M = 0.05–0.15 lb BOD₅/d/lb MLVSS = 0.05–0.15 kg BOD₅/d/kg MLVSS
- Aeration detention time (based on average daily flow) = 18–36 h
- Air supplied = 3,000–4,000 std. ft³/lb BOD₅ applied = 187–250 m³/kg BOD₅ applied
- Oxygen required = 2.0–2.5 lb O₂/lb BOD₅ applied = 2.0–2.5 kg O₂/kg BOD₅ applied
- Sludge retention time = 20–30 d
- Recycle ratio = 0.7–1.5
- Volatile fraction of MLSS = 0.6–0.7

The design criteria when using oxidation ditch modification are:

- Volumetric loading = 10–15 lb BOD₅/d/1,000 ft³ = 160–240 g BOD₅/d/m³
- MLSS = 3,000–5,000 mg/L
- F/M = 0.03–0.10 lb BOD₅/d/lb MLVSS = 0.03–0.10 kg BOD₅/d/kg MLVSS
- Aeration detention time (based on average daily flow) = 24 h
- Oxygen required = 2.0–2.5 lb O₂/lb BOD₅ applied = 2.0–2.5 kg O₂/kg BOD₅ applied
- Sludge retention time = 20–30 d

- Recycle ratio = 0.25–0.75
- Volatile fraction of MLSS = 0.6–0.7 mg/L

6. SEPARATE STAGE NITRIFICATION

6.1. Description

The process by which ammonia is converted to nitrate in wastewater is referred to as nitrification. In the process, *Nitrosomonas* and *Nitrobacter* act sequentially to oxidize ammonia (and nitrite) to nitrate. The biological reactions involved in these conversions may take place during activated sludge treatment (as in previous section) or as a separate stage following removal of carbonaceous materials. Separate stage nitrification may be accomplished via suspended growth or attached growth unit processes. In either case, the nitrification step is preceded by a pretreatment sequence to reduce the carbonaceous demand. Possible pretreatment schemes include: activated sludge, trickling filter, roughing filter, primary treatment with chemical addition, and physical–chemical treatment. In general, if the pretreatment effluent has a BOD₅/TKN ratio of less than 3.0, sufficient carbonaceous removal has occurred such that the following nitrification process may be classified as a separate stage. Low BOD₅ is required to assure a high concentration of nitrifiers in the nitrification biomass (1, 11).

The most common separate stage nitrification process is the plug flow suspended growth configuration with clarification (Fig. 13.7). In this process, pretreatment effluent is pH adjusted (as required) and aerated in a plug flow mode. Because the carbonaceous demand is low, nitrifiers predominate. A clarifier follows aeration, and nitrification biosolids are returned to the aeration tank. A possible modification is the use of pure oxygen in place of conventional aeration during the plug flow operation.

Less prevalent are attached growth separate stage nitrification processes. These processes may be operated analogously to trickling filter, packed bed or rotating biological contactor systems. Since the biomass is attached to the reactor surface and solids synthesis is low, a clarifier may not be required. Final filtration is sometimes practiced to reduce effluent suspended solids, although this is often not required.

6.2. Performance and Design Criteria

Conversions of ammonia (and nitrite) to nitrate of up to 98% are achievable. Properly designed systems have effluent ammonia in the 1–3 mg/L range. BOD₅ reductions are generally 70–80% (influent BOD₅ assumed as approximately 50 mg/L) (18).

Design criteria – suspended growth systems (1, 5, 6, 17):

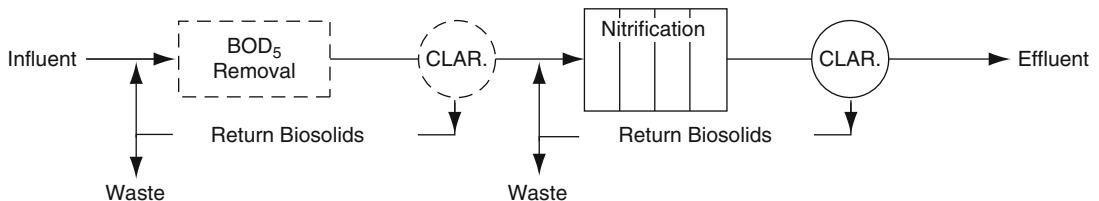


Fig. 13.7. Separate nitrification flow diagram (Source: U.S. EPA).

- Flow scheme: plug flow (preferable, but not mandatory)
- Optimum pH = 8.2–8.6
- MLVSS = 1,200–2,400 mg/L
- Min. aeration tank DO = 2.0 mg/L
- Clarifier surface loading rate = 400–600 gpd/ft² = 16.32–24.48 m³/d/m²
- Solids loading = 20–30 lb/d/ft² = 98–146 kg/d/m²
- Return biosolids rate = 50–100%
- Detention time = 0.5–3 h
- Mean cell residence time (MCRT) = 10–20 d

7. SEPARATE STAGE DENITRIFICATION

7.1. Description

Denitrification involves the reduction of nitrates and nitrites to nitrogen gas through the action of facultative heterotrophic bacteria (19, 20). In suspended growth, separate stage denitrification processes, nitrified wastewater containing primarily nitrates, is passed through a mixed anoxic vessel containing denitrifying bacteria. Since the nitrified feedwater contains very little carbonaceous material, a supplemental source of carbon is required to maintain the denitrifying biomass. This supplemental energy is provided by feeding methanol to the biological reactor along with the nitrified wastewater. Mixing in the anoxic denitrification reaction vessel may be accomplished using low speed paddles analogous to standard flocculation equipment. Following the reactor, the denitrified effluent is aerated for a short period (5–10 min) to strip out gaseous nitrogen formed in the previous step, which might otherwise inhibit sludge settling. Clarification follows the stripping step with the collected sludge being either returned to the head end of the denitrification system, or wasted. The flow diagram of this process is illustrated in Fig. 13.8.

Common modifications include the use of alternate energy sources, such as sugars, acetic acid, ethanol or other compounds. Nitrogen deficient materials, such as brewery wastewater, may also be used. An intermediate aeration step for stabilization (about 50 min) between the denitrification reactor and the stripping step may be used to guard against carryover of carbonaceous materials. The denitrification reactor may be covered but not air tight to assure anoxic conditions by minimizing surface reaeration (19).

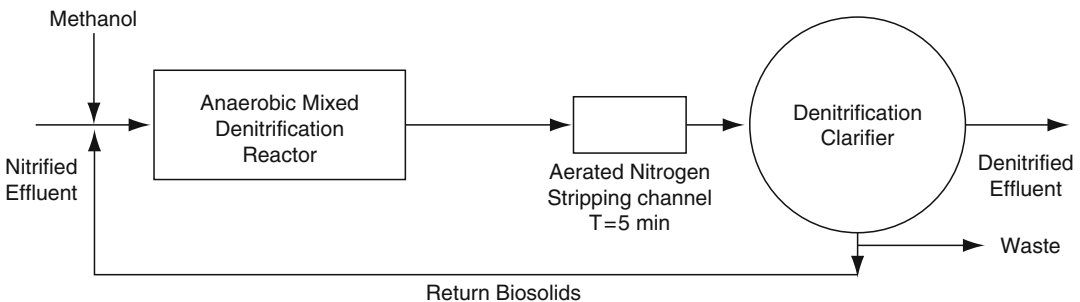


Fig. 13.8. Separate stage denitrification flow diagram (Source: U.S. EPA).

This process is used almost exclusively to denitrify municipal wastewaters that have undergone carbon oxidation and nitrification. It may also be used to reduce nitrate in industrial wastewaters.

7.2. Performance and Design Criteria

Separate stage denitrification is capable of reducing 50–98% of the nitrate and nitrite entering the system to gaseous nitrogen. Overall, nitrogen removals of 70–95% are achievable. Typical wastewater characteristics for $\text{NO}_3\text{-N}$: influent 20 mg/L, effluent 1 mg/L (18).

An energy source is needed which is usually supplied in the form of methanol. Methanol feed concentration may be estimated on the basis of 2.47 mg/L of methanol (CH_3OH) per mg/L of $\text{NO}_3\text{-N}$, 1.53 mg/L methanol/mg/L of $\text{NO}_2\text{-N}$ and 0.87 mg/L methanol/mg/L of DO (1, 11).

If supplemental energy feed rates are controlled, very little excess biosolids are generated. Biosolids production is in the range of 0.6–0.8 lb/lb (0.6–0.8 kg/kg) $\text{NH}_3\text{-N}$ reduced.

The design criteria for the denitrification process are listed below (5, 17):

- Flow scheme: plug flow (preferable, but not mandatory)
- Optimum pH = 6.5–7.5
- MLVSS = 1,000–3,000 mg/L
- Mixer power requirement = 0.25–0.50 HP/1,000 ft³ = 0.0066–0.0132 kW/m³
- Clarifier depth = 12–15 ft = 3.66–4.57 m
- Clarifier surface loading rate = 400–600 gpd/ft² = 16.32–24.48 m³/d/m²
- Solids loading = 20–30 lb/d/ft² = 98–146 kg/d/m²
- Return sludge rate = 50–100%
- Biosolids generation = 0.2 lb/lb CH_3OH or 0.7 lb/lb $\text{NH}_3\text{-N}$ reduced (1 lb/lb = 1 kg/kg)
- Detention time = 0.2–2 h
- Cell residence time = 1–5 d

8. EXTENDED AERATION

8.1. Description

Extended aeration is the “low rate” modification of the activated sludge process with the same flow diagram that was shown in Fig. 13.1. The F/M loading is in the range of 0.05–0.15 lb BOD_5 /d/lb MLVSS (0.05–0.15 kg BOD_5 /d/kg MLVSS), and the detention time is about 24 h (1). Primary clarification is rarely used. The extended aeration system operates in the endogenous respiration phase of the bacterial growth cycle because of the low BOD_5 loading. The organisms are starved and forced to undergo partial auto-oxidation. Volatile compounds are driven-off to a certain extent in the aeration process. Metals will also be partially removed, with accumulation in the sludge.

In the complete mix version of the extended aeration process, all portions of the aeration basin are essentially homogeneous, resulting in a uniform oxygen demand throughout the aeration tank. This condition can be accomplished fairly simply in a symmetrical (square or circular) basin with a single mechanical aerator or by diffused aeration. The raw wastewater and return biosolids enter at a point (e.g., under a mechanical aerator) where they are quickly

dispersed throughout the basin. In rectangular basins with mechanical aerators or diffused air, the incoming waste end return biosolids are distributed along one side of the basin and the mixed liquor is withdrawn from the opposite side ().

Extended aeration plants have evolved since the latter part of the 1940s. Most common applications are for plants with flows of less than 50,000 gpd (189,250 L/d) as well as for emergency or temporary treatment needs. Preengineered, package plants have been widely utilized for this process.

Some of the limitations or disadvantages of the process are (7):

- High power costs
- Operation costs
- Capital costs (for barge permanent installations where the preengineered plants would not be appropriate)

8.2. Performance and Design Criteria

The carbonaceous and ammonia-N removal rates are expected to be high since long detention times are employed in the process. Actual BOD₅ removal is in the range of 85–95% and NH₃-N removal between 50 and 90%.

Because of the low F/M loadings and long hydraulic detention times employed, excess biosolids production for the extended aeration process (and the closely related oxidation ditch process) is the lowest of any of the activated sludge process modifications, generally in the range of 0.15–0.30 lb excess biosolids/lb BOD₅ removed (7, 21). Here 1 lb/lb = 1 kg/kg.

The design criteria for the extended aeration modification of the activated sludge process is summarized as follows (1, 7, 22, 23):

- Volumetric loading = 5–10 lb BOD₅/d/1,000 ft³ = 80–160 g/d/m³
- MLSS = 3,000–6,000 mg/L
- F/M = 0.05–0.15 lb BOD₅/d/lb MLVSS = 0.05–0.15 kg BOD₅/d/kg MLVSS
- Aeration detention time = 18–36 h (based on average daily flow)
- Air required = 3,000–4,000 Std. ft³ air/lb BOD₅ applied = 187–250 m³ air/kg BOD₅ applied
- Oxygen required = 2.0–2.5 lb O₂/lb BOD₅ applied = (based on 1.5 lb O₂/lb BOD₅ removed + 4.6 lb O₂/lb NH₃-N removed) = 2.0–2.5 kg O₂/kg BOD₅ applied
- Biosolids retention time = 20–40 d
- Recycle ratio (R) = 0.75–1.5
- Volatile fraction of MLSS = 0.6–0.7

9. OXIDATION DITCH

9.1. Description

An oxidation ditch is an activated sludge biological treatment process; commonly operated in the extended aeration mode, although conventional activated sludge treatment is also possible. Typical oxidation ditch treatment systems consist of a single or closed loop channel 4–6 ft (1.22–1.83 m) deep, with 45° sloping sidewalls (Fig. 13.9).

Some form of preliminary treatment such as screening, comminution or grit removal normally precedes the process. After pretreatment (primary clarification is usually not practiced),

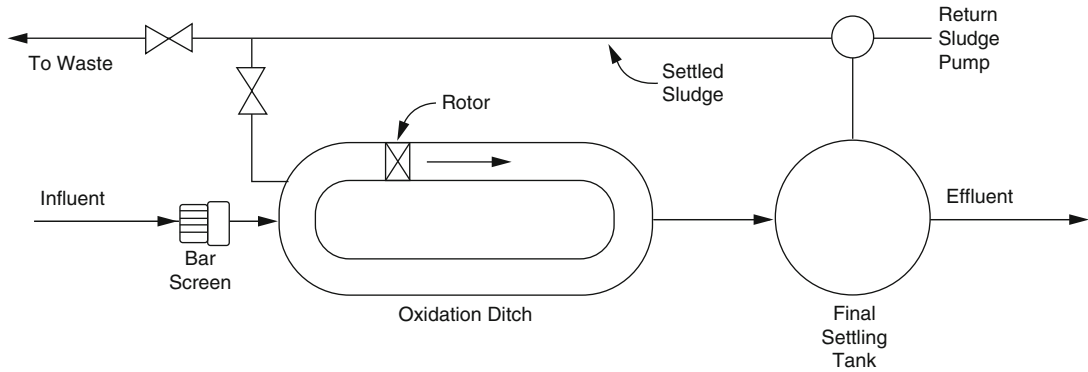


Fig. 13.9. Oxidation ditch (Source: U.S. EPA).

the wastewater is aerated in the ditch using mechanical aerators which are mounted across the channel. Horizontal brush, cage or disc-type aerators, specially designed for oxidation ditch applications, are normally used. The aerators provide mixing and circulation in the ditch, as well as sufficient oxygen transfer (24). Mixing in the channels is uniform, but zones of low, dissolved oxygen concentration can develop. Aerators operate in the 60–110 rpm range and provide sufficient velocity to maintain solids in suspension. A high degree of nitrification may occur in the process without special modification because of the long detention times and high solid retention times (10–50 d) utilized. Secondary Settling of the aeration ditch effluent is provided in a separate clarifier.

Ditches may be constructed of various materials, including concrete, gunite, asphalt, or impervious membranes. Concrete is the most common. Ditch loops may be oval or circular in shape. “Ell” and “horseshoe” configurations have been constructed to maximize land usage. Conventional activated sludge treatment, in contrast to extended aeration, may be practiced. Oxidation ditch systems with depths of 10 ft (3.05 m) or more with vertical sidewalls and vertical shaft aerators may also be used (25).

Oxidation ditch technology is applicable in any situation where activated sludge treatment (conventional or extended aeration) is appropriate. The process cost of treatment is generally less than other biological processes in the range of wastewater flows between 0.1 and 10 MGD (0.3785 and 3.785 MLD) (26).

9.2. Performance and Design Criteria

The average performance of shallow oxidation ditch plants is summarized below (24, 27–30):

- BOD₅ effluent = 10–15 mg/L
- BOD₅ removal = 90–95%
- Suspended solids effluent = 10–15 mg/L
- Suspended solids removal = 90–95%
- Ammonia–N removal = 40–80%

No primary biosolids are generated. Biosolids produced are less volatile due to higher oxidation efficiency and increased solids retention times.

Design criteria – extended aeration mode (24, 27–30):

- BOD_5 loading = 8.6–15 lb $BOD_5/d/1,000\text{ ft}^3$ of aeration volume = 138–240 g $BOD_5/d/m^3$
- Biosolids retention time = 10–33 d
- Channel depth = 4–6 ft = 1.22–1.83 m
- Channel geometry = 45° or vertical sidewalls
- Aeration channel detention time = 1 d

10. POWDERED ACTIVATED CARBON TREATMENT

10.1. Types of PACT Systems

The powdered activated carbon (PAC) activated sludge system is a process modification of the activated sludge process. PAC is added to the aeration tank where it is mixed with the biological solids (Fig. 13.10). The mixed liquor solids are settled and separated from the treated effluent. In a gravity clarifier, polyelectrolyte will normally be added prior to the clarification step to enhance solids–liquid separation. If phosphorus removal is necessary, alum is often added at this point also. Even with polyelectrolyte addition, tertiary filtration is normally required to reduce the level of effluent suspended solids. The clarifier underflow solids are continuously returned to the aeration tank. A portion of the carbon-biomass mixture is wasted periodically to maintain the desired solids inventory in the system (31).

There are six types of combined biological and physicochemical PAC process systems (32–39):

- (a) Continuous combined biological and physicochemical PAC process systems involving the use of sedimentation clarifiers

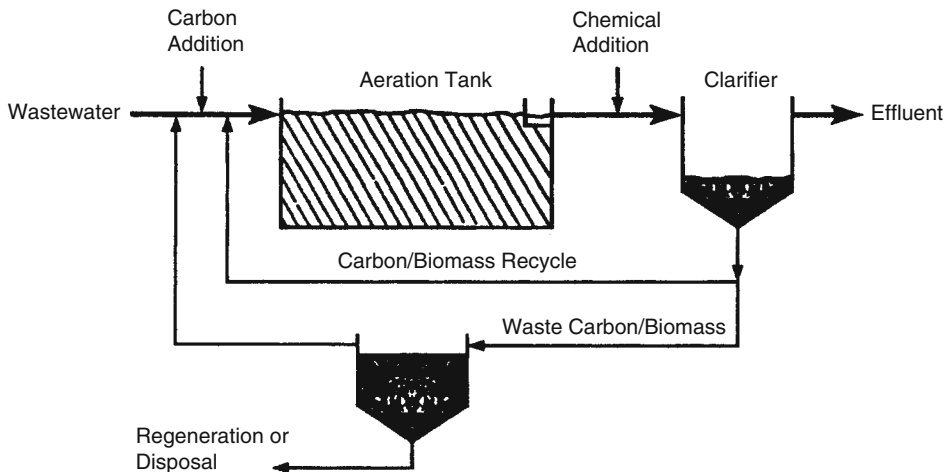


Fig. 13.10. Powdered activated carbon activated sludge process (PACT) (41, 45).

- (b) Combined biological and physicochemical PAC sequencing batch reactor systems involving the use of sedimentation clarifiers.
- (c) Continuous combined biological and physicochemical PAC process systems involving the use of dissolved air flotation (DAF) clarifiers
- (d) Combined biological and physicochemical PAC sequencing batch reactor systems involving the use of DAF clarifiers
- (e) Continuous combined biological and physicochemical PAC process systems involving the use of membrane filters (MF)
- (f) Combined biological and physicochemical PAC sequencing batch reactor involving the use of membrane filters (MF)

When PAC is dosed into an activated sludge process for combined adsorption and biochemical reactions, the combined process is also called powdered activated carbon treatment (PACT) process, in which PAC still stands for powdered activated carbon, while ACT stands for activated sludge.

10.2. Applications and Performance

The addition of PAC to plug flow and complete mix suspended growth reactors is a more common process modification for industrial wastewater treatment than for municipal systems. Demonstrated advantages of PAC addition to suspended growth reactors include (39):

- (a) Improved solids settling and dewatering characteristics
- (b) The ability of PAC to adsorb biorefractory materials and inhibitory compounds
- (c) Improving effluent quality and reducing the impact of organic shock loads
- (d) Reduction in odor, foaming, and sludge bulking
- (e) Improved color and 5-day BOD removal

Because PAC is wasted with excess biomass, virgin or regenerated PAC addition is required to maintain the desired concentration in the biological reactor. This can represent a significant cost factor for the system. When carbon addition requirements exceed 900–1,800 kg/d (2,400–4,000 lb/d), wet air oxidation/regeneration (WAR) is claimed to represent an economical approach to carbon recovery and waste biomass destruction (40). However, an ash separation step is needed in this case, affecting the economics of carbon regeneration and recovery (41). The economic analysis is further clouded by the inability to analytically differentiate powdered carbon from background refractory volatile materials, thus making it difficult to quantify the value of the volatile suspended material recovered after WAR. Although ash separation processes have been reported to be effective in at least two municipal PAC activated sludge plants, the economics of complete PAC/WAR systems relative to other activated sludge nitrification systems are unclear (38, 41, 42).

In the United States, PACT systems for nitrification have generally been applied at municipal treatment plants, where industrial sources contribute a significant fraction of the incoming wastewater. In all instances, PAC regeneration was included in the flowsheet (43). A summary of selected municipal PACT facilities is presented in Table 13.1.

The procedure to follow in designing PACT systems for nitrification involves a modification to those for complete mix or conventional plug flow systems in order to account for the effects of the addition of PAC (44). According to the major supplier of the technology (43, 45), most

Table 13.1
Summary of PACT process systems using wet air oxidation for PAC regeneration (41, 45)

Facility	Current design flow, m ³ /s	PAC/WAR status	Reason ^a for PAC	Permit Limits		
				BOD ₅ , mg/L	TSS, mg/L	NH ₄ -N, mg/L
Vemon, CT	0.18/0.28	MA	C	10	20	–
Mt. Holly, NJ	0.11/0.22	MA	C,S	30	30	20
E. Burlington, NC	0.31/0.53	MA	C,N,T	12–24	30	4.0–8.0
S. Burlington, NC	0.30/0.42	AS	C,N,T	12–24	30	4.0–8.0
Kalamazoo, MI	1.1/2.4	MA	C,N,T	7–30	20–30	2.0–10.0
Bedford Hts., OH	0.15/0.15	NAC	N,S	10	12	5.1
Medina Co., OH	0.31/0.44	MA	N	10	12	1.5–8.0
N. Olmsted, ^a OH	0.26/0.31	AS	N,S	30	30	2.3–6.9
Sauget, IL	0.70/1.2	AS	T	20	25	–
El Paso, TX	0.20/0.44	MA	N,O	Plan ^b	Plan	Plan

^aMA modified operation and/or design for ash control; AS converted to conventional activated sludge; NAC converted to the use of nonactivated carbon without regeneration.

C color removal; S space; N nitrification; T toxics; O organics.

^b Plan to convert to NAC without regeneration.

PAC process systems are designed at MLSS concentrations of approximately 15 g/L. The mixed liquor is composed of volatile activated carbon, biomass, nonvolatile PAC ash, biomass decay components, and influent inert material. The relative proportions of these materials are strongly influenced by whether carbon regeneration via wet air oxidation and a return of this material to the aerator is practiced. The intent is to maintain the PAC concentration at approximately 1.5 times the biomass level in nitrification PAC reactors (43, 45). The most appropriate PAC concentration will be dictated by the specific wastewater characteristics and often cannot be specified without bench or pilot scale studies. The PAC concentration to be added will depend on the design solids retention time, the hydraulic retention time, and the required PAC concentration in the reactor. According to the U.S. Environmental Protection Agency (45), for practical engineering design considering the loss, the PAC concentration to be added can be calculated from Eq. (1):

$$PACI = PACE + (PACR) HRT/SRT \quad (1)$$

where PACI is the influent PAC concentration (mg/L), PACR is the mixed liquor PAC concentration in the reactor (mg/L), PACE is the effluent PAC concentration (mg/L), HRT is the hydraulic retention time (d), and SRT is the design solids retention time (d).

The value of PACE in Eq. (1) can be estimated by assuming that the carbon fraction in the effluent TSS (total suspended solids) is the same as the fraction of PAC in the MLSS (mixed liquor suspended solids).

PACT nitrification systems are normally selected when the municipal wastewater contains compounds originating from industrial operations, as stated previously. Nitrifiers are susceptible to a number of organic and inorganic inhibitors found in many industrial wastewaters

(45). Researchers have provided evidence that the addition of PAC to nitrifying activated sludge systems receiving industrial wastewaters improved nitrification rates (45–47). More recent studies have been completed with the goal of determining the mechanism of nitrification enhancement in PAC activated sludge systems in the presence of adsorbable and nonadsorbable inhibitors (48). The results indicated that the addition of the proper amount of PAC can completely nullify the toxic effects of an adsorbable nitrification inhibitor. A minor positive effect on nitrification rates was observed when PAC was added to a nitrifying activated sludge system receiving nonadsorbable inhibitors. The activated sludge used in these studies was not acclimated to the inhibiting compounds. Another possible contributing factor to the enhancement of nitrification could be attributed to the fact that the addition of PAC provides particulate matter for attachment of the nitrifying microorganisms, thereby promoting nitrification (49).

10.3. Process Equipment

PAC can be fed in the dry state using volumetric or gravimetric feeders or can be fed in slurry form. There are more than three major PAC producers, over 50 manufacturers of volumetric and gravimetric feeders, and over 50 manufacturers of slurry feeders (50–52). There are also many manufacturers of sequencing batch reactors (SBR) (33), dissolved air flotation (DAF) clarifiers (38), and membrane filtration (MF) reactors (37).

10.4. Process Limitations

The process limitations of PACT Process Systems are identical to that of the PAC physicochemical process. PACT process will increase the amount of generated sludge. Regeneration will be necessary at higher dosages in order to maintain reasonable costs. Most systems will require postfiltration to capture any residual carbon particles. Some sort of flocculating agent, such as an organic polyelectrolyte, is usually required to maintain efficient solids capture in the clarifier.

About one pound of dry sludge will be generated per pound of carbon added. If regeneration is practiced, carbon sludge is reactivated and reused with only a small portion removed to prevent buildup of inert material. PAC physicochemical process systems are reasonably reliable. In fact, PAC systems can be used to improve process reliability of existing systems.

Additional information on carbon adsorption and combined biological and physicochemical PACT process systems can be found in refs. (53–62).

11. CARRIER-ACTIVATED SLUDGE PROCESSES (CAPTOR AND CAST SYSTEMS)

There has been a substantial interest in recent years in the potential benefits of high biomass wastewater treatment. The major obstacle for achieving this has been the inability of biosolids separation in secondary clarifiers. For the most part, this has been overcome by using various forms of support media or carriers that have the ability to attach high concentrations of aerobic bacterial growth (63–65). The increase in immobilized biomass reduces the process dependence on secondary settling basins for clarification. In such hybrid systems, where

attached growth coexist with suspended growth, one gets more stable systems which possess the combined advantages of both fixed and suspended growth reactors.

11.1. *Advantages of Biomass Carrier Systems*

The performance of carrier systems is dependent on the amount of attached biomass, the characteristics of attached and suspended microorganisms, and the type of carriers. The advantages of such hybrid systems are (31):

- (a) Heterogeneity of the microbial population. This is brought about by the differences in the microhabitat of organisms attached to the surface of a carrier and those in the bulk of the solution with respect to pH, ionic strength, and concentration of organics (66–70)
- (b) Increased persistence in reactor. This leads to increase in biomass of organisms, reduction of hydraulic retention time and thus smaller reactor volumes (71–73)
- (c) Higher growth rate (74–76)
- (d) Increased metabolic activity. This leads to increase in respiration and substrate utilization, hence higher removal rates (77–80)
- (e) Better resistance to toxicity (81–84)

11.2. *The CAPTOR Process*

One interesting concept of hybrid systems is the CAPTOR process developed jointly by the University of Manchester Institute of Science and Technology (UMIST) and Simon-Hartley, Ltd., in the United Kingdom. This high biomass approach uses small reticulated polyurethane pads as the bacterial growth medium (85). The pads are added to standard activated sludge aeration reactor, and the system is operated without sludge recycle, essentially combining suspended growth with a fixed film in one process. Excess growth is removed from the pads by periodically passing them through specially designed pressure rollers.

The British Water Research Centre (WRC) and Severn-Trent Water Authority conducted a full-scale evaluation of the CAPTOR process for upgrading the activated sludge plant at the Freehold Sewage Treatment Works, in the West Midlands area of England, to achieve year-round nitrification. This full scale study was jointly sponsored by the U.S. Environmental Protection Agency (86, 87).

11.3. *Development of CAPTOR Process*

As mentioned earlier, the CAPTOR process originated from research work on pure systems in the Chemical Engineering Department of UMIST. Single strands of stainless steel wire were woven into a knitted formation and then crushed into a sphere of about 6 mm (0.25 in.) diameter. These particles of known surface area were used for modeling liquid-fluidized bed systems. From this work derived the idea of using porous support pads for growing biomass at high concentrations that could be used in wastewater treatment systems. The idea was jointly developed and patented by UMIST and their industrial partner Simon-Hartley, Ltd. The present form of the CAPTOR process uses 25 mm × 25 mm × 12 mm (1 in. × 1 in. × 0.5 in.) reticulated polyether foam pads containing pores nominally of about 0.5–0.9 mm (0.02–0.035 in.) diameter and 94% free space (88–90).

11.4. Pilot-Plant Study

The conducted pilot-plant work indicated that it was possible to achieve the following (86, 87):

- (a) Biomass concentrations of 7,000–10,000 mg/L
- (b) Waste sludge concentrations of 4–6% dry solids using a special pad cleaner
- (c) Improved oxygen transfer efficiencies, and
- (d) High BOD volumetric removal rates

11.5. Full-Scale Study of CAPTOR and CAST

The full-scale evaluation of the CAPTOR process was undertaken at the Freehold Sewage Treatment Works near Stourbridge, West Midlands. The Freehold plant did not achieve any nitrification in the winter and only partial nitrification in the summer. Freehold's activated sludge system consisted of five trains equipped with tapered fine bubble dome diffusers arranged in a grid configuration. The system was modified as shown in Fig. 13.11 to split the wastewater flow into two equal volumes. Half went to two trains that were modified by adding CAPTOR pads to the first quarter of two aeration basins, and the other half went to two trains that remained unaltered and served as a control. The CAPTOR modified trains were each equipped with a CAPTOR pad cleaner (Fig. 13.12), and the CAPTOR pads were prevented from escaping into the remainder of the experimental system aeration basins by screens placed at the effluent ends of the CAPTOR zones.

The Simon-Hartley design predicted that, with a concentration of 40 pads/L, an annual average removal of 75% of the BOD₅ coming into the plant could be achieved in the CAPTOR zones, resulting in a reduced food-to-microorganism (F/M) loading on the follow-on activated sludge stage of 0.08 kg BOD₅/d/kg MLSS. With the reduced load, it was predicted that the modified system would achieve year-round nitrification with an effluent ammonia nitrogen concentration of 5 mg/L or less (87).

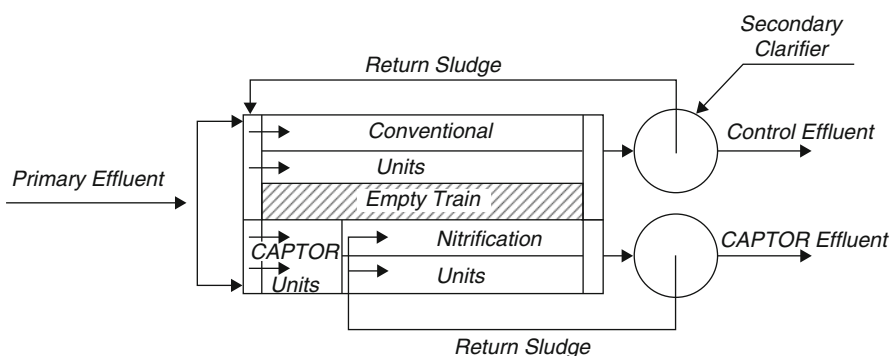


Fig. 13.11. Schematic of treatment plant showing incorporation of CAPTOR (87).

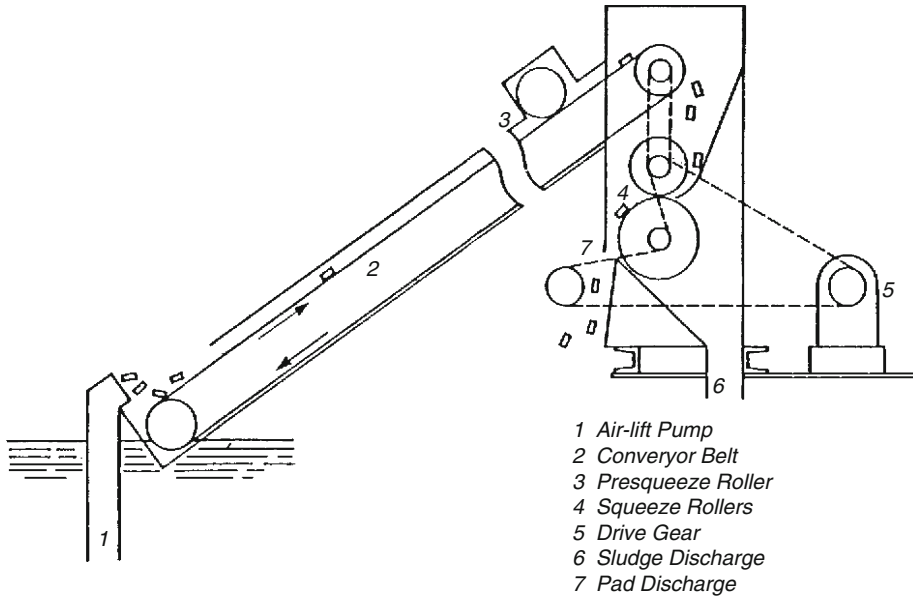


Fig. 13.12. CAPTOR pad cleaner (87).

11.5.1. Full-Scale Plant Initial Results

The Freehold modified CAPTOR activated sludge system was put in operation and immediately encountered a major problem. The CAPTOR pads floated on the surface of the tanks and would not become incorporated into the tank liquor. A solution was found by removing three of the seven longitudinal rows of fine bubble diffusers in the CAPTOR aeration basins. This was done to create a spiral roll in the tanks, which leads to areas of rising and falling liquid with quite large channels down, where the pads can fall. The spiral roll modification provided the necessary falling zone and produced complete mixing of the CAPTOR pads.

Another problem that occurred was mal distribution of the pads. The flow of wastewater tended to push the CAPTOR pads to the outlet of their zones, resulting in a concentration of 50–60 pads/L at the outlet and only 10–20 pads/L at the inlet end.

An other disturbing feature was the rapid deterioration in the CAPTOR pads. The CAPTOR pads used initially were black and were wearing at such a rate that they would not have lasted for more than 3 years, rendering the process uneconomical.

It had also become evident by this time that with the Freehold wastewater it would be possible to achieve the concentration of 200 mg biomass/pad predicted in the design. However, it was found that if the biomass was allowed to grow beyond 180 mg/pad, the biomass in the center of the pad became anaerobic. The control of pad biomass was difficult because the pad cleaners provided were not reliable and were situated at the CAPTOR zone inlets, while most of the pads gravitated to the outlet ends of the zones.

During this early period, while the above problems were being tackled on the full-scale plant, there were some occasions when the effluent from the CAPTOR units was reasonable

(BOD removals of 40–50%), but BOD removal never approached the average of 75% predicted based on the earlier pilot-plant results. Poor BOD removals were being experienced because the suspended solids concentration in the effluent was always high (> 80 mg/L).

Consequently more pilot-scale studies were used to find solutions to the operating problems described above before attempting further full-scale evaluation at Freehold.

11.5.2. Pilot-Scale Studies for Project Development

It was decided to evaluate two variations of the CAPTOR process. The new variation differed from the original CAPTOR, in that the pads were placed directly into the mixed liquor of the activated sludge aeration tank rather than in a separate stage before the activated sludge tank. WRC named this process variation CAST (CAPTOR in activated sludge treatment). The CAST system had been applied to upgrade several overloaded wastewater treatment plants in Germany and France, and was found to be useful in improving the treatment efficiency and plants performance (91–93).

In addition, a single aeration tank filled with 40 CAPTOR pads/L was fed effluent from the above activated sludge control unit to assess the potential of CAPTOR as a second-stage nitrification process. Neither pad cleaning nor final clarification was necessary with this process variation because of the low sludge yields characteristic of nitrifier growth.

Studies were conducted using two well-mixed CAPTOR tanks in series. A range of loading and pad cleaning rates were used to evaluate process removal capabilities for CAPTOR. The intermediate effluent was used as a measure of process efficiency of the primary reactor and the final effluent for the entire system. This permitted plotting (Fig. 13.12) of % BOD₅ removal (total and soluble) vs. volumetric organic loading rate over the range of 1–3.5 kg BOD₅/d/m³ (62–218 lb/d/1,000 ft³). High and low pad cleaning rates are differentiated in Fig. 13.13 as $\geq 16\%$ and $< 16\%$ of the total pad inventory/d, respectively (87).

Total BOD₅ removal efficiency was less than soluble BOD₅ removal efficiency because of the oxygen demand exerted by the biomass solids lost in the process effluent. The higher pad cleaning rates are believed to have contributed to the improved total and soluble BOD removals shown in Fig. 13.13, although low bulk liquid DO's may have adversely affected removals on some of the low cleaning runs. Low cleaning rates ($< 16\%/d$) were detrimental to soluble BOD₅ removal efficiency because of a gradual decline in activity of the biomass remaining in the pad. Cleaning rates greater than 24%/day, however, resulted in reduced biomass levels in the pads and a reduction in performance.

The problem of mal distribution of CAPTOR pads in the aeration tank (i.e., crowding of pads into the effluent end of the tank when operated in plug flow fashion as at Freehold) was solved by modifying the flow pattern to transverse flow (across the width of the tank rather than down the length). When implemented later at Freehold, this pattern resulted in a fourfold decrease in flow velocity.

Several mixing intensities and diffuser arrangements were tried to decrease biomass shedding into the process effluent. It became obvious, however, that production of effluent biomass solids was not significantly affected by changes in mixing intensity or diffuser arrangement.

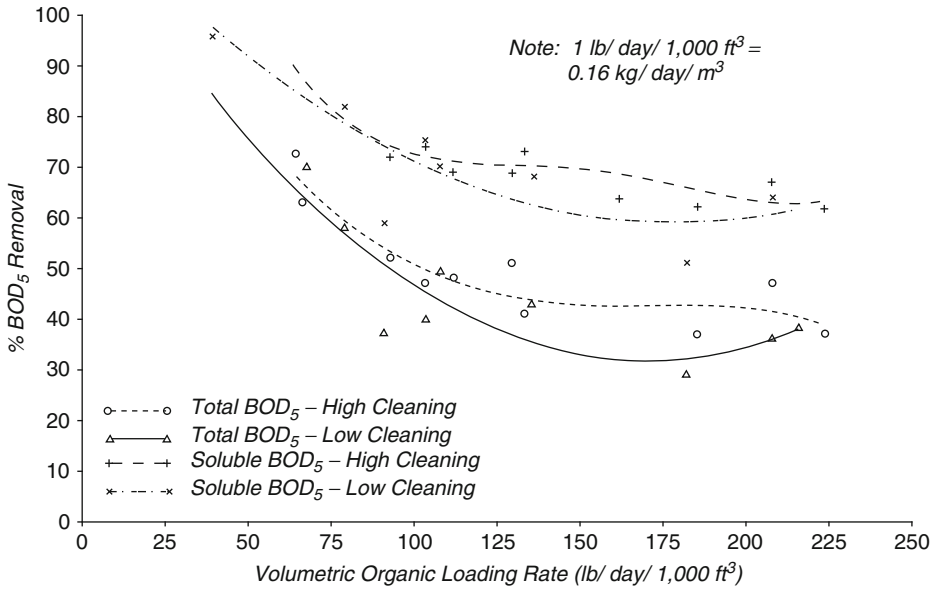


Fig. 13.13. Pilot-scale CAPTOR BOD₅ removals as a function of organic loading rate (87).

High effluent suspended solids proved to be far more dependent on pad cleaning rate, biochemical activity of the biomass, and biomass growth directly in the liquor.

Using the transverse flow scheme and a regular pad cleaning regimen, CAPTOR process performance was similar to that experienced in the small tanks. Operating parameters and process performance are summarized in Table 13.2 for two different volumetric loading rates (87).

Respiration studies conducted on pads indicated that biomass held within the pads respire at up to 40–50% less than equivalent biomass in free suspension. Any increase in net biomass concentration achieved in a CAPTOR reactor above that in a conventional activated sludge reactor may not produce noticeable benefits, therefore, due to the lower specific activity. These observations suggest that diffusion limitations were occurring in the CAPTOR pads.

The CAST variation of CAPTOR was operated in conjunction with a final clarifier to settle the mixed liquor solids component of the total biomass inventory and return it to the aeration tank. CAPTOR pads and biomass retained therein were kept in the reactor by screens. Operating and performance data are compared in Table 13.3 for the CAST unit and the parallel activated sludge control unit for a 25-day period when the volumetric loadings and hydraulic residence times (HRTs) for both units were identical.

In the nitrification experiments conducted on the CAPTOR process, the biomass concentrations per pad ranged from 99 to 124 mg. This is within the range of 100–150 mg/L reported by other researchers (94). With a pad concentration of 40/L, equivalent MLSS levels varied from 3,960 to 4,960 mg/L. Liquor DO concentrations were maintained between 6.4 and 8.4 mg/L, and liquor temperature ranged from 11.50 to 6.5°C.

Table 13.2
Pilot-scale operating conditions and process performance (87)

Parameter	Period			
	1		2	
Volumetric loading (lbBOD ₅ /d/1,000 ft ³) ^a	113		213	
HRT(h)	2.32		1.52	
Pads/L	40		40	
Biomass/pad(mg)	121		126	
Equivalent MLSS (mg/L)	4,840		5,040	
F/M loading (kg BOD ₅ /d/kg MLSS)	0.37		0.68	
SRT (days)	3.23		1.72	
DO (mg/L)	4.2		4.7	
	In	Out	In	Out
Total BOD ₅ (mg/L)	175	93	216	129
Soluble BOD ₅ (mg/L)	86	24	85	33
SS (mg/L)	116	120	178	160
Total BOD ₅ removal (%)	47		40	
Soluble BOD ₅ removal (%)	72		61	
SS removal (%)	-3		10	

$$^a 1 \text{ lb/d/1,000 ft}^3 = 0.016 \text{ kg/d/m}^3$$

Secondary effluent from the control activated sludge pilot unit used in the CAST experiments was applied to the nitrification reactor over a range of loading conditions. Essentially, complete nitrification was achieved at TKN and ammonia nitrogen loadings of approximately 0.25 kg/d/m³ (15.6 lb/d/1,000 ft³) and 0.20 kg/d/m³ (12.5 lb/d/1,000 ft³), respectively.

11.5.3. Full-Scale Plant Results After Modifications

Following the successful testing of the transverse mixing arrangement in the pilot-scale study, the two Freehold CAPTOR trains were modified. The modifications involved the following (87):

- Splitting each of the CAPTOR trains, C1 and C2, into two compartments, C1A and C1B and C2A and C2B, as shown in Fig. 13.14
- Feeding influent flow along long weirs at the side of the trains instead of at the narrow inlet ends
- Modifying the aeration pipe work to place all three rows of dome diffusers directly below the outlet screens (covering about 25% of the width of the tanks), thereby creating a spiral roll of pads and liquid counter-current to the flow of wastewater entering along the weirs on the sidewalls
- Installing two extra pad cleaners so that each CAPTOR subunit was provided with a cleaner, and

Table 13.3
Pilot-scale CAST and activated sludge operating conditions and performance (87)

Parameter	System			
	CAST		Activated sludge	
Volumetric loading (lb BOD ₅ /d/1,000 ft ³) ^a	148		148	
HRT (h)	1.8		1.8	
Pads/L	34		–	
Biomass/pad (mg)	116		–	
Equivalent MLSS in pads (mg/L)	3,930		–	
MLSS in suspension (mg/L)	3,720		6,030	
Total MLSS (mg/L)	7,650		6,030	
F/M loading (kg BOD ₅ /d/kg total MLSS)	0.31		0.39	
SRT, based on total MLSS (days)	3.6		3.0	
DO (mg/L)	2.5		3.0	
	In	Out	In	Out
Total BOD ₅ (mg/L)	178	12	178	20
Soluble BOD ₅ (mg/L)	101	5	101	4
SS (mg/L)	121	15	121	23
Total BOD ₅ removal (%)	93		89	
Soluble BOD ₅ removal (%)	95		96	
SS removal (%)	88		81	

$$^a 1 \text{ lb/d/1,000 ft}^3 = 0.016 \text{ kg/d/m}^3$$

- (e) Installing fine screens at the outlet from the primary clarifiers to reduce the quantity of floating plastic material entering the CAPTOR units that created problems with the cleaners

The objective of the first three modifications was to achieve uniform mixing of the pads in the CAPTOR units and prevent the situation that had occurred previously where high concentrations of pads (50–60 pads/L) collected at the outlet end and very low concentrations (10–20 pads/L) at the inlet end. Pads were removed from the tanks during the modifications. After the modifications were completed, the number of pads in each compartment was equalized at about 35/L.

The changes were completely successful in obtaining uniform distribution and complete mixing of the CAPTOR pads. A lithium chloride tracer test conducted on the modified tanks indicated that no dead zone was occurring in the “eye” of the roll. Formation of floating pad rafts (which had occurred at the outlet end of the tank with the original arrangement) was completely eliminated. The modifications, however, had no effect on the high level of suspended solids present in the liquor. The modified CAPTOR system was operated at an

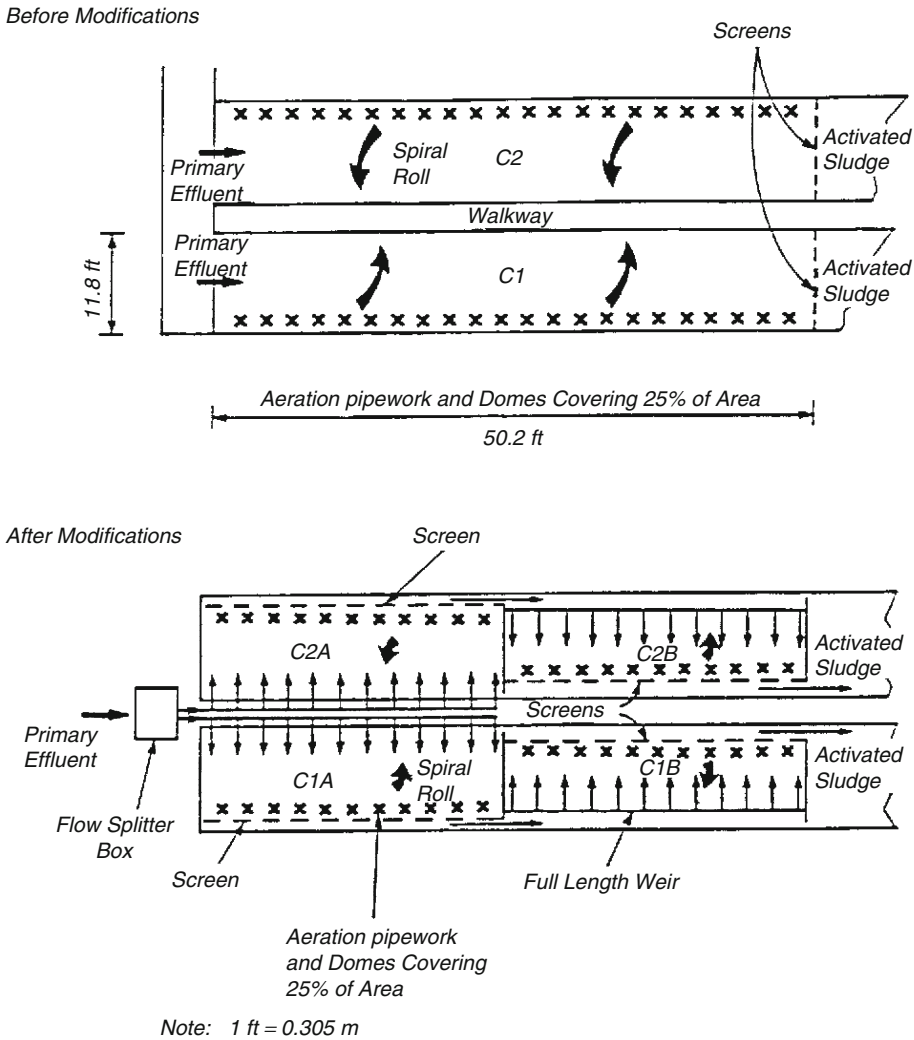


Fig. 13.14. Modifications to full-scale CAPTOR system flow pattern (87).

average volumetric loading rate of $1.24 \text{ kg BOD}_5/\text{d}/\text{m}^3$ ($77 \text{ lb}/\text{d}/1,000 \text{ ft}^3$), an average HRT (excluding sludge recycle) of 2.55 h and an overall biomass concentration of $4,830 \text{ mg}/\text{L}$.

The CAST variation of the CAPTOR process, which had exhibited somewhat better performance than conventional activated sludge in the small tank experiments, was also field evaluated at Freehold. The CAPTOR trains were further modified so that return sludge could be introduced to the CAPTOR zones (35 pads/L), providing an activated sludge component throughout the entire aeration tanks, not just in the nitrification stage. The average volumetric organic loadings and HRTs (excluding sludge recycle) were $1.11 \text{ kg BOD}_5/\text{d}/\text{m}^3$ ($69 \text{ lb}/\text{d}/1,000 \text{ ft}^3$) and 3.40 h, respectively.

Table 13.4
Full-scale modified CAPTOR performance results (87)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
Total BOD ₅	128	22	83
Soluble BOD ₅	40	4	90
SS	138	32	77
NH ₄ -N	24	24.4	0

Table 13.5
Full-scale modified CAST performance results (87)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
Total BOD ₅	138	16	88
Soluble BOD ₅	56	2	96
SS	120	27	78
NH ₄ -N	26.7	17.2	36

Performance data summarized in Tables 13.4 and 13.5 indicate that the CAST system exhibits somewhat better performance than the CAPTOR version. In the CAST process, the removal of soluble BOD₅ is 96% compared to 90% in CAPTOR; the removal of total BOD₅ is 88% compared to 83%; and the removal of SS is about the same at about 78%.

11.5.4. Overall Conclusions

The US EPA conclusions and recommendations for the CAPTOR/CAST treatment systems are as follows (86, 87, 95):

- (a) In the initial phase, when the CAPTOR process was installed at the Freehold Sewage Treatment Works, several problems were immediately evident. There were major problems with respect to pad mixing, suspension, and distribution, and the process performance was adversely affected by the high-level of suspended solids in the CAPTOR stage effluent. The problems of pad mixing and distribution were solved by pilot- and full-scale development work
- (b) The performance of the CAPTOR process was still adversely affected by the high level of suspended solids in the CAPTOR stage effluent after correction of the pad mixing, suspension, and distribution problems. This prevented the achievement of nitrification in the follow-on activated sludge stage
- (c) The presence of CAPTOR pads in the tank liquid did not improve oxygen transfer efficiency
- (d) The durability of the CAPTOR pads was solved by switching to different pads
- (e) The peak biomass concentration in the pads is unpredictable. It does not appear to be related to the BOD concentration of the wastewater. There were indications in the various studies, however, that the frequency of pad cleaning (and, hence, the biomass/pad concentration) was critical to the performance of the process. Regular pad cleaning is essential to prevent anaerobic conditions from developing in the pads
- (f) It is possible to raise the biomass concentration in a CAPTOR stage to 6,000–8,000 mg/L, but the respiration rate of the biomass in the pads is lower than the respiration of the same biomass if freely suspended and less than that of normal activated sludge. These data suggest that the

geometry of the CAPTOR pads results in diffusion limitations, which demands further pad design improvement to enhance the potential for economic utilization of the CAPTOR process in wastewater treatment

- (g) The CAST variation of the CAPTOR process performs well
- (h) CAPTOR has the potential as an add-on package for tertiary nitrification
- (i) The CAPTOR option was projected to be more cost effective than extending the activated sludge plant for upgrading Freehold to complete year-round nitrification.
- (j) For CAPTOR and CAST to achieve their full potential, as predicted by the pilot-scale studies, further design development and improvements are needed.

12. ACTIVATED BIO-FILTER

12.1. Description

Activated bio-filters (ABF) are a recent innovation in the biological treatment field. This process consists of the series combination of an aerobic tower (biocell) with wood or other packing material, followed by an activated sludge aeration tank and secondary clarifier. Settled sludge from the clarifier is recycled to the top of the tower. In addition, the mixture of wastewater and recycle sludge passing through the tower is also recycled around the tower, in a similar manner to a high rate trickling filter (31). No intermediate clarifier is utilized. Forward flow passes directly from the tower discharge to the aeration tank (Fig. 13.15). The use of the two forms of biological treatment combines the effects of both fixed and suspended growth processes in one system. The microorganisms formed in the fixed growth phase are passed along to the suspended growth unit, whereas the suspended growth microorganisms are recycled to the top of the fixed media unit (96). This combination of the two processes results in the formation of a highly stable system that has excellent performance and good settling biological floc when treating wastewaters that have variable loads (97).

The biomedia in the biocell consists of individual racks made of wooden laths fixed to supporting rails. The wooden laths are placed in the horizontal direction, permitting wastewater to pass downward, and air horizontally and vertically. The horizontal surfaces reduce premature

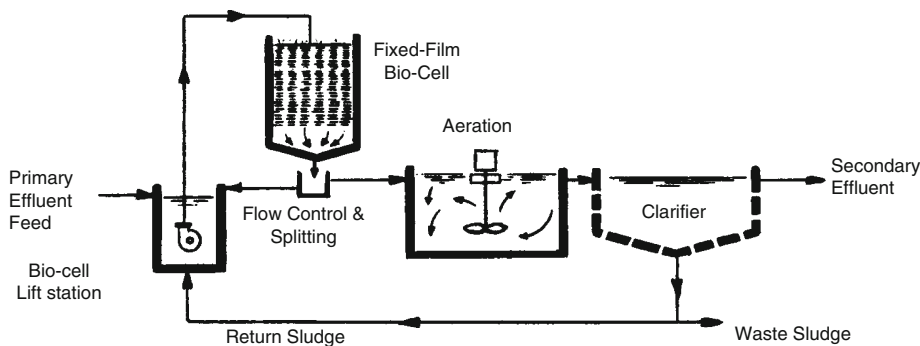


Fig. 13.15. ABF process flow diagram (96).

sloughing of biota. Droplet formation and breakup induced by wastewater dripping from lath to lath enhances oxygen transfer. Other types of material for the biomedial have also been reported by other researchers and equipment manufacturers (98–101). The aeration basin is a short detention unit that can be designed for either plug flow or complete mix operation. The effluent from the aeration basin passes to a secondary clarifier where the activated sludge is collected and recycled to the top of the biocell tower and to waste.

ABF units can be used for the removal of either carbonaceous material or for carbonaceous removal plus nitrification by appropriately modifying the detention time of the aeration basin. When nitrification is desired, the biocell acts as a first-stage roughing unit and the aeration basin as a second-stage nitrification unit (102, 103). ABF biocells can be either rectangular or round. Various types of aeration equipment can be used in the aeration system, including both surface and diffused aerators. The detention time of the aeration tank can be modified, depending on influent quality and desired effluent quality. ABF units can be supplied with mixed media effluent filters for enhanced treatment.

12.2. Applications

Activated bio-filters can be used for treating municipal wastewater and biodegradable industrial wastewater. ABF systems are especially useful where (96, 97):

- (a) Both BOD₅ removal and nitrification are required
- (b) Land availability is low
- (c) Raw wastewater organic loadings fluctuate greatly, due to its ability to handle shock conditions
- (d) Existing trickling filter facilities and overloaded existing secondary plants need to be upgraded at reduced cost

A typical ABF application is the Burwood Beach Wastewater Treatment Works in Australia (104). The plant was upgraded in the 1990s using ABF at a cost of \$48M. The facility currently serves a population of 180,000 with a flow of 43ML a day, and has the capacity to treat 53 ML/day for a population of 220,000 in the year 2020. The Biofilter is 30 m in diameter and has a design organic loading of 3.2 kg BOD₅/m³/d. The aeration tank is designed for 1.5 h of hydraulic detention time. The plant has been in operation for around 10 years producing an effluent that is consistently within the required US EPA set limits.

12.3. Design Criteria

The design criteria for the ABF system are reported to be as follows (96, 105, 106):

- (a) Biocell organic load: 100–200 lb BOD₅/d/1,000 ft³ = 1.6–3.2 kg BOD₅/d/m³
- (b) Return sludge rate: 25–100%
- (c) Biocell recycle rate: 0–100%
- (d) Biocell hydraulic load: 1–5.5 gpm/ft² = 40.7–224 Lpm/m²
- (e) Aeration basin detention time: 0.5–3.0 h for BOD₅ removal only, 5.8–7.5 h for two-stage nitrification
- (f) System F/M: 0.25–1.5 lb BOD₅/d/lb MLVSS for BOD removal, 0.18 lb BOD₅/d/lb MLVSS for two-stage nitrification. Here 1 lb/d/lb = 1 kg/d/kg.

Table 13.6
Performance of BAF systems (96)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
BOD ₅	153	14	91
COD	330	58	82
TSS	222	20	91
NH ₄ -N ^a	20	1	90

^aWhen used for nitrification.

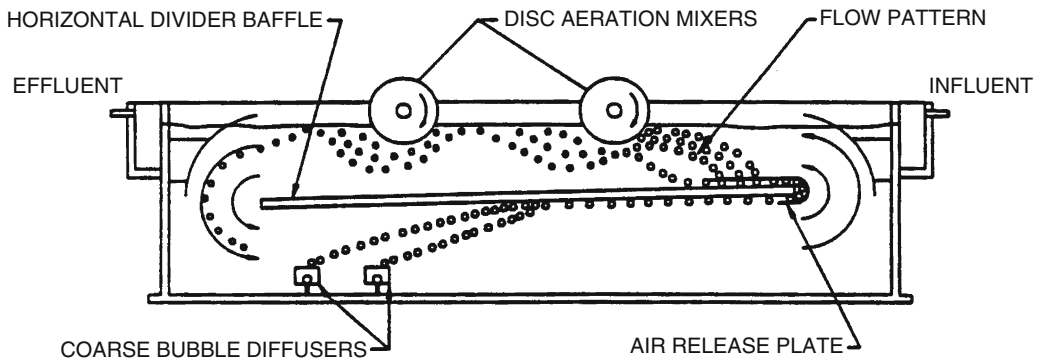


Fig. 13.16. Diagram of the vertical loop reactor (108, 109).

12.4. Performance

ABF systems are quite stable and highly reliable. They can treat standard municipal, combined municipal/industrial, or industrial wastewaters to BOD₅ and suspended solids levels of 20 mg/L or less. Test study on a package system showed at least 90% removal of BOD₅, TSS and NH₄-N (96). The detailed results are shown in Table 13.6.

Sludge production was reported at 0.25–1.0 lb of waste VSS per lb of BOD₅ removed. The mean yield, over the course of the study, was 0.60 lb VSS per lb of BOD removed. Here 1 lb/lb = 1 kg/kg.

13. VERTICAL LOOP REACTOR

13.1. Description

A vertical loop reactor (VLR) is an activated sludge biological treatment process similar to an oxidation ditch (107, 108). The wastewater in an oxidation ditch circulates in a horizontal loop; the water in a VLR circulates in a vertical loop around a horizontal baffle, as shown in Fig. 13.16 (109). A typical VLR consists of an 18 ft deep concrete or steel basin, with a horizontal baffle extending the entire width of the reactor and most of its length. Operating basins are reported to have side-wall depths which range from approximately 10–22 ft (3.05–6.71 m) (110). The length and width of the VLR are determined by the required capacity but, as a rule, the length is at least twice the width. The baffle is generally 5–11 ft (1.52–3.35 m)

below the surface of the water. Because a VLR is typically deeper than an oxidation ditch, the VLR requires less land area (31).

Aeration in a VLR is provided by coarse bubble diffusers, which are located below the horizontal baffle and by disc aeration mixers. The disc mixers also circulate the wastewater around the baffle at a velocity of 1–1.5 ft/s (0.30–0.46 m/s) (111). Because the diffusers are positioned below the baffle, the air bubble residence time in a VLR is as much as six times longer than the bubble residence time in a conventional aeration system. This extended bubble contact time increases the process aeration efficiency. Denitrification in an anoxic zone also reduces oxygen requirements.

The VLR process is usually preceded by preliminary treatment such as screening, comminution or grit removal. Secondary settling of the VLR effluent is typically provided by a separate clarifier. An intrachannel clarifier may be used for secondary settling in place of a separate clarifier.

Vertical loop reactors may be operated in parallel or series. When a series of VLRs are used, the dissolved oxygen profile can be controlled to provide nitrification, denitrification, and biological phosphorus removal at hydraulic detention times of 10–15 h.

13.2. Applications

VLR technology is applicable in any situation where conventional or extended aeration activated sludge treatment is appropriate. The technology is applicable for nitrification and denitrification. Biological phosphorus removal may be incorporated in the system design. Power costs may be lower for a VLR system than for other aerated biological treatment systems, due to improved oxygen transfer efficiency. There are currently more than ten municipal wastewater treatment facilities in the United States with VLRs. One such example is the City of Willard, OH waste water treatment plant (112). The facility is designed for an average daily flow of 4.5 MGD (17.03 MLD), and is capable of handling a peak flow of 7.2 MGD (27.25 MLD).

The following advantages have been reported for VLR systems (113):

- (a) Land area required for VLRs is about 40% less than for oxidation Ditches
- (b) The VLR aeration basin cost is about 30% less than for oxidation ditches
- (c) The multiple tank basin series arrangement is an advantage for facilities with highly variable flow
- (d) VLRs are useful for retrofitting existing basins for plant upgrade to suit increased flows or more stringent effluent requirements

13.3. Design Criteria

The design criteria for the VLR process are reported to be as follows (107):

- BOD loading: 14–22 lb BOD₅/1,000 ft³/d = 224–353 g BOD₅/d/m³
- SRT: 17–36 d
- Detention time: 12–24 h

13.4. Performance

The average effluent BOD₅ and TSS concentrations for the five studied operating VLR facilities are 4.2 and 7.1 mg/L, respectively. The average effluent ammonia concentration is 0.8 mg/L. Only one of the VLRS studied was designed for biological phosphorus removal; the average effluent phosphorus concentration for this plant was 1.45 mg/L, and alum was added in the final clarifiers. A second VLR facility was not designed for biological phosphorus removal but was required to monitor phosphorus. This plant had an average effluent phosphorus concentration of 2.19 without any chemical addition.

The VLR system is quite reliable. Table 13.7 indicates the percent of time the monthly average effluent concentration of the given pollutants was less than the concentration given in the first column. No significant difference in results was observed between winter and summer data.

13.5. EPA Evaluation of VLR

The following summarizes the major findings and conclusions of US EPA evaluation of VLRS (108). The information is based on analysis of available information from site visits, a detailed design of a full scale VLR system, and information from consultants and manufacturers.

- (a) The VLR is a modification of the conventional activated sludge process. The unique features of the process are circulating mixed liquor around a horizontal baffle with a dual aeration system, bubble diffused air beneath the horizontal baffle, and disc aerators at the surface of the aeration tank. The process operates as a plug flow reactor with capability for varying dissolved oxygen profiles to achieve biological, phosphorus and nitrogen removal. The VLR process also features a stormwater by-pass design for treatment of high peak to average flows
- (b) There are currently over ten operating VLRS in the U.S. ranging in size from 0.22 to 5.0 MGD (1.06–219 L/s)
- (c) Performance data from operating VLRS show that this process is capable of achieving effluent carbonaceous biochemical oxygen demand levels of less than 10 mg/L; effluent total suspended

Table 13.7
Reliability of the VLR treatment process (107)

Concentration, mg/L	BOD ^a	NH ₃ -N ^a	TSS ^a	P ^a
0.2	0	30	0	2
0.5	0	63	1	10
1.0	0	83	1	24
2.0	20	88	5	63
3.0	71	95	43	93
10.0	97	96	75	100
20.0	100	100	96	100
Number of plants	5	5	5	1

^aPercentage of time the monthly average concentration of the pollutant was less than the stated value in the first column.

- solids levels of less than 10 mg/L; and effluent ammonia-nitrogen levels of less than 1.0 mg/L. The process is further capable of achieving total nitrogen and phosphorus removals of 60–80%
- (d) The VLR process is applicable for flows ranging from 0.05 to over 10 MGD (2.19 to over 438 L/s)
 - (e) The claimed advantages of this process by the manufacturer include the following:
 - Higher dissolved oxygen transfer than conventional equivalent technology
 - Improved response to peak flows due to a stormwater bypass feature
 - A credit for oxygen release due to denitrification with the credit based on 80% denitrification
 - Increased mixed liquor settleability and process stability
 - (f) The design criteria for the existing VLRs are conservative. HRTs range from 11.9 to 24 h. Volumetric loading ranged from 13.6 to 23.1 lb CBOD/1,000 ft³ (218 to 370 g CBOD/d/m³). This loading is similar to that used for extended aeration systems and is about 1/3 to 1/2 of that normally used for conventional activated sludge designs
 - (g) The VLR technology has been designated as Innovative Technology by the US EPA for three plants due to a 20% claimed energy savings
 - (h) Based on this assessment, the 20% energy savings over competing technology could not be verified
 - (i) The VLR was compared to oxidation ditches as “Equivalent Technology.” The results of this comparison indicated:
 - The VLR technology produces comparable to slightly improved effluent levels of BOD, TSS and NH₃-N than oxidation ditch plants
 - Total removal of phosphorus and total nitrogen are equivalent to oxidation ditches designed for the same level of treatment
 - The energy requirements for aeration were found to be similar to 10% less than for oxidation ditches
 - The land area required for VLRs was found to be approximately 40% less than for oxidation ditches based on equivalent aeration tank loadings
 - The VLR aeration basin cost was found to be approximately 30% less than for oxidation ditches for situations where rock excavation is not required for the deeper VLR basin
 - A definitive comparison of total VLR plant costs to total oxidation plant costs could not be made. Data submitted from both manufacturers indicated a comparable cost for plants in the 0–2 MGD (0–87.6 L/s) range. The reported VLR costs at plants ranging from 2 to 10 MGD (87.6 to 438 L/s) were significantly less than oxidation ditch plant costs. This would be expected because of the modular design and common wall construction of the VLR compared to oxidation ditches
 - The total operation and maintenance costs of the two technologies were found to be similar

13.6. Energy Requirements

The VLR energy requirements are shown in Fig. 13.17. The requirements are based on the following assumptions (107):

- (a) Water Quality
 - Influent BOD₅ = 200 mg/L, Effluent = 20 mg/L
 - Influent TKN = 35 mg/L, Effluent = 1 mg/L

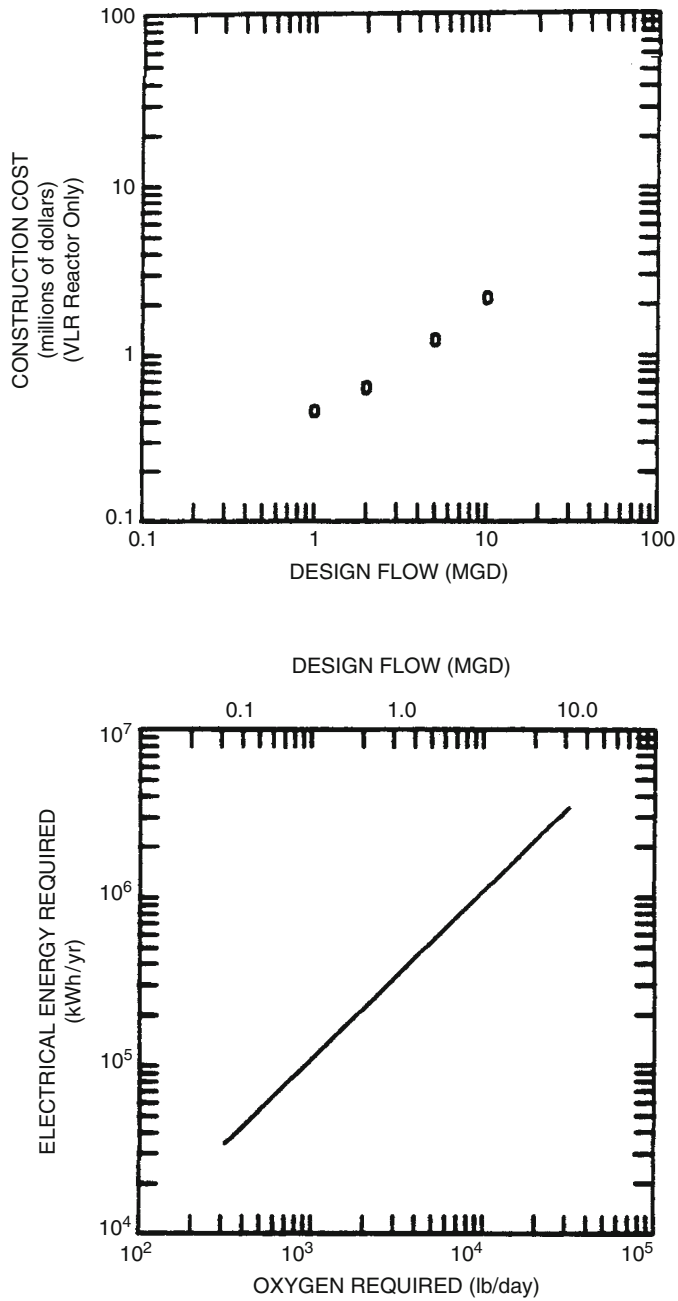


Fig. 13.17. VLR energy requirements and construction cost (107, 108). 1 MGD = 3.785 MLD = 43.8 L/s; 1 lb = 453.6 g

- (b) Design Basis
 - Oxygen transfer efficiency = 2.5 lb O₂/HP hour = 1.52 kg/kWh
 - Nitrification occurs
- (c) Operating Parameters
 - Oxygen Requirement = 1.5 lb O₂/lb BOD₅ removed, 4.57 lb O₂/lb TKN removed (1 lb/lb = 1 kg/kg)
- (d) Type of energy: Electrical

13.7. Costs

Construction costs (1991 Dollars, Utilities Index = 392.35) for VLR are shown in Fig. 13.17. To obtain the values in terms of the present 2009 U.S. Dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of $570.38/392.35 = 1.45$ (114). The operation costs are similar to oxidation ditch type treatment plant.

14. PHOSTRIP PROCESS

14.1. Description

“PhoStrip” is a combined biological–chemical precipitation process based on the use of activated sludge microorganisms to transfer phosphorus from incoming wastewater to a small concentrated substream for precipitation. As illustrated in Fig. 13.18, the activated sludge is subjected to anoxic conditions to induce phosphorus release into the substream and to provide phosphorus uptake capacity when the sludge is returned to the aeration tank. Settled wastewater is mixed with return activated sludge in the aeration tank. Under aeration, sludge microorganisms can be induced to take up dissolved phosphorus in excess of the amount required for growth. The mixed liquor then flows to the secondary clarifier where liquid effluent, now largely free of phosphorus, is separated from the sludge and discharged. A portion of the phosphorus-rich sludge is transferred from the bottom of the clarifier to a thickener-type holding tank: the phosphate stripper. The settling sludge quickly becomes anoxic and, thereupon, the organisms surrender phosphorus, which is mixed into the supernatant. The

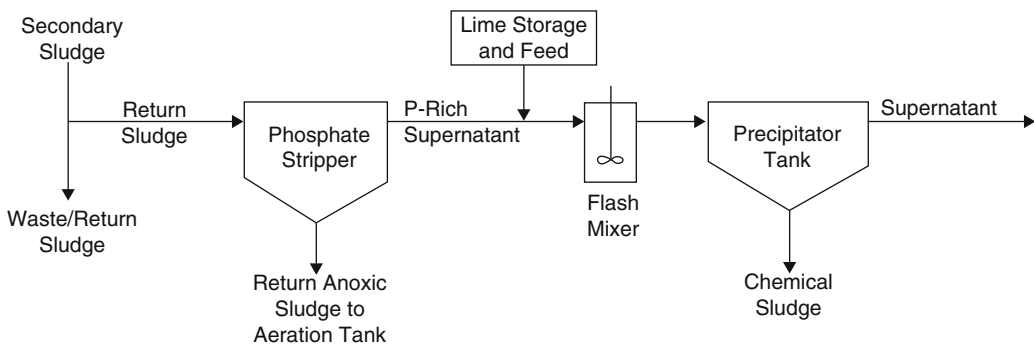


Fig. 13.18. PhoStrip process flow diagram (96).

phosphorus-rich supernatant, a low volume, high concentration substream, is removed from the stripper and treated with lime for phosphorus precipitation. The thickened sludge, now depleted in phosphorus, is returned to the aeration tank for a new cycle (96).

The PhoStrip process has demonstrated a compatibility with the conventional activated sludge process and is compatible with its modifications. The process can operate in various flow schemes, including full or Split flow of return activated sludge through the phosphate stripper, use of an elutriate to aid in the release of phosphorus from the anoxic zone of the stripper, or returning lime-treated stripper supernatant to the primary clarifier for removal of chemical sludge.

This technique is a new development in municipal wastewater treatment and has been demonstrated in pilot plant and full-scale studies. Notable large scale evaluations have been conducted at Seneca Falls, New York and, more recently, Reno/Sparks, Nevada. Nearly a dozen commercial installations are reported to be in the operational phase (31).

14.2. Applications

This method, which involves a modification of the activated sludge process, can be used in removing phosphorus from municipal wastewaters to comply with most effluent standards. Direct chemical treatment is simple and reliable, but it has the two disadvantages of significant sludge production and high operating costs. The PhoStrip system reduces the volume of the substream to be treated, thereby reducing the chemical dosage required, the amount of chemical sludge produced, and associated costs. Lime is used to remove phosphorus from the stripper supernatant at lower pH levels (8.5–9.0) than normally required. The cycling of sludge through an anoxic phase may also assist in the control of bulking by the destruction of filamentous organisms to which bulking is generally attributed (96).

On the negative side, it should be pointed out that more equipment and automation, along with a greater capital investment, are normally required than for conventional chemical addition systems. Since this method relies on activated sludge microorganisms for phosphorus removal, any biological upset that hinders uptake ability will also affect effluent concentrations. It has been found that sludge in the stripper tank is very sensitive to the presence of oxygen. Anoxic conditions must be maintained for phosphorus release to occur.

14.3. Design Criteria

The fraction of the total sludge flow that must be processed through the stripper tank is determined by the phosphorus concentration in the influent wastewater to the treatment plant and the level required in the treated effluent. Required detention time in the stripper tank ranges from 5 to 15 h. Typical phosphorus concentrations produced in the stripper are in the range of 40–70 mg/L. The volume of the phosphorus-rich supernatant stream to be lime treated is 10–20% of the total flow (96). Typical design criteria for the PhoStrip process are shown in Table 13.8 (105)

Table 13.8
Typical design criteria for the PhoStrip process (96)

Design parameter	Unit	Value
Food-to-microorganisms ratio (F/M)*	lb BOD/lb MLSS/d	0.3–0.5
Solids retention time (SRT)	d	10–30
Mixed liquor suspended solids (MLSS)	mg/L	600–5,000
Hydraulic retention time in stripper (t)	h	8–12
Hydraulic retention time in aeration tank (t)	h	4–10
Return activated sludge (RAS)	% of influent	20–50
Internal recycle (stripper underflow)	% of influent	10–20

* 1 lb/lb/d = 1 kg/kg/d

14.4. Performance

Pilot and full-scale studies of the process have shown it to be capable of reducing the total phosphorus concentration of typical municipal wastewaters to 1.5 mg/L (105), or even to 0.5 mg/L or less (106). A plant-scale evaluation of the method treating 6 MGD (262.8 L/s) of municipal wastewater at the Reno/Sparks Joint Water Pollution Control Plant in Nevada demonstrated satisfactory performance for achieving greater than 90% phosphorus removal. Results showed that the process enhanced the overall operation and performance of the activated sludge process, since it produced a more stable, better settling sludge. Regular maintenance of mechanical equipment, including pumps and mixers, is necessary to ensure proper functioning of entire system.

14.5. Cost

14.5.1. Construction Cost

Construction costs (1980 Dollars, Utilities Index = 277.60) for PhoStrip are shown in Fig. 13.19. To obtain the values in terms of the present 2009 U.S. Dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of $570.38/277.60 = 2.05$ (114). Construction costs include: stripper (10 h detention time at 50% of return sludge); flash mixer; flocculator/clarifier; thickeners; lime feed and storage facilities (96).

14.5.2. Operation and Maintenance Cost

The electrical energy required for operation of pumps, lime mixing equipment, and clarifiers, is shown in Fig. 13.11. Operation and maintenance costs (1980 Dollars, Utilities Index = 277.60) for PhoStrip are shown in Fig. 13.20. To obtain the values in terms of the present 2009 U.S. Dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of $570.38/277.60 = 2.05$ (114). Operation and maintenance costs include: labor for operation, preventive maintenance, and minor repairs; materials to include replacement parts and major repair work; lime and power costs based on the electrical energy requirement shown in Fig. 13.21 (96).

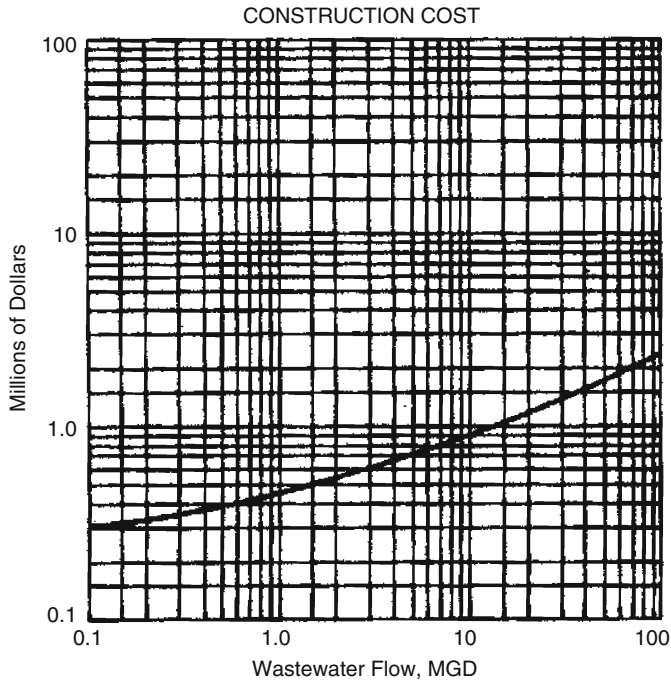


Fig. 13.19. PhoStrip construction cost (96). 1 Mgal/d = 1 MGD = 3.785 MLD = 43.8 L/s

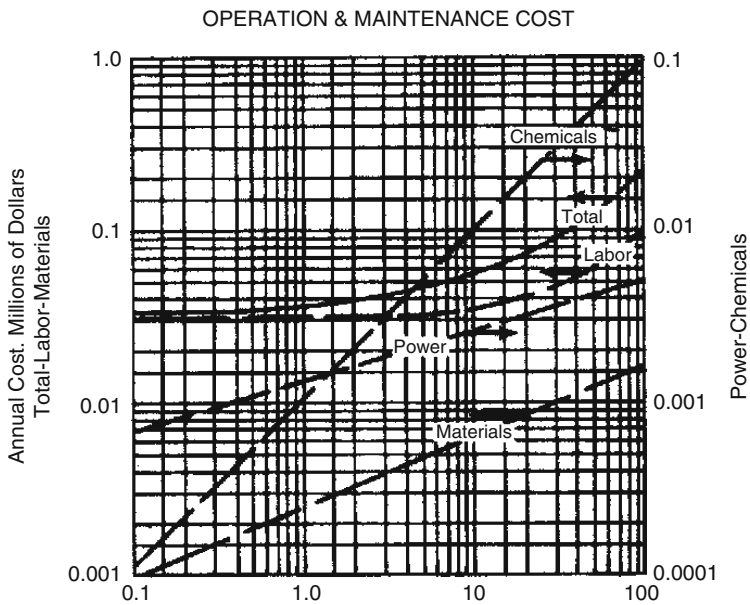


Fig. 13.20. PhoStrip operation and maintenance cost (96). 1 Mgal/d = 1 MGD = 3.785 MLD = 43.8 L/s

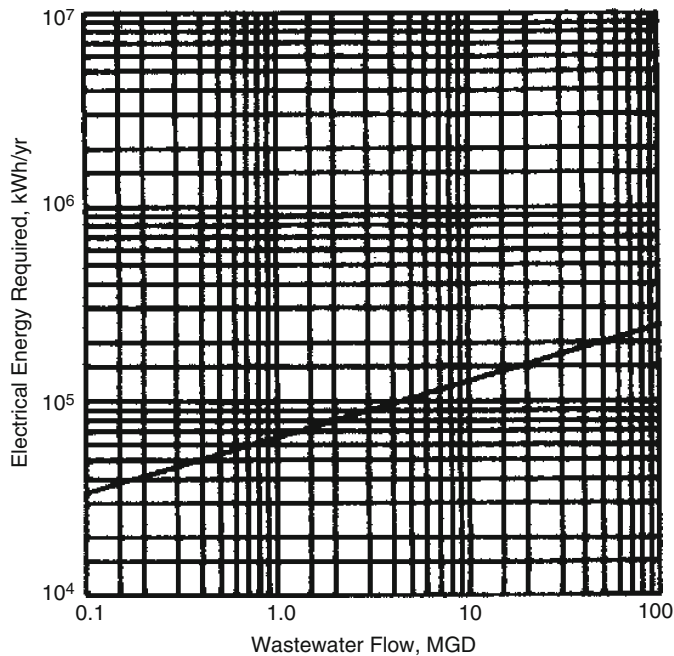


Fig. 13.21. PhoStrip electrical energy requirement (96). 1 Mgal/d = 1 MGD = 3.785 MLD = 43.8 L/s

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APPENDIX

U.S. Army Corps of Engineers Civil Works Construction Yearly Average Cost Index for Utilities (114))

Year	Index	Year	Index
1967	100	1989	383.14
1968	104.83	1990	386.75
1969	112.17	1991	392.35
1970	119.75	1992	399.07
1971	131.73	1993	410.63
1972	141.94	1994	424.91
1973	149.36	1995	439.72
1974	170.45	1996	445.58
1975	190.49	1997	454.99
1976	202.61	1998	459.40
1977	215.84	1999	460.16
1978	235.78	2000	468.05
1979	257.20	2001	472.18
1980	277.60	2002	484.41
1981	302.25	2003	495.72
1982	320.13	2004	506.13
1983	330.82	2005	516.75
1984	341.06	2006	528.12
1985	346.12	2007	539.74
1986	347.33	2008	552.16
1987	353.35	2009	570.38
1988	369.45		