

Chapter 3

MBBR and IFAS systems

H. Ødegaard

Scandinavian Environmental Technology (SET) and Norwegian University of Science and Technology (NTNU), Trondheim, Norway; SET AS, Hellemsveien 421, N 7165 Oksvoll, Norway

ABSTRACT

Even though the moving bed biofilm reactor (MBBR)- and integrated fixed film activated sludge system (IFAS)-processes are designed and used more or less as they were from the beginning (in the 1990s), there have been several advances. This chapter focuses on advances that have been made during the last 10 years in biochemical oxygen demand (BOD)- and nutrient removal, and particularly in the use of MBBR-based processes for nitrogen removal by de-ammonification, MBBRs for biological phosphate removal and MBBR-based membrane bioreactors (MBRs). Advances in MBBR biomass separation including the use of dissolved air flotation, micro-sand ballasted lamella settling (ACTIFLO®) and micro-screening (disc filtration) are presented. MBBR-based membrane bioreactors (including MBBR MBR and IFAS MBR) are discussed and a comparison is carried out in terms of volumes and footprints of a pure MBBR plant, a conventional MBR plant (based on activated sludge) and an IFAS-based MBR plant.

Keywords: Bio-P removal, de-ammonification, IFAS, MBBR, MBBR-based MBR

3.1 INTRODUCTION

The moving bed biofilm reactor (MBBR) was invented in Norway in the late 80s when focus was on nitrogen removal caused by eutrophication of the North Sea (Ødegaard *et al.*, 1994; Ødegaard, 2006). The concept was developed in close cooperation between the water treatment research group at NTNU/SINTEF in Trondheim and Kaldnes Miljøteknologi AS in Tønsberg, a company founded and based on the MBBR technology. The development was enhanced by the establishment of a research programme on nitrogen removal (Removal of nitrogen (FAN), 1988–1992 – in Norwegian) that was financed by The Norwegian State Pollution Control Authority (SFT)(Ødegaard, 1992). The research led to design criteria that are still in use (Hem *et al.*, 1994; Rusten *et al.*, 1995a,b). After a series of obstacles, the MBBR concept (reactor as well as carriers) was patent filed by Kaldnes Miljøteknologi in 1991 (with this author as the inventor) and the first plant was commissioned the same year as a small pre-denitrification plant (Lardal wastewater treatment plant [WWTP]) – the first nitrogen removal plant in Norway. Over the years the original patent was owned, and the process marketed, by a series of companies; Kaldnes Miljøteknologi AS (1991–1995), Anglian Water Ltd (1995–2002), AnoxKaldnes AB (2002–2006) and Veolia Water Systems (2006–still). Several other companies around the world are now offering MBBR systems and MBBR-based processes are used on all continents in municipal plants (including on-site) as well as industrial plants.

The MBBR is a flow-through, pure biofilm reactor, i.e. there is no recycle of biomass from the downstream separation reactor back to the MBBR. The biofilm (or attached biomass) grows on carriers that are suspended in the reactor and moving freely around with the currents set up by aeration (in aerobic reactors) and mixing (in anoxic and anaerobic reactors) – see Figure 3.1.

At steady state the net biomass growth is sloughed off the carriers due to erosion caused by the movement and mixing, and the produced biomass is separated from the water in a downstream clarification reactor. The suspended solids (SS) that are to be separated, are composed of the biomass produced as well as the incoming suspended solids that are not degraded in the reactor. And since there is no recycle, the concentration of this suspended solids is low (typically $150\text{--}250\text{ g SS} \cdot \text{m}^{-3}$), which allows for the use of any separation reactor alternative. Frequently compact separation alternatives are used, such as dissolved air flotation (DAF), lamella settling, floc blanket lamella settling (possibly micro-sand enhanced) as well as various kinds of filters (micro-screens, cloth filters, sand filters and membrane filters). Used in this way, MBBR plants may become extremely compact with a footprint for nitrogen removal of only 10–20% of that of a conventional activated sludge (AS) plant based on settling.

The pure MBBR is used for the removal of organic matter and nitrogen (see Figure 3.2) and is applied in municipal and industrial wastewater treatment as well as in drinking water treatment and treatment of recycled water in aquaculture.

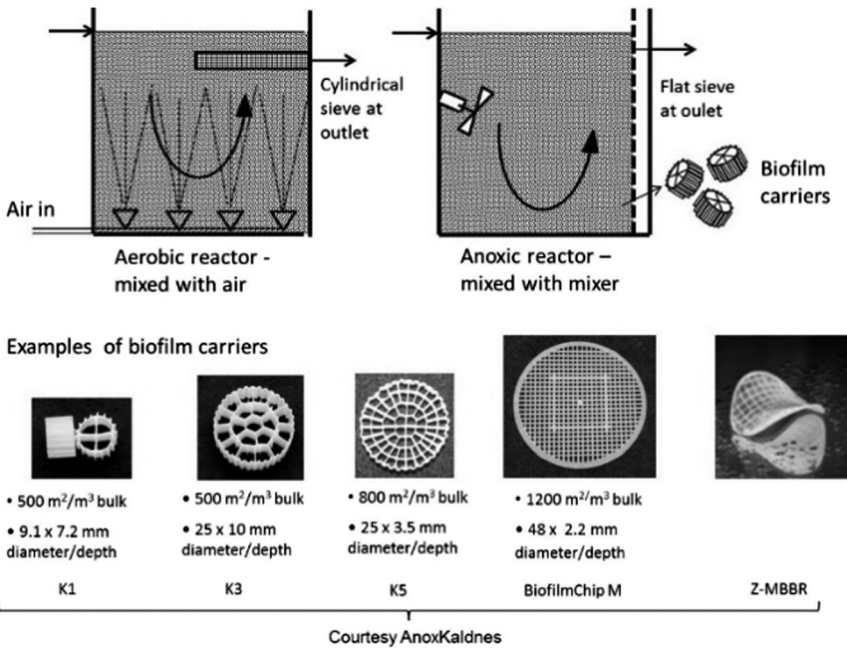


Figure 3.1 Principle of the MBBR and examples of carriers (Ødegaard, 2016).

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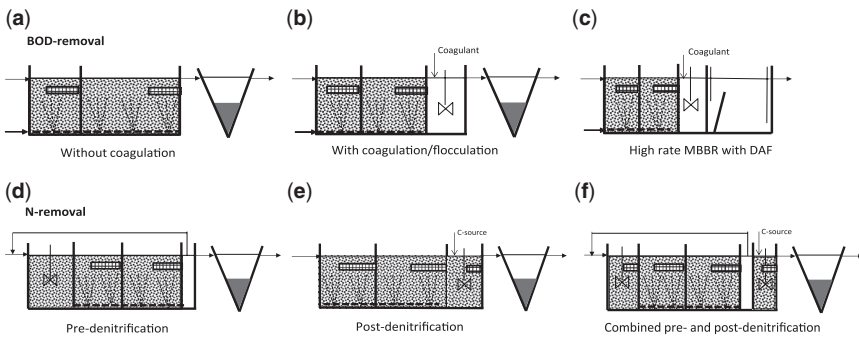


Figure 3.2 The most common principles of MBBR systems for biochemical oxygen demand (BOD)- and N-removal in pure MBBR plants.

The MBBR concept may also be used in hybrid systems, together with activated sludge, in a so-called integrated fixed film activated sludge system (IFAS for short) where MBBR carriers, with attached biomass, are present in the activated sludge tank containing suspended biomass (see Figure 3.3). In IFAS plants nitrification

can be achieved at a much lower mixed liquor suspended solids sludge retention time (SRT_{MLSS}) than in conventional activated sludge plants and hence nitrification (and N-removal) is achieved at a much lower bio-reactor volume.

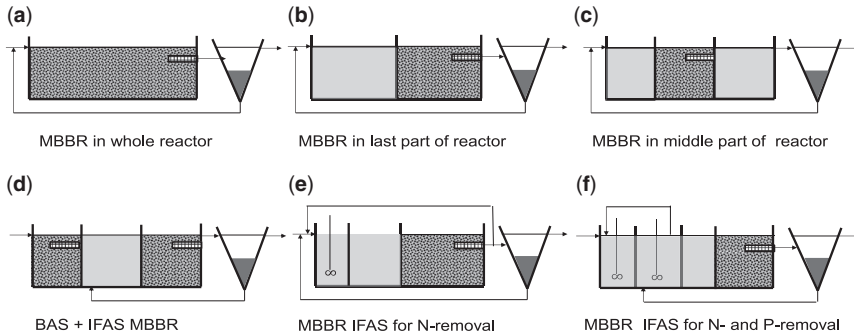


Figure 3.3 Common MBBR-based IFAS systems for different purposes (BAS – biofilm before activated sludge).

MBBR-based IFAS systems are used for a variety of reasons, including enhanced BOD removal, enhanced nitrification, enhanced N removal, enhanced biological P removal, improved settling, reduced footprint and improved operational stability (Ødegaard *et al.*, 2014).

It is fair to say that the MBBR-based processes (pure biofilm MBBR and MBBR-based IFAS) are now well established as compact, easy-to-operate wastewater treatment solutions and at this time there are least 1200 plants (about 50/50 MBBR and IFAS) of significant size around the world. The MBBR is also used in several thousand mini/on-site MBBR plants around the world.

Even though the MBBR-based processes are designed and used as they were from the beginning (in the 1990s), there have been several developments over the last 10 years. In this chapter, advances in the use of MBBR and IFAS for traditional secondary treatment (BOD- and SS-removal) and tertiary treatment (nutrient removal) will be given first, and subsequently advances in new applications of MBBR and IFAS for de-ammonification, biological P-removal and MBBR-based membrane bioreactors (MBRs) will be dealt with.

3.2 BOD-REMOVAL

The development in the area of secondary treatment (BOD- and SS-removal) has gone in the direction of increasingly more compact plants. This has been made possible partly because of the knowledge gained through research into hydrolysis of particles (Helness & Sjøvold, 2001) and partly because compact separation methods have to an increasing extent been applied. Lately the interest

in de-ammonification processes for nitrogen removal (anammox) has revived the focus on high-rate BOD-removal, to be used as the C-stage in a de-ammonification process (see below).

Biodegradable, soluble organic matter is quickly degraded. Particulate organic matter is partly hydrolysed and degraded, and partly passes more or less unchanged through the reactor. The most efficient use of the MBBR for secondary treatment only is, therefore, to design for a high soluble organic loading in the first stage of the bioreactor and a lower loading in the later stages (see Figure 3.2a) – combined with subsequent coagulation/flocculation and high-rate biomass/floc separation (see Figure 3.2b). In this way one takes advantage of the fact that the biomass will operate at a higher removal rate at a high organic loading and remove the easily biodegradable organic matter first (in the first stage) before starting to hydrolyse particulate organic matter that may take place in the later stages. Subsequent coagulation/flocculation and floc separation ensures that the biomass produced by biodegradation of soluble organic matter, as well as not biodegraded, incoming particles, are removed from the water. This combines well with phosphate removal by chemical precipitation by coagulant dosing.

Design of pure MBBR systems is based on surface area loading rates, r_A ($\text{g} \cdot \text{m}^{-2}$ carrier surface \cdot d) and the volume of the reactor is determined as:

$$V_{\text{MBBR}} = L_s \cdot (r_A \cdot A_{\text{spec}} \cdot f)^{-1} \cdot 100$$

V_{MBBR} = volume of MBBR (m^3)

L_s = substrate load ($\text{g} \cdot \text{d}^{-1}$)

r_A = design surface area loading rate ($\text{g} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$)

A_{spec} = specific area of carrier in bulk ($\text{m}^2 \cdot \text{m}^{-3}$)

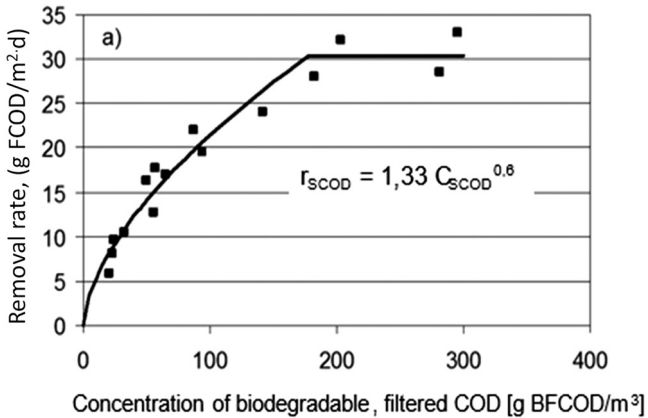
f = filling fraction of media (%)

Typical design values for BOD-removal are $8 \text{ g BOD}_5 \cdot \text{m}^{-2} \cdot \text{d}$ at 10°C for the whole reactor (divided in at least two steps) without any post-coagulation/flocculation and $11.5 \text{ g BOD}_5 \cdot \text{m}^{-2} \cdot \text{d}$ at 10°C with post-coagulation/flocculation. A typical design temperature coefficient is $\theta = 1.08$ (Norsk Vann, 2009).

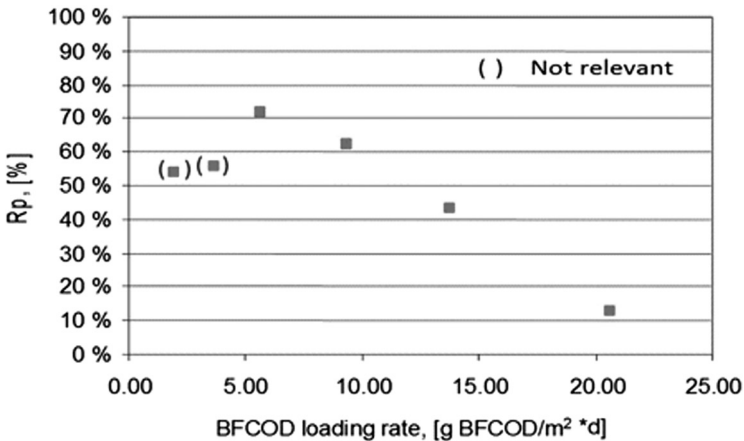
Because of the compactness, the hydraulic residence time (HRT) in moving bed reactors for carbonaceous matter removal will be quite low (15–60 minutes), depending on the strength of the wastewater and the effluent requirement. It is recommended always to have a minimum of two reactor stages. Since the first stage should have a higher organic loading than the second stage, a division of the total volume of around 35% in the first reactor step and around 65% in the second is found to be optimal, based on 0.6 order kinetics in completely mixed reactors (see below).

A more sophisticated design procedure for the removal of organic matter is based on knowledge about the organic matter characteristics of the incoming water, especially with respect to particulate (suspended) organic matter versus soluble (filterable, assumed to be biodegradable) organic matter. The sizing of the

reactor is made dependent on the degradation rate of the soluble BOD (SBOD₅) which is assumed to be 0 order at bulk SBOD₅ concentrations above 125 g · m⁻¹ and 0.6 order at SBOD₅ concentrations below 125 g · m⁻¹ (transformed to SBOD₅ from biodegradable, filterable COD (BFCOD) based on Figure 3.4). The SBOD₅ to be designed for, is composed of the incoming SBOD₅ and the SBOD₅ that is the result of hydrolysis of the incoming particulate BOD (PBOD₅). The latter is dependent on the soluble organic matter load (see Figure 3.5) and the percentage of the incoming PBOD₅ being hydrolysed, given by the formula in Table 3.1.



AQ1 Figure 3.4 The removal rate of filtered (1 μm filtered) COD versus the concentration of filtered biodegradable COD (Helness & Ødegaard, 2005).
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AQ8 Figure 3.5 Extent of hydrolysis (R_p, %) of particulate COD being hydrolysed to soluble COD versus BFCOD load (Helness & Sjøvold, 2001).

Table 3.1 Design values for BOD-removal in MBBRs based on soluble BOD (SBOD₅) degradation.

Effluent Concentration, c (g SBOD ₅ · m ⁻³)	Reaction Order	Reaction Rate (15°C) g SBOD ₅ · m ⁻² · d ⁻¹	Biodegradation of PBOD ₅ (%)
>125	0	$r_0 = k_0 = 20$	0
<125	0.6	$r_{0.6} = 1.15 \cdot [c]^{0.6}$	40–0.125 F^2

$F = \text{SBOD}_5 \text{ load (g SBOD}_5 \cdot \text{m}^{-2} \cdot \text{d}^{-1})$

The necessary area (and hence volume based on carrier characteristics and filling fraction) is determined reactor stage by reactor stage towards the effluent SBOD₅ targeted (for instance SBOD₅ = 5 g · m⁻³ if the BOD_{5,total} target is 10 g · m⁻³).

3.2.1 High-rate MBBR for BOD-removal

As the first stage for upgrading of existing (AS) plants (for instance in order to achieve nitrification, or as the carbon removal stage of a de-ammonification process), the so-called high-rate MBBR may be used (see Figure 3.2c) (Melin *et al.*, 2004; Helness *et al.*, 2005; Brosseau *et al.*, 2016).

The idea behind the high-rate process is that the attached biomass only degrades the soluble, most easily biodegradable organic matter, while the particulate, more slowly biodegradable organic matter passes through the reactor, more or less unchanged, and is separated from the water by coagulation/floc separation. This requires an organic loading on the MBBR that is low enough for truly soluble organic matter to be biodegraded and high enough to prevent significant hydrolysis and degradation of suspended and colloidal organic matter. In designing such a high-rate process, therefore, it is especially important to know the distribution of soluble and particulate organic matter.

In order to avoid hydrolysis of particulate organic matter, the biodegradable, filtered (i.e. soluble) organic load has to be higher than around 20–25 gBFCOD · m⁻² _{carrier-area} · d⁻¹ (equivalent to about 15–25 g SBOD₅ · m⁻² · d⁻¹ in a normal wastewater), as demonstrated in Figure 3.4 (Helness & Sjøvold, 2001). The reaction rate at this high load will be at its maximum (0 order) of around 30 gBFCOD · m⁻² · d⁻¹, as demonstrated in Figure 3.4, resulting in an MBBR effluent concentration of BFCOD >150–200 g · m⁻³ equivalent to 100–150 gSBOD₅ · m⁻³. Below this, the reaction order will be close to half order, but even at 100 gBFCOD · m⁻³ the removal rate will be as high as 20 gBFCOD · m⁻² · d⁻¹ corresponding to around 60 gCOD_{total} · m⁻² · d⁻¹. At this high load only a negligible fraction of the particulate organic matter will be converted to soluble organic matter (Helness & Sjøvold, 2001) – see Figure 3.5.

Particulate matter entering the MBBR as well as biomass produced by biodegradation of easily degradable soluble organic matter in the MBBR, are

separated from the water by coagulation/flocculation/biomass separation. By the use of a compact biomass separation method, such as dissolved air flotation or micro-sand ballasted lamella settling, an extremely low plant footprint will be obtained.

The separated sludge will have a high biogas potential since it consists of non-biodegraded incoming particulate organic matter and biomass resulting from a high-rate biological degradation of easily biodegradable, soluble organic matter. A combination of cationic polymer plus a small dose of iron is used in order to minimize the precipitation of metal hydroxide that does not contribute to biogas production.

A pilot plant study of the high-rate MBBR process (with DAF for biomass separation) demonstrated that secondary treatment standard ($25 \text{ gBOD}_5 \cdot \text{m}^{-3}$ in effluent) could be reached at a total HRT from inlet to outlet of around 1.0 h (Melin *et al.*, 2004; Helness *et al.*, 2005). The pilot plant was operated on a quite concentrated municipal wastewater at relatively low temperatures ($10\text{--}15^\circ\text{C}$). Based on this study, the following design criteria were proposed for the high-rate MBBR process:

- *Load:* $20\text{--}25 \text{ gCOD}_{\text{filt}} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ ($15\text{--}20 \text{ gBOD}_{\text{filt}} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$), $65\text{--}85 \text{ gCOD}_{\text{tot}} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ ($45\text{--}60 \text{ gBOD}_{\text{tot}} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$)
- *Sludge production:* $0.5 \text{ gDS} \cdot \text{gCOD}_{\text{rem}}^{-1}$ in MBBR step, $1.0 \text{ gDS} \cdot \text{gSS}_{\text{rem}}^{-1}$ in biomass separation step
- *Coagulant dosage:* $5 \text{ g cationic polymer} \cdot \text{kgSS}^{-1} + 35 \text{ gFe} \cdot \text{kgSS}^{-1}$

3.3 N-REMOVAL BY NITRIFICATION/DENITRIFICATION

The MBBR was developed during a time of nitrogen removal focus and several publications address this ambition (Ødegaard *et al.*, 1994; Hem *et al.*, 1994; Rusten *et al.*, 1995a,b; Ødegaard, 2014; 2016; 2017). The design criteria developed in the early 90s are, however, still valid.

3.3.1 Nitrification

After BOD-removal, nitrification design of pure MBBRs is normally based on the following model (Hem *et al.*, 1994; Rusten *et al.*, 1995a):

$$r_N = k \cdot (S_n)^n$$

$$r_N = \text{nitrification rate (gNH}_4\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1})$$

$$k = \text{reaction rate coefficient (estimated at } 0.5\text{--}0.75 \text{ – depending on pretreatment)}$$

$$n = \text{reaction order constant, estimated at } n = 0.7$$

$$S_n = \text{rate-determining ammonium concentration, gNH}_4\text{-N} \cdot \text{m}^{-3} \text{ (} S_n \sim (\text{DO}_{\text{bulk}} - 0.5)/3.2)$$

k will be dependent upon the C/N-ratio ($\text{gBOD}_5/\text{gNH}_4\text{-N}$) of the incoming water to the bioreactor and vary from 0.7 at C/N = 0.5 to 0.5 at C/N = 4.5. If the preceding

MBBR reactors are designed to remove essentially all BOD, the nitrification rate coefficient for the final nitrification reactors may be set at 0.75.

The nitrification rate is linearly dependent on the oxygen concentration up to more than $10 \text{ gO}_2 \cdot \text{m}^{-3}$ (Hem *et al.*, 1994; Æsøy *et al.*, 1998), see Figure 3.6. The dissolved oxygen (DO) strongly limits the nitrification rate even at concentrations as low as $2\text{--}3 \text{ gO}_2 \cdot \text{m}^{-3}$. Normally one is designing for DO concentrations of $4\text{--}6 \text{ gO}_2 \cdot \text{m}^{-3}$ at peak load.

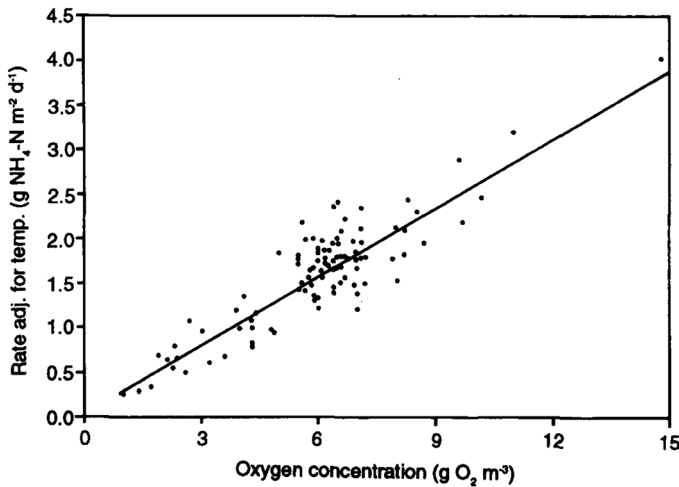


Figure 3.6 Nitrification rate (adjusted to 15°C vs. DO $\text{DO}/\text{NH}_4\text{-N}$ ratio $<2 \text{ gO}_2 \cdot \text{gNH}_4\text{-N}^{-1}$ (Hem *et al.*, 1994).

An advantage of the linear relationship between oxygen concentration and nitrification rate is that it may favourably be used for process control. If extremely low values of ammonium are required, the last reactor segment must be designed for an ammonium limited rate – in which case oxygen concentration may be low since it is no longer governing the rate of nitrification.

Oxygen transfer is enhanced by the presence of carriers as demonstrated in Figure 3.7 (Christensson, 2011). The higher the filling fraction (up to around 60%), the better the oxygen transfer. The benefit of the carriers is dependent on the design of the carrier used. It seems as if tube-like carriers are better than coin-like carriers in this respect. Carrier suppliers should, therefore, provide Standard Oxygen Transfer Rate (SOTR) data. This implies that good oxygen transfer is achieved even with a robust, medium coarse bubble aeration system, that is often used in order to achieve even air distribution and complete mixing without clogging at low head-loss (Ødegaard *et al.*, 1994).

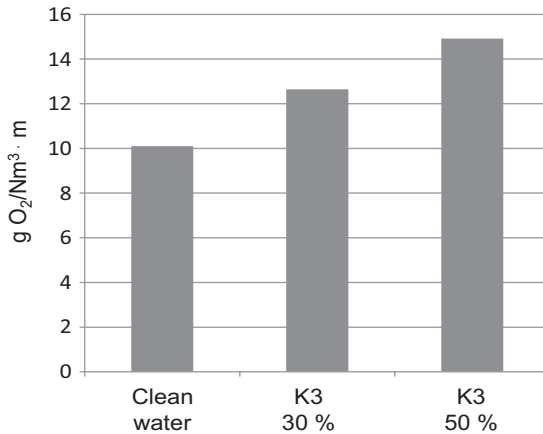


Figure 3.7 Influence of carrier filling fraction on oxygen transfer (Christensson, 2011).

3.3.2 Denitrification

Pure MBBRs are used for pre-denitrification, post-denitrification and combined pre- and post-denitrification (see Figure 3.2d–f). Too low C/N-ratio in the raw water is a common challenge when stringent effluent nitrogen concentrations are to be met in traditional pre-denitrification plants. In such situations the so-called combined pre- and post-denitrification MBBR process (see Figure 3.8) is often used. This process configuration gives a superior possibility to control towards the effluent standard aimed for – however stringent that is (Rusten & Ødegaard, 2007; Ødegaard, 2014).

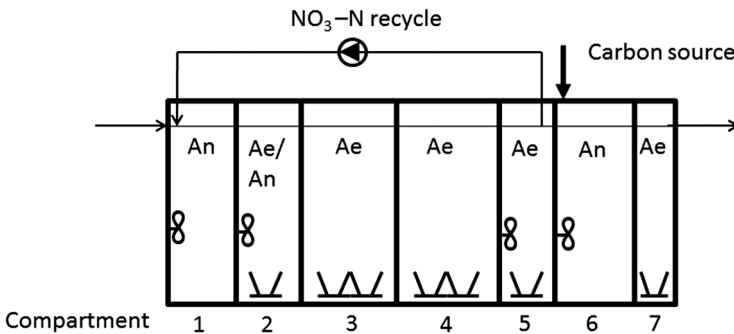


Figure 3.8 The typical build-up of a combined pre- and post-denitrification MBBR (Ødegaard, 2016).

3.3.2.1 Combined pre- and post-denitrification MBBR

The combined-denitrification (DN) MBBR plants normally have a division between aerobic and anoxic compartments of the MBBR – as indicated in Figure 3.8.

The first compartment (comp. 1) is anoxic (pre-denitrification) and receives recycled nitrate from the last aerobic compartment. The second compartment (comp. 2) may be operated both anoxic (no air) and aerobic (with air). Then follow two aerobic reactors (comp. 3 and 4), used for residual organic matter removal and nitrification. Compartment 5 also contributes to nitrification and may be operated with or without aeration. Compartment 6, to which carbon is added, is anoxic and for post denitrification. Finally there is optionally a small, aerobic compartment for oxygenation as well as for removal of any residual biodegradable organic matter, possibly caused by excess addition of external carbon source.

This bioreactor scheme may give very high nitrogen removal efficiencies and optimal flexibility of operation at minimized consumption of external carbon source (Rusten & Ødegaard, 2007). During summer operation, for instance, less nitrification volume is needed because of higher temperatures and the raw water normally has a higher DO concentration than in the winter time. In this situation compartment 2 may be used for denitrification (stirred and not aerated). More nitrate may be returned to the pre-DN tank and the need for added carbon source in the post-DN tank is lower. In summer time, therefore, most of the nitrogen can be removed by pre-denitrification, reducing the consumption of external carbon source to just that which is needed to ensure that the nitrogen removal goal is achieved. During winter operation, more nitrification capacity is needed and compartment 2 is operated with aeration. More carbon source will have to be added to the post-DN tank in order to achieve the treatment goal. The reason for the option of not aerating compartment 5, is de-oxygenation in order to minimize the return of oxygen to the pre-denitrification reactor. Because the ammonia concentration here will be low ($<2 \text{ g NH}_4\text{-N} \cdot \text{m}^{-3}$), the oxygen is no longer rate limiting for the nitrification rate, but rather the ammonia concentration is. Hence it does not matter that the DO is reduced which minimizes oxygen respiration in the pre-DN compartment caused by the recirculation of oxygen.

The extent of nitrification may be controlled by the oxygen level in the nitrification reactor (comp. 4) since there is a linear relationship between nitrification rate and oxygen concentration (see Figure 3.6). The extent of denitrification in the post denitrification step is governed by the availability of carbon source. Nitrate measurement in the post denitrification outlet may therefore control carbon source addition and hence the final denitrification result.

Figure 3.9 shows post-denitrification rates in a combined pre- and post-denitrification MBBR plant (Gardermoen WWTP in Norway) where three different external carbon sources were tested (Rusten *et al.*, 1996). The tests were carried out in a situation where the nitrate concentration was not rate limiting. Use of methanol and monopropylene glycol (MPG) gave similar denitrification rates. Use of ethanol

doubled the denitrification rates. All three external carbon sources showed a temperature coefficient of $\theta = 1.07$. Necessary C/N-ratios were also temperature dependent, which can be explained by higher specific sludge production (lower decay) at lower temperatures. The necessary C/N-ratio increased with an average of 45% when the temperature was reduced from 15 to 5°C.

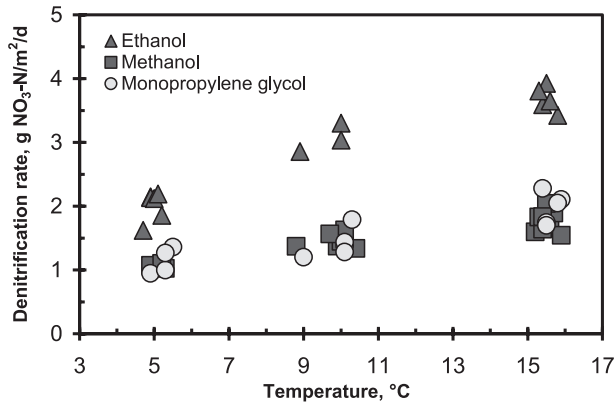


Figure 3.9 Denitrification rate versus temperature with various external carbon sources (Rusten *et al.*, 1996).

For 95% removal of nitrate, ethanol had the lowest carbon demand, with a necessary C/N-ratio of $3.9 \text{ gCOD}_{\text{added}} \cdot \text{gNO}_3\text{-N}_{\text{equivalents}}^{-1}$ at 10°C. MPG had the highest necessary C/N-ratio at $5.6 \text{ gCOD}_{\text{added}} \cdot \text{gNO}_3\text{-N}_{\text{equivalents}}^{-1}$ at 10°C (Rusten *et al.*, 1996).

3.3.2.2 Post-denitrification MBBR

Pure MBBR post-denitrification is also frequently used downstream of an existing nitrifying activated sludge plant (Aspegren *et al.*, 1998; Stinson *et al.*, 2009; Ericsson, 2011).

The post denitrification rate is totally governed by the availability of biodegradable organic matter as long as the effluent $\text{NO}_3\text{-N}$ concentration is $>5 \text{ gNO}_3\text{-N} \cdot \text{m}^{-3}$. The $K_{s,\text{COD}}$ -value is quite low, around $3 \text{ gSCOD}_{\text{biodegradable}} \cdot \text{m}^{-3}$, so as long as the $\text{SCOD}_{\text{biodegradable}}$ -concentration is above $\sim 10 \text{ g biodegradable, soluble COD (BSCOD)} \cdot \text{m}^{-3}$, the rate is not limited by organic matter (Rusten *et al.*, 1995b). If very low effluent nitrate concentrations are to be met ($<3\text{--}5 \text{ gNO}_3\text{-N} \cdot \text{m}^{-3}$), the nitrate concentration will limit the denitrification rate. Reported $K_{s,\text{NO}_3\text{-N}}$ -values are in the range of $0.5\text{--}3.0 \text{ gNO}_3\text{-N} \cdot \text{m}^{-3}$ (Christensson, 2008; Rusten, 2008; Stinson *et al.*, 2009). One may conservatively design for a $K_{s,\text{NO}_3\text{-N}}$ value of $1.5 \text{ mgNO}_3\text{-N} \cdot \text{m}^{-3}$ and a maximum design rate of $3 \text{ gNO}_x\text{-N} \cdot \text{m}^{-2}\text{d}^{-1}$ at 15°C at no substrate limitation

(BSCOD $>10 \text{ g} \cdot \text{m}^{-3}$ and $\text{NO}_3\text{-N} >5 \text{ g} \cdot \text{m}^{-3}$. At a moderate nitrate limitation (effluent $3\text{--}5 \text{ gNO}_3\text{-N} \cdot \text{m}^{-3}$), the rate is still quite high ($>2 \text{ gNO}_x\text{-N} \cdot \text{m}^{-2}\text{d}^{-1}$ at 15°C) while at strong limitation ($\text{NO}_3\text{-N}$ concentrations $<3 \text{ gNO}_3\text{-N} \cdot \text{m}^{-3}$), the design rates should probably not be set above $1 \text{ gNO}_3\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ at 15°C . Ericsson (2011) reported maximum full-scale denitrification capacities at Rya WWTP in Sweden, of $2.2 \text{ gNO}_x\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ at around 10°C , with methanol as carbon source.

This has, above all, an impact on how the post denitrification reactor should be divided in segments. The following thumb rule is proposed:

- (1) If the effluent design $\text{NO}_3\text{-N}$ value is $>3 \text{ mg} \cdot \text{l}^{-1}$, use only one post-DN reactor.
- (2) If the effluent design $\text{NO}_3\text{-N}$ value is $<3 \text{ mg} \cdot \text{l}^{-1}$, divide the DN-step in two reactors, designed for maximum rate in the 1st reactor and half of the maximum rate in the 2nd reactor. If only one reactor is possible, design it for half of the maximum design rate.
- (3) If the effluent design $\text{NO}_3\text{-N}$ value is $<1 \text{ mg} \cdot \text{l}^{-1}$, one should consider three reactors in the post-DN step, with the 3rd designed for a quarter of the maximum rate.

3.3.3 N-removal in MBBR-based IFAS plants

The most frequently used objective for using MBBR-based IFAS is to enhance nitrification, with the goal of being able to fully nitrify at a lower aerobic SRT_{MLSS} than in a conventional AS system. This results in a reduced footprint for nitrifying-, N removal- and biological nutrient removal (BNR) plants. Positive side-effects are improved biomass settling and operational stability. IFAS systems are therefore, frequently used for upgrading of conventional AS (CAS) plants.

One of the benefits of the MBBR IFAS system is that one needs only add the amount of attached growth area (i.e. the amount of carriers) that is necessary. The maximum recommended filling fraction in a pure MBBR system is 67% (Ødegaard *et al.*, 1994). In MBBR IFAS systems filling fractions $>60\%$ are seldom used, and one may adapt the biofilm growth area necessary to the volume available (Ødegaard *et al.*, 2014).

As shown in Figure 3.3a–c one may add carriers to the entire reactor volume or to one or more segments of the reactor. Figure 3.10a shows a common pre-denitrification application where carriers are used in the aerobic part of the reactor train. Carriers may also be used, however, in anoxic reactors (as shown in Figure 3.10b), although this is currently less common.

Nitrification in MBBR-based IFAS systems is mainly influenced by the aerobic SRT_{MLSS} , the organic carrier area load (relative to ammonium load), the reactor DO concentration, the ammonium concentration and the temperature. In order to demonstrate this, data from pilot plants at Tau WWTP, Norway and from Broomfield WWTP, Colorado, USA are shown in Figure 3.11 (Rusten *et al.*, 2003).

The Tau pilot plant, with a filling fraction of only 18% ($64 \text{ m}^2 \cdot \text{m}^{-3}$ carrier surface), achieved complete nitrification at aerobic SRT_{MLSS} as low as 2.2 d at 15°C , about 40% lower aerobic SRT_{MLSS} than without carriers (Figure 3.11a). The carrier biomass nitrification rate in the Broomfield plant (see Figure 3.11b) agreed well with the nitrification design model (see above) for pure MBBR systems, indicating that this model may also be used in MBBR IFAS systems. This pilot demonstrated that a DO concentration of $4 \text{ g} \cdot \text{m}^{-3}$ was needed to achieve an effluent of $1 \text{ gNH}_4 \cdot \text{m}^{-3}$ (Figure 3.11b) indicating the important role of the attached biomass in nitrification.

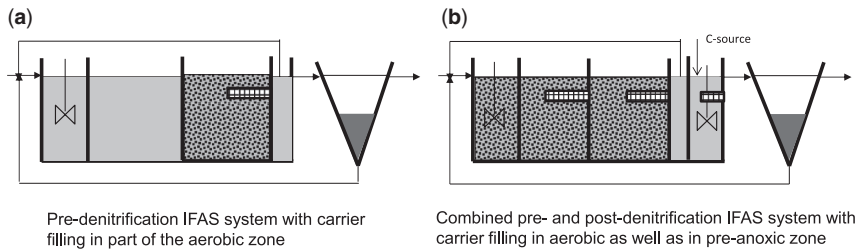


Figure 3.10 Examples of IFAS systems for nitrification/denitrification.

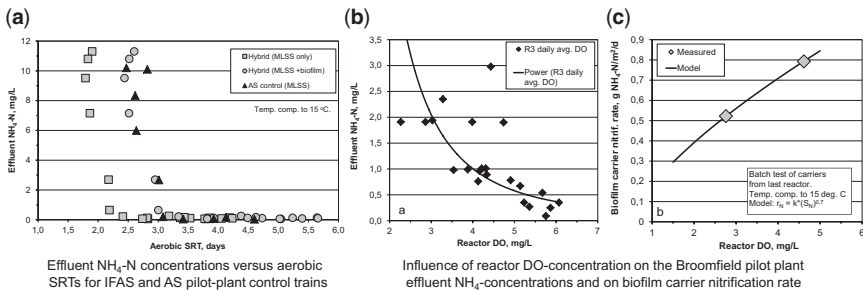


Figure 3.11 Influence of aerobic SRT_{MLSS} and DO in MBBR-based IFAS pilot plants (Rusten *et al.*, 2003).

Figure 3.12 shows the relationship between design aerobic SRT_{MLSS} and temperature for various US-plants designed for nitrification (Johnson, 2009) together with the curves for nitrifier growth rate and the Abwassertechnische Vereinigung (ATV) design curve for nitrification in AS plants. The SRTs for these full-scale MBBR IFAS systems are generally far lower than the German design curve.

Different models have been used to explain the enhancement of nitrification in IFAS systems: (a) the high attached biomass SRT , (b) an increased nitrifier

biomass per unit volume, (c) an increased nitrifier activity in the attached biomass compared to the suspended biomass, and (d) the seeding of nitrifiers into the suspended biomass from the attached biomass (Ødegaard *et al.*, 2014). McQuarrie (2009) studied the nitrification activity in the suspended biomass and the attached biofilm in an MBBR IFAS system based on data from Thomas (2009). They found that the specific nitrification activity ($\text{mgNH}_4\text{-N} \cdot \text{gSS}^{-1} \cdot \text{h}^{-1}$) was much higher in the attached biomass than in the suspended biomass (see Figure 3.13a).

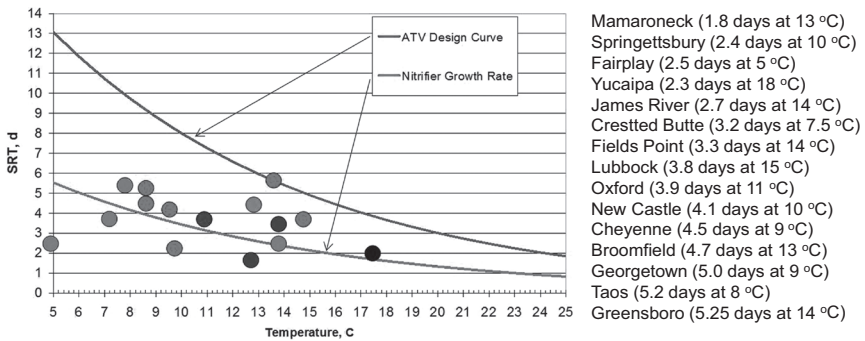


Figure 3.12 Design SRT versus temperature for full-scale IFAS systems (Johnson, 2009).

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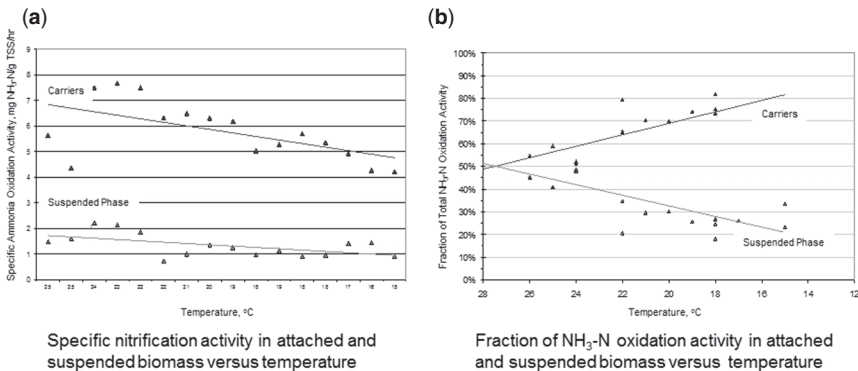


Figure 3.13 Distribution of nitrification activity in attached and suspended biomass as a function of temperature (McQuarrie, 2009; Thomas, 2009).

A probable explanation for this, is that the nitrifiers in the attached biomass are always exposed to low soluble biodegradable carbon to ammonia ratio, resulting in an attached biofilm enriched with nitrifying bacteria. McQuarrie (2009) also demonstrated that the fraction of $\text{NH}_4\text{-N}$ oxidation activity taking place on the

carriers, increased with decreasing temperature, while the fraction in the suspended biomass decreased with decreasing temperature. This demonstrates that the benefit of the carriers in the overall process is greater at low temperatures when it is most needed.

Denitrification (DN) in IFAS systems may be by pre-DN, post-DN and combined pre- and post-DN (Ødegaard, 2014). Pre-DN-systems, in which the anoxic reactor does not contain carriers, are most common, but there also is an interesting potential for using carriers in anoxic reactors (Choi *et al.*, 2007).

In the previously mentioned Tau pilot plant, Rusten *et al.*, (2003) found a >50% higher DN rate in the IFAS pre-anoxic reactor biomass than in the CAS pre-anoxic reactor biomass ($3.15 \text{ g NO}_x\text{-N} \cdot \text{kg MLSS}^{-1} \cdot \text{h}^{-1}$ and $2.00 \text{ g NO}_x\text{-N} \cdot \text{kg MLSS}^{-1} \cdot \text{h}^{-1}$ respectively). So did Onnis-Hyden *et al.* (2011). This indicates that the low aerobic SRT of hybrid processes leads to higher DN-rates than in CAS plants. In systems with higher MLSS and SRT, much of the carbon is lost during the aerobic phase; hence, the shorter the aerobic SRT, the more carbon is available for DN.

3.4 N-REMOVAL BY DE-AMMONIFICATION IN MBBR-BASED PLANTS

Traditionally, biological nitrogen removal in wastewater is carried out through nitrification/denitrification. The challenges with this process are the high energy demand for the aerobic (autotrophic) part of the process (nitrification), and the carbon source requirement for the anoxic (heterotrophic) part of the process (denitrification). Over the last 10 years research focus has been on an alternative nitrogen removal process (de-ammonification) as a consequence of the desire to overcome the challenges of nitrification/denitrification.

De-ammonification is a biological treatment process that converts ammonia to nitrogen gas without the need for a carbon source and with much less air needed than in nitrification (Water Environment Research Foundation [WERF], 2014). De-ammonification is accomplished through two biological process steps. The first is termed nitrification, which is the aerobic oxidation of ammonia-N ($\text{NH}_4\text{-N}$) to nitrite-nitrogen ($\text{NO}_2\text{-N}$) by autotrophic, aerobic, ammonia-oxidizing bacteria (AeAOB). Nitrification is well known in wastewater treatment, as it is the initial step in biological nitrification of ammonia-N to nitrate-N. Since stoichiometrically only about half of the ammonia needs to be converted to nitrite, this is also known as partial nitrification. In the second step of de-ammonification, which is the anammox (anaerobic ammonia oxidation) reaction, $\text{NH}_4\text{-N}$ is the electron donor and oxidized under anaerobic conditions by autotrophic, anaerobic ammonia-oxidizing bacteria (AnAOB) that can use $\text{NO}_2\text{-N}$ as the electron acceptor. Hence the overall process is most correctly referred to as the partial nitrification/anammox process, but in this chapter the expression de-ammonification is used (for short).

There are several obvious advantages of the de-ammonification process; the energy for air is about 60% lower, no carbon source is needed, the alkalinity

consumption is about 50% lower, and the sludge production is in the order of 70% lower (if organic particles are removed upstream). About 89% of the inorganic nitrogen ($\text{NH}_4\text{-N} + \text{NO}_2\text{-N}$) fed to the process ends up as N_2 gas and about 11% as $\text{NO}_3\text{-N}$ (WERF, 2014). So if very low tot N-concentrations in the effluent ($< \sim 3 \text{ gN} \cdot \text{m}^{-3}$) are aimed for, a small nitrification/post-denitrification step has to be applied downstream of the de-ammonification step.

Since it is implemented most easily when the temperature is high (preferably $> \sim 25^\circ\text{C}$), the ammonium concentration is high ($> 500 \text{ mg NH}_4\text{-N/l}$) and the C/N-ratio is low, the process has been successfully implemented for sludge reject water (in side-stream), that normally represents about 25% of the nitrogen load on a typical BNR plant. It is not yet established as a proven process for nitrogen removal in the main-stream, but will represent a break-through in wastewater treatment when it is.

3.4.1 De-ammonification in the side-stream

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The MBBR has demonstrated itself to be a robust and compact biofilm technology for de-ammonification in sludge reject water (side-stream) (Jardin & Hennerkes, 2012; Rosenwinkel & Cornelius, 2012; Christensson *et al.*, 2013; Malovanyy, 2017). The first full scale MBBR demonstration plant for treatment of reject water was established in Hattingen, Germany (Rosenwinkel & Cornelius, 2012) as a multi-stage MBBR process. The demonstration project here proved the MBBR de-ammonification process to be cost-effective in comparison to conventional N-removal (Jardin & Hennerkes, 2012).

In a biofilm reactor it is possible carry out both nitrification and de-nitrification by anammox in the same biofilm. Partial nitrification to nitrite and autotrophic N-removal (i.e. anammox) may occur simultaneously within the biofilm, where aerobic and anoxic zones results from oxygen mass transfer limitation under limited dissolved oxygen (DO) conditions (see Figure 3.14).

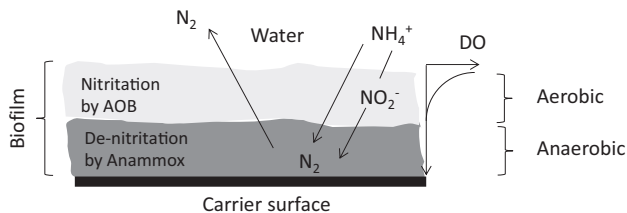


Figure 3.14 Schematic of 1-stage nitrification/anammox biological processes occurring inside a carrier’s biofilm.

In an MBBR the ammonium oxidizing bacteria (AeAOB) as well as the anammox (AnAOB) bacteria are maintained in the attached biofilm on the suspended carriers

retained in the reactor by the sieves, with no risk of biomass wash-out (Cema, 2009; Christensson *et al.*, 2013). AeAOB oxidize NH_4 to NO_2 in the aerobic zone of the biofilm (i.e. outer part) while AnAOB bacteria located in the anoxic zone of the biofilm (i.e. inner part) consume NO_2 produced by AeAOB together with the excess $\text{NH}_4\text{-N}$ (see Figure 3.14). The main challenge of single-stage de-ammonification is to prevent further oxidation of nitrite to nitrate by nitrite oxidizing bacteria (NOB). One strategy is to use intermittent aeration in order to prevent nitrite oxidation by lowering the DO intermittently.

ANITATMMox is a single-stage MBBR de-ammonification process (Christensson *et al.*, 2013; Veuillet *et al.*, 2013; 2014). In the ANITATMMOX system, a DO control system is used to prevent nitrite oxidation in the aerobic zone of the biofilm while maximizing the amount of nitrite available for the anammox bacteria. The DO set-point is automatically adjusted based on online inlet and outlet concentrations of NH_4 and NO_3 to control the NO_3 production below 11% of $\text{NH}_4\text{-N}$ removed (i.e. stoichiometric NO_3 production by anammox) while at the same time keeping high NH_4 -oxidation performance in the reactor. This real-time DO control strategy reduces the need for a mechanical mixer in the MBBR due to the continuous aeration pattern (Christensson *et al.*, 2013).

To shorten the start-up phase, new ANITATMMox plants are seeded with a small fraction of colonized carriers, which reduce the time required for the development of a mature de-ammonification biofilm on the brand new carriers. Seeding has proven to dramatically decrease the start-up time from up to a year down to 2–3 months depending of the amount of seeding (Christensson *et al.*, 2013). A full-scale study indicated, however, that inoculum was not necessary in order to start up a full-scale de-ammonification process in the reject water from thermophilic digestion if nitrification biomass is already established. The start-up took 120 days in full-scale and 72 days in lab-scale (Kanders *et al.*, 2016).

When comparing 2-stage MBBR and 1-stage MBBR for de-ammonification on sludge reject water, Cema *et al.* (2010) found excellent nitrification in the 2-stage system, but also that it was more difficult to control the anammox stage. Nitrite is a substrate for anammox bacteria but nitrite accumulation can also inhibit their activity (Fernandez *et al.*, 2012; Lotti *et al.*, 2012). Inhibition was experienced in a 2-stage pilot-plant, while the nitrites produced by AOB were consumed immediately by anammox in the 1-stage process (Plaza, 2015). In order to prevent anammox inhibition in a 2-stage process the $\text{NO}_2\text{-N}/\text{NH}_4\text{-N}$ ratio in the reactor effluent (influent to anammox reactor) should be kept as close as possible to the stoichiometric optimum of 1.32 (0.57 $\text{NO}_2\text{-N}/0.43 \text{NH}_4\text{-N}$).

In an attempt to overcome the inhibition challenge and to improve the ANITATMMox process performances under different operating conditions, it was hypothesized that substrate transport could be enhanced by combining suspended cultures and fixed biomass into one system, i.e. an IFAS system (see Figure 3.15) (Veuillet *et al.*, 2013; 2014).

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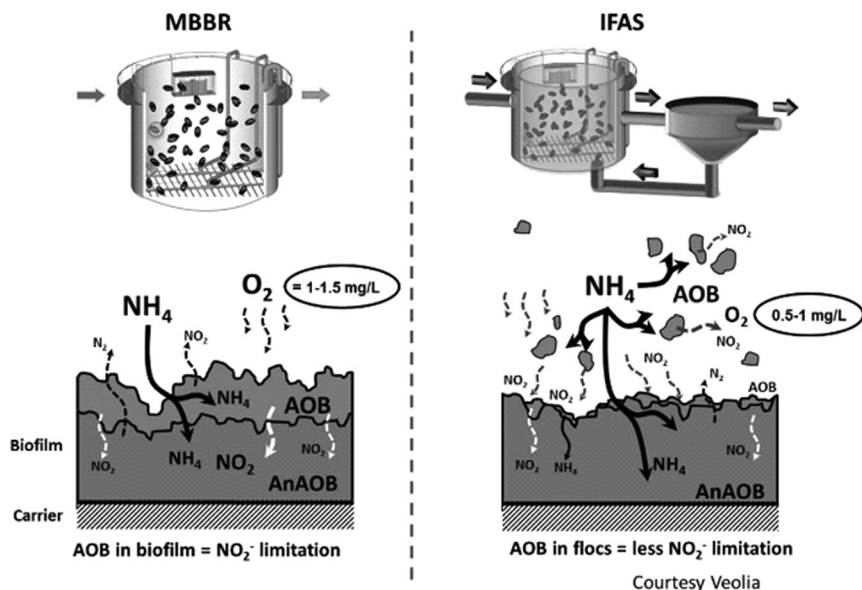


Figure 3.15 Pure MBBR versus MBBR-based IFAS for de-ammonification (Veuillet *et al.*, 2013).

In lab-scale tests on anaerobic sludge digester reject water, it was demonstrated that the nitrogen removal capacity was almost four times higher when comparing the pure 1-stage MBBR version with the IFAS 1-stage version (Christensson *et al.*, 2013; Veuillet *et al.*, 2013). With reference to Figure 3.15 the authors offered the following explanation:

“In an IFAS system the NO₂ level in the bulk liquid may be high enough to be able to fully diffuse into the deep layer of the biofilm where the AnAOB are mostly located, but not too high in order to avoid inhibition of AnAOB. Increased AOB activity in the IFAS configuration is obtained by the increased growth of AOB in the MLSS. In the IFAS mode the sludge retention time (SRT) can be increased from less than a day to several days, thereby preventing wash-out of AOBs. Substrate diffusion limitation in flocs is less apparent than in biofilms that are thicker and denser. Hence this should lead to better substrate accessibility for AOB (oxygen, NH₄) in the suspended solids in the IFAS configuration, meaning that AOB that pre-existed on the outer-layer of the biofilm gradually disappear from the biofilm due to a lack of oxygen (i.e. DO in bulk liquid is lower in IFAS than MBBR) that is now mostly consumed by AOB in the liquid phase. The biofilm is therefore almost exclusively composed of AnAOB with a very fine top layer of oxygen scavengers like AOB or normal heterotrophs as indicated in Figure 3. The larger AOB population in IFAS configuration improves the overall flux of NO₂ produced for AnAOB but also the residual concentration in the bulk improving the diffusion of the NO₂ through the basal layer of the biofilm

where AnAOB are located and therefore increasing their active fraction in the biofilm” (Veuillet *et al.*, 2014).

In the experiments, the global amount of AnAOB in the IFAS and the MBBR configuration was found to be similar. The improvement of nitrites flux and production in the IFAS mode led, however, to a N-removal rate near the maximum that was obtained during fully anoxic AnAOB activity batch tests, with non-limiting substrate levels (i.e. $\text{NH}_4\text{-N}$ and $\text{NO}_2\text{-N}$) (Veuillet *et al.*, 2013).

Full-scale experiences with the ANITA™Mox process, both in pure biofilm as well as in IFAS mode, from the demonstration plant at Sjölanda WWTP in Sweden, confirm the findings in the lab- and pilot-scale tests (Veuillet *et al.*, 2014).

3.4.2 De-ammonification in the main-stream

While a number of wastewater treatment plants have implemented the side-stream de-ammonification process as a cost-effective, efficient and reliable option to treat reject water, the techniques required to sustain de-ammonification in the colder and more dilute main-stream wastewater are yet to be developed. The challenges of mastering main-stream de-ammonification have been summarized as follows (Trela *et al.*, 2014):

- (1) The dominance of NOB growth at lower temperatures makes the selection of AOB over NOB challenging. Moreover, NOB suppression by free ammonia inhibition is not possible because of low ammonium concentration. If NOB out-competes AOB, nitrate accumulation results, which will significantly decrease nitrogen removal efficiency. This seems to be the biggest challenge for reaching high efficiency of nitrogen removal treating main-stream wastewater.
- (2) The possible solution for out-competing NOB is to optimize operation parameters (DO, intermittent aeration phase duration, pH, inorganic carbon concentration) for stimulating AOB growth and suppression of NOB growth. Using the IFAS system instead of pure MBBR may also be a possible way to overcome NOB out-competition – as discussed above.
- (3) An effective retention of the anammox biomass in a reactor is required. This is because inflow nitrogen concentration in main-stream wastewater is low ($25\text{--}50 \text{ gNH}_4\text{-N} \cdot \text{m}^{-3}$) and this, together with low yield and growth rate of anammox bacteria, leads to low anammox biomass production. This is the main reason why biofilm systems (i.e. MBBR) are especially suitable.
- (4) Nitrogen transformation rates are about 70–80% lower for main-stream wastewater with a yearly average temperature of 15°C , compared to supernatant treatment at 30°C based on activation energies of de-ammonification reactions. Moreover, because of lower ammonium and nitrite concentrations in the process, even lower rates can be expected.

At this time most developers and researchers work with MBBR-based IFAS systems for main-stream de-ammonification (Lemaire *et al.*, 2015; Malovanny *et al.*, 2015; Malovanny, 2017). IFAS-based ANITA™Mox is proposed for use according to two different strategies (Lemaire *et al.*, 2015), see Figure 3.16.

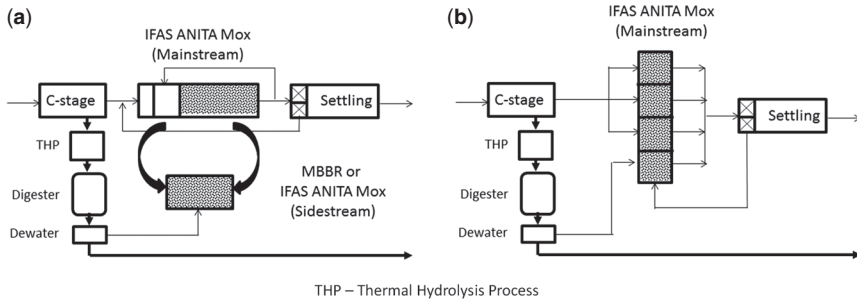


Figure 3.16 Schematic of main-stream ANITA™Mox WWTP (a) with carrier recycling concept between side-stream and main-stream ANITA™Mox reactors and (b) with alternating feed concept between side-stream and main-stream water (Ødegaard, 2016 after Lemaire *et al.*, 2015).

The main aim is to obtain bio-augmentation by moving carriers from the more robust side-stream process to the less robust main-stream process. Anammox-rich biofilm carriers can easily be transferred back and forth from the side-stream to the main-stream, for instance by the use of an air-lift pump (Figure 3.16a). This action will allow for partial regeneration of the anammox biofilm due to the more favourable conditions in the side-stream reactor (higher temperature, higher NH_4 level, lower COD level). The carrier transfer flow rate (Figure 3.16a) will have to be controlled in order to maximize the bio-augmentation effect to the main-stream IFAS ANITA™Mox reactor (Lemaire *et al.*, 2015).

In the alternative approach (Figure 3.16b) the feeding is alternated between the main-stream COD-treated effluent and the side-stream reject water to a multi-celled IFAS ANITA™Mox system. As shown in Figure 3.16b, one of the cells will receive the side-stream reject water for a period of time providing optimal conditions for anammox and is, therefore, temporarily considered as a side-stream system. The side-stream reject water is then switched to feed another IFAS cell while the main-stream COD-treated effluent is switched back to the cell which has previously been fed with the side-stream reject water. The period of time during which a given IFAS cell receives a side-stream effluent will have to be controlled.

Results from two main-stream IFAS ANITA™Mox studies were presented – one pilot-scale study in Paris treating chemically enhanced primary effluent and one full-scale prototype (reactor volume 50 m^3) study at ANITA™Mox plant in Malmö, Sweden treating effluent from a high-rate activated sludge step in the full-scale

Sjölunda plant (Lemaire *et al.*, 2015). In the prototype plant the bio-augmentation strategy using alternating feed between side-stream and main-stream effluent was tested. From these studies it was concluded that the IFAS ANITA™Mox process was successfully applied as a main-stream process with removal rates obtained in the pilot-plant of $1.4 \text{ gN} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ (on average) during the summer (at 23°C) and $0.5\text{--}0.8 \text{ gN} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ during the start of winter (at 17°C). The bio-augmentation strategy in the prototype plant seemed to be efficient in suppressing the NOB activity while achieving N-removal rates up to $0.1\text{--}0.3 \text{ kgN} \cdot \text{m}^{-3} \cdot \text{d}^{-1}$ (equal to $0.23\text{--}0.68 \text{ gN} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$) at $13\text{--}18^\circ\text{C}$ (Lemaire *et al.*, 2015).

Malovanyy (2017), who studied both pure MBBR and IFAS single-stage systems for main-stream de-ammonification, found an advantage of IFAS systems. He presented the mass balance for the IFAS single stage system shown in Figure 3.17 that demonstrates that heterotrophic denitrification played a significant role. The heterotrophic denitrification activity presupposes availability of organic matter that is in contradiction to the enhancement of anammox bacteria. This demonstrates the challenge in design and operation of one-stage systems.

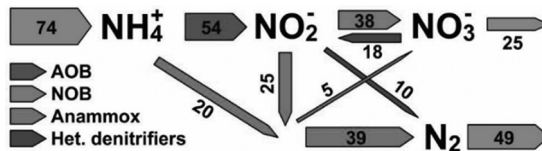


Figure 3.17 Nitrogen conversion routes in a single stage IFAS de-ammonification experiment (Malovanyy, 2017).

The development of the MBBR system over the years has demonstrated that better control and operability has been achieved when dividing the bacterial cultures in separate stages – making it possible to better optimize process conditions for each stage. The single-stage pure MBBR seems to be inferior to the single-stage MBBR-based IFAS system for de-ammonification. However, a two-stage pure MBBR-process, where nitrification and anammox is separated in two different stages, may be competitive in a main-stream process if NOB control can be secured. This may require extended BOD-removal in the C-stage, biofilm thickness control and bio-augmentation from the side- to the main-stream.

A two-stage system with one or more continuously aerated MBBRs for nitrification and an anoxic MBBR for anammox, was tested out in lab-scale by Piculell *et al.* (2016). It was hypothesized that the wash-out of NOB from the biofilm would be enhanced by having a very thin biofilm. The biofilm thickness was maintained below $200 \mu\text{m}$ by the use of a carrier especially designed for de-ammonification (AnoxKaldnes Z-200 carriers – see Figure 3.1), hence exposing a large fraction of the biomass close to bulk liquid free ammonia (FA) and/or free nitrous acid (FNA)

concentrations. By doing so, it was expected that NOB establishment in the deeper biofilm layers would be prevented, and the chances of washing out NOBs from the biofilm would hence increase. The thin biofilm would also ensure high oxygen availability, which was expected to improve nitrification rates. The feed was switched periodically from low-strength, low-temperature main-stream wastewater to reject water at high temperatures and concentrations (in agreement with the strategy shown in Figure 3.16b). This sudden exposure to high substrate concentrations and temperatures was expected to inhibit NOB growth in the thin biofilm, and possibly also to boost AOB activity (Piculell *et al.*, 2016).

The lab-study demonstrated indeed that stable nitritation and anammox was achievable and the results demonstrated that periodic exposure of the biomass to high concentration reject water in order to favour AOB activity and suppress NOB growth was successful (Piculell *et al.*, 2016). In the nitritation stage, ammonia removal rates ranged between 0.25 and 0.50 $\text{kgNH}_4\text{-N} \cdot \text{m}^{-3} \cdot \text{d}^{-1}$ (equal to 1.25–2.5 $\text{gNH}_4\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$), with a nitrite accumulation ratio above 75%, and the effluent was fed to the second stage anammox reactor, with removal rates reaching 0.20 $\text{kgNH}_4\text{-N} \cdot \text{m}^{-3} \cdot \text{d}^{-1}$ (equal to 1.0 $\text{gNH}_4\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$). At an average influent concentration of $51.6 \pm 5.2 \text{ gNH}_4\text{-N} \cdot \text{m}^{-3}$ effluent concentrations below $15 \text{ gTotN} \cdot \text{m}^{-3}$ could be achieved. Further studies are needed in order to optimize the scheme and evaluate the fate of the NOB, but once fully developed, this concept is expected to enable new solutions for main-stream anammox applications (Piculell *et al.*, 2016).

In a proposed flow diagram for an MBBR-based, energy neutral, advanced wastewater treatment plant for the future, Ødegaard (2016) proposed a scheme for de-ammonification as shown in Figure 3.18.

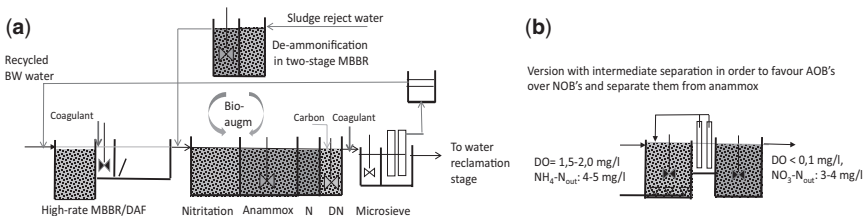


Figure 3.18 Proposed scheme for de-ammonification in an MBBR-based energy-neutral process solution for advanced wastewater treatment (based on Ødegaard, 2016).

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The strategy is based on:

- Good removal (>85%) of BOD up-front of the de-ammonification reactor by a high-rate MBBR process, in order to suppress heterotrophs during de-ammonification

- De-ammonification in the main stream in two-stage MBBR (with thin biofilm carrier) in order to be able to optimize the two parallel processes:
 - Nitritation at $DO = 1.5\text{--}2.0 \text{ gO}_2 \cdot \text{m}^{-3}$ and $\text{NH}_4\text{-N}_{\text{out}} = 4\text{--}5 \text{ g} \cdot \text{m}^{-3}$ (in order to maximize AeAOB and suppress NOB)
 - Anammox at $DO < 0.1 \text{ gO}_2 \cdot \text{m}^{-3}$ and $\text{NO}_3\text{-N}_{\text{out}} = 3\text{--}4 \text{ g} \cdot \text{m}^{-3}$ (in order to make AnAOB overrule NOB for NO_2)
- Use of bio-augmentation, as presented above by moving carriers back and forth between the de-ammonification MBBRs in side-stream and main-stream by the use of an air-lift pump (or similar) as well as returning the side-stream water to the main-stream de-ammonification MBBR
- Use of a two-step nitrification/denitrification MBBR directly after the main-stream de-ammonification MBBR in order to get down to very low Tot N-concentrations in the effluent ($< 3 \text{ g TotN} \cdot \text{m}^{-3}$)

Post denitrification (with carbon source addition) would have to be used in the post nitrification/denitrification step since there is very little $\text{SCOD}_{\text{biodegradable}}$ left in the recycled side-stream after anaerobic digestion and dewatering, and very little $\text{SCOD}_{\text{biodegradable}}$ left in the main-stream after the nitrification step.

Since it has been shown that the AOB activity is higher in a suspended biomass, it might be beneficial to insert a compact separation step (for instance a micro-screen, see Figure 3.18b) in between the aerobic and the anaerobic step. This may enhance AOB activity and suppress NOB activity in the aerobic reactor.

3.5 P-REMOVAL

In MBBR plants P-removal is normally carried out through chemical precipitation. Also in IFAS plants chemical P-removal may be implemented (normally through simultaneous precipitation), but in IFAS plants biological P-removal may also quite easily be implemented in the same way as in conventional activated sludge (CAS) plants. Chemical P-removal in MBBR- and IFAS plants is quite established and below we shall focus on advances in biological P-removal in pure MBBR plants and only touch briefly upon chemical P-removal.

3.5.1 Chemical P-removal in MBBR and IFAS plants

In MBBR plants P-removal is normally carried out through chemical precipitation, usually by adding Al^{3+} or Fe^{3+} to the effluent directly after the MBBR and before flocculation and floc/biomass separation reactors, see Figure 3.19.

Since the SS-concentration after the MBBR is relatively low ($150\text{--}250 \text{ gSS} \cdot \text{m}^{-3}$) there is no need for a separate biomass separation step between the MBBR and the chemical precipitation/flocculation step. Flocculation is normally carried out in a multi-compartment flocculator with residence time (HRT) of 10–30 min depending on the upstream process. The floc/biomass separation step may be based on any particle separation method (see below), but normally it is based on

settling or dissolved air flotation (DAF). If the effluent P standard is very stringent ($<0.3 \text{ mg Tot P} \cdot \text{l}^{-1}$) it is common to use a polishing step based on micro-straining (i.e. disc filtration), sand filtration or membrane filtration.

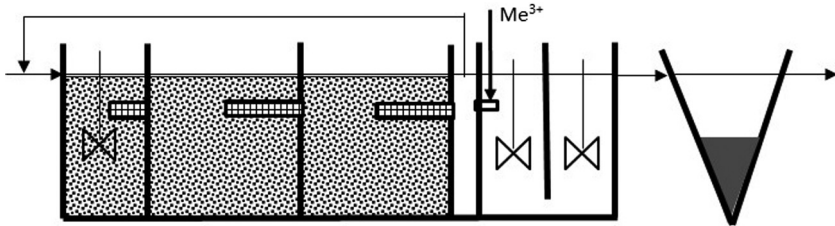


Figure 3.19 Typical principle for chemical P-removal in a pre-denitrification (as an example) MBBR plant.

In IFAS plants, chemical P-removal is commonly carried out through simultaneous precipitation by adding the precipitant directly to (or after) the IFAS-reactor, normally without any flocculation reactor in between.

3.5.2 Biological P-removal in MBBR plants

When conditions are favourable (sufficiently high concentration of easily biodegradable COD in incoming water), biological P-removal may quite easily be implemented in IFAS plants in the same way as in conventional activated sludge (CAS) plants. A University of Cape Town (UCT)-configuration is preferable. At this time it is not known whether or not any plants have been built using carriers in the anaerobic reactor.

3.5.2.1 Discontinuously operated (SBR) MBBR

Through experiments in small pilot scale, Helness demonstrated that biological phosphorus and nitrogen removal could be achieved in an MBBR operated as an SBR (Helness, 2007; Helness & Ødegaard, 2001; 2005). It was found that the SBR cycle should be tuned to achieve near complete removal of easily biodegradable soluble COD in the anaerobic phase and complete nitrification in the aerobic phase.

Pilot experiments with municipal wastewater (influent concentrations after pretreatment: $4.2 \text{ gPO}_4\text{-P} \cdot \text{m}^{-3}$, $18 \text{ gNH}_4\text{-N} \cdot \text{m}^{-3}$, $104 \text{ gSCOD} \cdot \text{m}^{-3}$ and $205 \text{ gCOD} \cdot \text{m}^{-3}$) were carried out. It was demonstrated that at an SBR cycle of 6 hours (with 1 hour and 40 minutes anaerobic phase followed by a 4 hour and 20 minutes aerobic phase), less than $10 \text{ gN}_{\text{soluble}} \cdot \text{m}^{-3}$ and $0.3 \text{ gPO}_4\text{-P} \cdot \text{m}^{-3}$ in the SBR effluent could be achieved. However, with this wastewater (low in biodegradable COD), removals as good as this could only be achieved, when dosing with an external carbon source (acetate). A ratio of $15 \text{ gBSCOD/gPO}_4\text{-P}$ was required in order to

achieve the effluent concentrations reported above. For an SBR MBBR process like this, the design criteria in Table 3.2 were recommended for a wastewater with COD:N:P ~ 100:10:2 (Helness, 2007).

Table 3.2 Proposed design values for an SBR MBBR for a wastewater with COD:N:P ~ 100:10:2 (Helness, 2007).

Parameter	Design Value
Total COD-loading rate, (gCOD · m ⁻² · d ⁻¹)	<5
Anaerobic BSCOD ¹ -loading rate, (gBSCOD · m ⁻² · d ⁻¹)	<5
Aerobic ammonia loading rate, (gNH ₄ -N · m ⁻² · d ⁻¹)	<0.4
Required influent BSCOD/PO ₄ -P	20

¹BSCOD – biodegradable soluble COD.

3.5.2.2 Continuously operated MBBR

Saltnes *et al.* (2017) presented an alternative, continuously operated, pure MBBR process for nutrient removal, based on bio-P and simultaneous nitrification/denitrification (SND). In this MBBR process the biomass experiences succeeding anaerobic and aerobic conditions by physically moving the carriers from the aerobic to the anaerobic zone by a conveyer belt – as demonstrated in Figure 3.20.

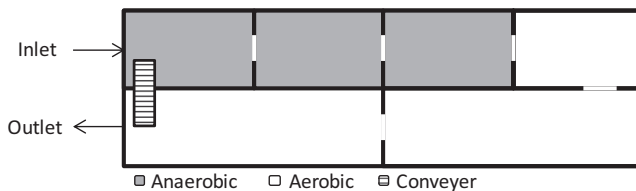


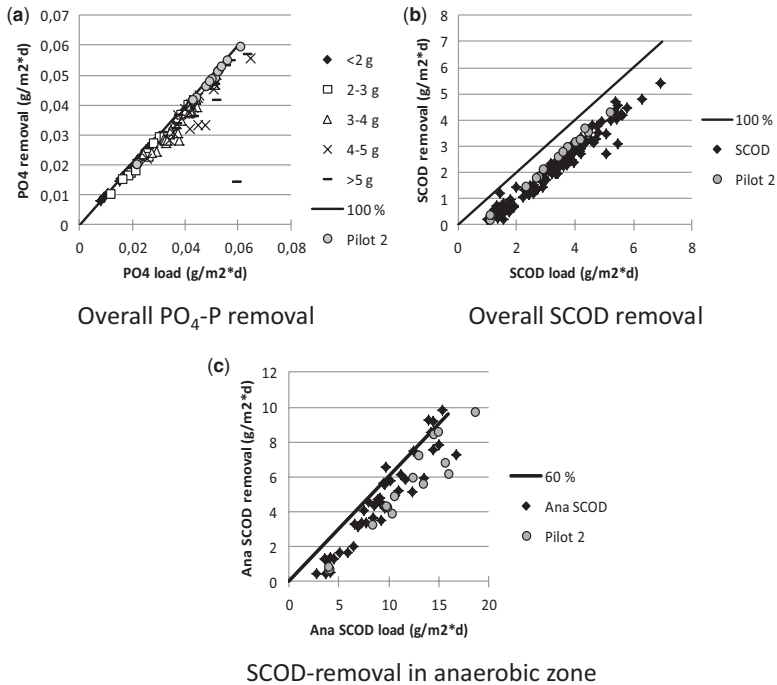
Figure 3.20 Flow scheme of the pilot used in the study of a continuous MBBR-plant for bio-P and SND-N removal (Saltnes *et al.*, 2017).

The MBBR is partitioned in several anaerobic and aerobic zones – all containing carriers. There are no sieves between the zones and the carriers are flowing, together with the water, through openings in the partition walls between the zones, first through the mechanically mixed, anaerobic zones and thereafter through the aerated zones. From the last aerobic zone the carriers (with biofilm on them) are transported mechanically by a conveyer sieve, that also separate the carriers from the wastewater, back to the first anaerobic zone, while the treated wastewater flows out of the bioreactor from the last aerobic zone.

In the experiments carried out at HIAS WWPT in Norway, the MBBR of the pilot plant had a total reactor volume of 7 m³ and was operated with an effective carrier

area of $300 \text{ m}^2 \cdot \text{m}^{-3}$ reactor volume. The anaerobic volume was varied between 30 and 40% of the total volume, and the total hydraulic residence time (HRT) was varied between 5 and 10 hours. The pilot plant was operated in two main experimental periods (referred to as pilot 1 and pilot 2). In pilot 2 the pilot plant design was improved, within the same total volume of the pilot plant, with optimized flow passages between zones and division of the aerobic volume into several volumes in order to enhance SND. The DO in the aerated zones was kept between 4 and $8 \text{ gO}_2 \cdot \text{m}^{-3}$.

The inlet wastewater to HIAS WWTP is influenced by effluent from the food processing industry in the area and has a relatively high ammonia- as well as biodegradable COD-concentration ($61 \text{ gNH}_4\text{-N} \cdot \text{m}^{-3}$, $312 \text{ gSCOD} \cdot \text{m}^{-3}$, $3.6 \text{ gPO}_4\text{-P} \cdot \text{m}^{-3}$). The load and removal of $\text{PO}_4\text{-P}$ (in $\text{gPO}_4\text{-P} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$) over several months of operation in pilot 1 and for 3 weeks of operation in pilot 2, after stable enhanced biological phosphorus removal (EBPR) activity had been established in each pilot, are shown in Figure 3.21a. The results for pilot 1 are divided into groups according to the SCOD load ($\text{gSCOD} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$). The removal of $\text{PO}_4\text{-P}$ seemed to be negatively influenced by high organic loads ($>5 \text{ gSCOD} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$) in pilot 1, while in pilot 2 the removal was $>95\%$ even at an SCOD load of $5.2 \text{ g} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ (Saltnes *et al.*, 2017).



AQ4 Figure 3.21 Removals of $\text{PO}_4\text{-P}$ and SCOD in the pilot plant (Saltnes *et al.*, 2017).

Based on the experiments it was concluded that a total P concentration in the effluent of $0.4 \text{ gP} \cdot \text{m}^{-3}$ (the full-scale plant standard) could be reached in the full-scale plant provided that a particulate P concentration of less than $0.2 \text{ gP} \cdot \text{m}^{-3}$ could be maintained after biomass separation.

A steady EBPR process is dependent on phosphate accumulating organisms (PAO) to take up volatile fatty acids (VFA; produced by hydrolysis of easily biodegradable organic matter, SBCOD in the anaerobic zone), for which they use stored polyphosphate resulting in P release. The VFA taken up and stored as polyhydroxyalkanoates (PHA) are utilized for growth and P uptake in the aerobic zone and the new generation of PAO also take up P in excess resulting in P removal. If VFA or SBCOD leak to the aerobic zone, one might experience a reduction in P uptake due to reduced PAO growth. A good removal of SCOD in the anaerobic phase is important, therefore, for PAO both under anaerobic and aerobic conditions.

Figure 3.21b shows the overall SCOD load versus removal and Figure 3.21c shows the SCOD load versus removal in the anaerobic phase for some months of operation in pilot 1 and for 3 weeks of operation in pilot 2. Because of the inert SCOD-fraction (difference between the points and the 100% line), the total SCOD removal is only about 80%, but the biodegradable SCOD-removal is $>90\%$ at a removal rate $>5 \text{ g} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$. About 60% of the SCOD was removed in the anaerobic zone, with an anaerobic SCOD load of up to $18 \text{ g} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ (Saltnes *et al.*, 2017).

During the pilot and laboratory testing of the new bio-P process (pilot 2), a significant nitrogen removal potential was discovered. Both nitrification and denitrification took place in the aerobic zone. This indicates that simultaneous nitrification/denitrification (SND) took place in different layers of the biofilm as proposed by Helness and Ødegaard (2001). The process enhances SND because most of the BOD is removed before the aerobic stage thereby improving nitrification. The BOD uptake in the anaerobic zone by denitrifying PAO secures an efficient use of wastewater carbon by anoxic uptake of $\text{PO}_4\text{-P}$ and denitrification of $\text{NO}_3\text{-N}$ using the same carbon. However, when PAO participate in denitrification, their P removal per COD taken up decreases. As opposed to a standard activated sludge EBPR-plant, $\text{NO}_3\text{-N}$ from the aerobic zone is not returned to the anaerobic/anoxic zone in this process which could otherwise lead to inefficient use of carbon (Saltnes *et al.*, 2017).

Simultaneous nitrification/denitrification (SND) was studied only in the pilot 2 experimental phase and experiments are still going on. The average ammonia load, the nitrification rate and the denitrification rate in pilot 2 was found to be $0.86 \text{ gN} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, $0.54 \text{ gN} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ and $0.33 \text{ gN} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$ respectively (Saltnes *et al.*, 2017).

3.6 ORGANIC MICRO-POLLUTANT REMOVAL

Biofilm processes, and in particular MBBR-based processes, have been found to be more efficient for the removal of organic micro-pollutants (OMP) than activated sludge processes (Falås *et al.*, 2012; 2013; 2016; Torresi *et al.*, 2017).

Through batch experiments Falås *et al.* (2012) assessed, the removal of seven active pharmaceutical substances (ibuprofen, ketoprofen, naproxen, diclofenac, clofibrac acid, mefenamic acid, and gemfibrozil) by the biomass on MBBR carriers and in suspended sludge from several full-scale wastewater treatment plants (including activated sludge, MBBR-based IFAS and MBBR plants). A distinct difference between nitrifying activated sludge biomass and biomass on suspended biofilm carrier in removal of several pharmaceuticals was demonstrated. The biomass on MBBR carriers gave considerably higher removal rates per unit biomass (i.e. suspended solids for the activated sludges and attached solids for the carriers) of diclofenac, ketoprofen, gemfibrozil, clofibrac acid and mefenamic acid, as compared to the suspended sludges. Even carrier biomasses with the poorest results show higher removal rates than the activated sludge biomasses with the best results (Falås *et al.*, 2012).

In a later study at the Bad Ragaz IFAS wastewater treatment plant in Switzerland, batch experiments with carriers and activated sludge from the same full-scale reactor were carried out in order to assess the micro-pollutant removal rates of the carrier biofilm under oxic conditions and the suspended sludge under oxic and anoxic conditions (Falås *et al.*, 2013). Clear differences in the micro-pollutant removal kinetics of the attached and suspended growth were demonstrated, often with considerably higher removal rates for the biofilm compared to the sludge. The results from model predictions and plant measurements showed that the removal efficiency of the process could be predicted with acceptable accuracy (~25%) for most of the modelled micro-pollutants. The model estimations indicated that the attached growth in hybrid (IFAS) processes can contribute significantly to removal of individual pharmaceutical compounds (Falås *et al.*, 2013).

Torresi *et al.* (2017) investigated different optimization strategies using MBBRs towards the removal of 23 commonly detected micro-pollutants (i.e. pharmaceuticals) in municipal wastewater. Overall it was found that biotransformation rates were significantly enhanced by the post-denitrification system as compared to the pre-denitrification and nitrification only system for the majority of micro-pollutants (~60%) suggesting the positive impact of easily biodegradable carbon sources (such as methanol, glycol or ethanol) on micro-pollutant removal. The removal of compounds such as propranolol, atenolol, citalopram, venlafaxine (under post-denitrifying conditions) and diclofenac (under aerobic conditions) was improved compared to conventional activated sludge. It was concluded that MBBR can offer a suitable technology that can be optimized for the removal of micro-pollutants in municipal wastewater under a range of operating conditions (nitrifying, pre- and post-denitrifying) (Torresi *et al.*, 2017).

3.7 SEPARATION OF BIOMASS FROM MBBR AND IFAS SYSTEMS

Separation of biomass from MBBR systems is different from that in activated sludge systems or IFAS systems. In contrast to activated sludge and IFAS systems,

the biomass to be separated includes only the growth since there is no recycle. This means that the SS to be separated is in the order of $150\text{--}250 \text{ mgSS} \cdot \text{l}^{-1}$ in normal municipal wastewater – around 20 times less than in activated sludge and IFAS systems. This makes it possible to combine the MBBR with any of the commonly used separation technologies; settling, flotation, micro-screening, media filtration and membrane filtration. A review is given by Ødegaard *et al.* (2010).

This section focuses on advances in separation from MBBRs through compact separation methods, such as micro-sand ballasted lamella sedimentation, dissolved air flotation and micro-screening as well some comments on separation of the suspended biomass in IFAS systems. Separation by membranes is discussed in section 8.

3.7.1 Separation characteristics of MBBR biomass

The characteristics of the biomass leaving an MBBR are different from that of activated sludge with respect to morphology as well as particle size distribution.

In an activated sludge plant, the process of developing an active biomass begins with the growth of zoogeal bacterial floc which then becomes colonized by protozoa which feed on the free swimming bacteria to produce a clarified effluent. In the moving bed process, the order of colonization seems to be reversed (Mosey, 1996). High loading rates in an MBBR (around $30 \text{ gCOD} \cdot \text{m}^{-2}\text{d}^{-1}$) produce compact bacterial biofilms, with protozoan population either absent or limited to small free-swimming protozoa and *Vorticella* spp. Moderate loading rates, around $10\text{--}15 \text{ gCOD} \cdot \text{m}^{-2}\text{d}^{-1}$ promote a more “fluffy” biofilm with a rich variety of ciliated protozoa. Low loading rates ($<5 \text{ gCOD} \cdot \text{m}^{-2}\text{d}^{-1}$) promote a biofilm generally dominated by stalked ciliates (Mosey, 1996).

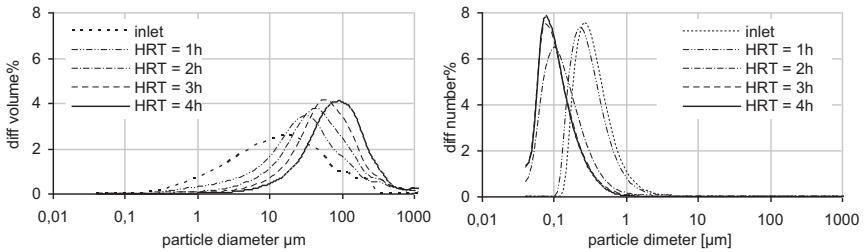
To a lesser extent than in activated sludge processes, therefore, one will find well defined aggregates of bacteria (flocs) and to a larger extent either small particles (single bacteria or clusters of bacteria) or rather large particles (biofilm flakes), which is a characteristic also for other biofilm processes (trickling filters, rotating biological contactors - RBCs etc).

Åhl *et al.* (2006) studied the particle size distributions in the bulk liquid of various reactor steps in a pilot plant, obtaining different curves at different MBBR loadings (i.e. different HRTs in the MBBR). In Figure 3.22 the particle size distributions presented as differential volume (large particles) and differential number (small particles) are shown.

It is demonstrated that particle aggregation takes place with a clear shift towards larger particles with increasing HRT (decreasing organic load) as shown in the differential volume % analysis. Aggregation of the smaller particles also results in a redistribution of the numbers of particles in the various size fractions where there is a relative increase in the number of smallest size fractions, particularly at HRTs higher than 2 hours (right graph in Figure 3.22).

In another study using the same pilot plant, Melin *et al.* (2005) analysed how the relative amount of COD in different filtered size fractions changed in the MBBR

effluents including the soluble fraction from the membrane permeates. It was demonstrated (Figure 3.23a) that the majority of the COD was found in particles >1 μm and that there was a clear decrease in colloidal particles (0.1–1 μm fraction) when the HRT increased. The results also show (Figure 3.23b) that the majority of the particles in the MBBR effluents are above 1 μm in size and it was confirmed that there is a clear shift towards smaller particles when the HRT in the MBBR is reduced. It was observed throughout the experiments that the low-rate (long HRT) MBBR produced flocs that settled much better than the high-rate.



AQ5 Figure 3.22 Development of particle size distributions as a function HRT (Åhl *et al.*, 2006).

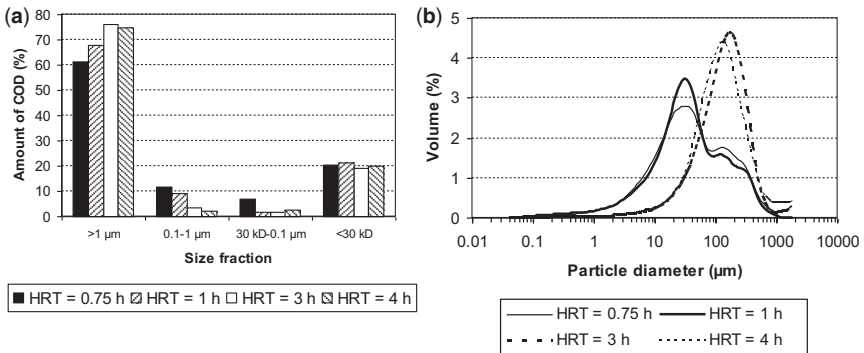


Figure 3.23 Relative amounts of COD in the different size fractions (a) and the particle size distributions (b) in the MBBR effluent (Melin *et al.*, 2005).

The above demonstrates that separation of biomass from MBBRs by conventional settling is not straight-forward and that flocculation of some sort is required to ensure optimal separation. The flocculation may be physical/chemical or biological. Quite often P-removal is required and metal salt coagulation followed by flocculation takes care of the fine particle challenge. Biological flocculation will take place in IFAS processes or in processes where the MBBR is directly followed by activated sludge (without any separation in between).

3.7.2 High-rate biomass separation after MBBRs

Settling is probably the most widely used method of biomass separation also after MBBRs. As demonstrated above, however, the characteristics of MBBR biomass are not favourable for sedimentation without pre-coagulation. The major portion of the MBBR sludge settles well, but especially in highly loaded MBBRs, the fine fractions ought to be coagulated in order to achieve a low SS in the effluent. In most countries, however, phosphate removal is required and coagulation/flocculation directly after the MBBR is implemented (see Figure 3.17).

Over the last decades some very compact, high-rate particle separation reactors have been more increasingly used in the water treatment sector, such as lamella settlers, flotation units and micro-screens. Since the SS concentration to be separated is much lower in biofilm systems than in activated sludge systems, these separation techniques have also been used for MBBR biomass separation in wastewater treatment. In particular dissolved air flotation (DAF) units, micro-sand ballasted lamella settlers (ACTIFLO™), and micro-screens (various types of drum-, disc- and belt-filters) have been used, and experiences from these applications will be focused on here.

3.7.2.1 Dissolved air flotation (DAF)

Dissolved air flotation (DAF) is favourably used for separating biomass from MBBRs. In Europe and North America alone there are more than 60 MBBR/DAF plants and additionally at least 25 such plants are used on-board cruise-ships (Wessman, 2017). Especially when combined with pre-coagulation, flotation gives a very good separation result (normally below $10 \text{ gSS} \cdot \text{m}^{-3}$) in addition to a concentrated sludge phase (at least 4% DS). The combination MBBR + DAF results in a very compact overall plant.

In most municipal wastewater cases the plants consist of pretreatment, MBBR, metal salt addition for phosphate removal, flocculation (optionally with polymer addition) and flotation. In Norway alone there are around 20 such plants and typically $<10 \text{ gSS} \cdot \text{m}^{-3}$ and $<0.5 \text{ gTotP} \cdot \text{m}^{-3}$ is achieved on average annually.

Flotation reactors are usually designed based on surface overflow rate and dispersion water (air saturated) flow. Typically, but quite conservatively, an overflow rate of $5.0\text{--}7.5 \text{ m} \cdot \text{h}^{-1}$ at design flow and $10 \text{ m} \cdot \text{h}^{-1}$ at maximum design flow is used (Norsk Vann, 2009). The amount of air precipitated is proportional to the product of saturated water flow and pressure. The typical saturator pressure is 400–600 kPa (4–6 bar) and the typical saturated water flow (in % of maximum design flow) is 10–25% (depending on SS_{in} and air saturation). When using the lower limit for pressure, the higher limit for saturated water flow should be used – and vice versa. The dispersion equipment should have a minimum capacity equivalent to 10% of maximum design flow at a saturator pressure of 600 kPa.

Flocculation is important and normally a multi-stage (minimum two stages) flocculator with a total residence time of 15–20 min and relatively high mixing intensity ($G = 70\text{--}90 \text{ s}^{-1}$) in all stages should be used.

In traditional DAF plants the dispersion water is made by pressurizing a return flow by the use of a compressor, a static mixer and a pressure tank (Figure 3.24a) creating bubbles in the 20–40 μm range. Nowadays it is more common to use a high pressure turbine/mixer pump together with an air saturator that can produce finer bubbles ($<10 \mu\text{m}$).

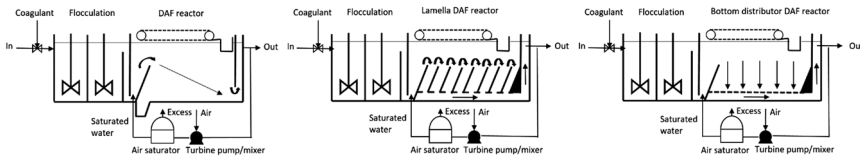


Figure 3.24 Traditional DAF and newer designs of DAF-reactors – with flow distributors.

Some newer designs of DAF-reactors are using lamellas in the clarification zone (Figure 3.24b) – mainly to ensure good hydraulic distribution. Another way of achieving the same is by introducing a reactor bottom with holes for flow distribution (Figure 3.24c).

3.7.2.2 Micro-sand ballasted lamella sedimentation

When combined with a micro-sand ballasted lamella sedimentation reactor (such as ACTIFLO™), the MBBR plant gets an extremely low footprint. Therefore, this process combination is becoming of increasing interest. Around 20 plants based on MBBR and ACTIFLO™ are in operation, especially in Norway, France and Canada, ranging in size from 4000 to 160,000 $\text{m}^3 \cdot \text{d}^{-1}$ (Wessman, 2017).

The principle of the process is shown in Figure 3.25. After metal (Al or Fe) hydroxide/phosphate precipitation, an anionic polymer is added, as well as micro-sand to the flocculation reactor. The micro-sand is enmeshed in the hydroxide/phosphate floc and kept together by the anionic polymer and flocculated (at relatively high G-values) before being separated from the water in an up-flow lamella clarifier. The main purpose of the lamellas is to ensure good hydraulic distribution. This is required because the weight of the flocs is so high, due to the enmeshed micro-sand, that it results in a very high settling velocity, in the 50–150 $\text{m} \cdot \text{h}^{-1}$ range. The micro-sand attached to the sludge is separated from the sludge in a hydro-cyclone, returned to the flocculation chamber and hence reused. However, there is a certain consumption of micro-sand, estimated by the supplier (Veolia Water Technologies) to be 2.5 $\text{g} \cdot \text{m}^{-3}$ in MBBR/ACTIFLO™ plants which means 3.75 tons per month for a 50,000 $\text{m}^3 \cdot \text{d}^{-1}$ plant. The average rise rate in the clarifier tank of the plants that are in operation is 72 $\text{m} \cdot \text{h}^{-1}$, ranging from 35 to 120 $\text{m} \cdot \text{h}^{-1}$ (Wessman, 2017). The coagulation/flocculation zone (which is especially designed for the different versions of the ACTIFLO™, is about of the same size (both in footprint and volume) as the settling zone.

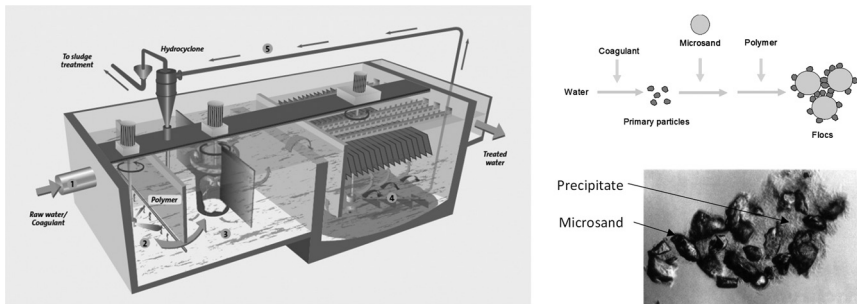


Figure 3.25 The ACTIFLO™ Turbo separator and working principle (Courtesy Veolia Water Technologies).

Most of the Norwegian MBBR/ACTIFLO™ plants are designed for secondary treatment plus P-removal and some are used in areas of extreme variation in load, for instance in touristic towns. Typically one is achieving $<10 \text{ gSS} \cdot \text{m}^{-3}$ and $<0.5 \text{ gTotP} \cdot \text{m}^{-3}$ at dosages of around $10\text{--}15 \text{ gFe} \cdot \text{m}^{-3} + 1\text{--}2 \text{ g anionic polymer} \cdot \text{m}^{-3}$. The operational experiences with ACTIFLO™ in combination with MBBR are quite good, but it is a vulnerable process. On the one hand, if the dosage of any one of the additives (metal coagulant, anionic polymer or micro-sand) fails, the whole separation process fails. On the other hand, the process is quickly restarted after such a failure. Operation costs are high because of all the additives.

3.7.2.3 Micro-screening (disc filtration)

Micro-screening, i.e. drum and disc filtration, is mainly based on sieving of particles and thus the particle size distribution is crucial to the separation result. Nominal pore size openings normally range from $10 \mu\text{m}$ to $50 \mu\text{m}$ depending upon application. The principle of operation of most devices is relatively simple and straight-forward. There are a number of different micro-screening units (drum-, disc- and belt-filters) available. Here the experiences with the so-called Hydrotech™ disc filter are referred to, see Figure 3.26.

Disc filters are normally used as a polishing step after another separation method such as sedimentation (including ACTIFLO™) or DAF. When disc filtration follows an MBBR directly without any previous separation steps, the fine particle fraction mentioned above, is a challenge that requires pre-coagulation/flocculation in order to get low effluent SS values. This was confirmed when disc filtration without any pre-coagulation/flocculation was tested in two separate pilot tests, at Nordre Follo and at Gardermoen WWTP in Norway (Ødegaard *et al.*, 2010). In order for the capacity to be sufficiently high, considering the backwash water consumption, it was found that a mesh $>20 \mu\text{m}$ ought to be used. The results in Figure 3.27 and 3.28 stem from tests with a $40 \mu\text{m}$ mesh. In these tests the applied filtration rates were in the order of $2\text{--}6 \text{ m} \cdot \text{h}^{-1}$ and in most cases not corresponding to maximum hydraulic capacity.

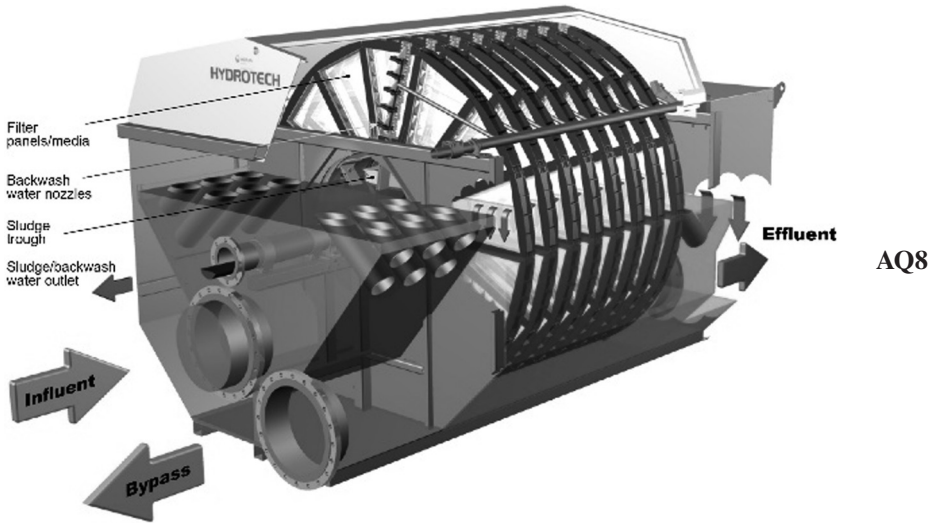


Figure 3.26 Hydretech disc filter (Courtesy Veolia Water Technologies).

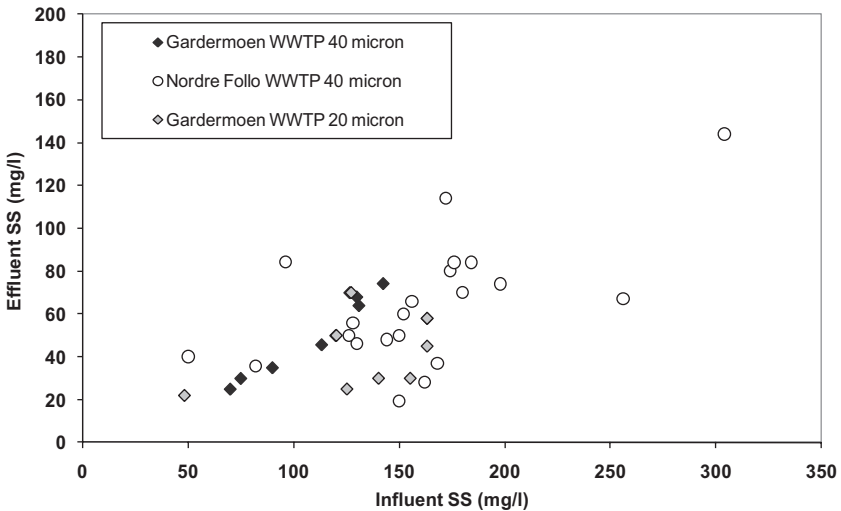


Figure 3.27 Effluent SS versus influent SS (MBBR effluent SS) without any pre-coagulation/flocculation (Ødegaard *et al.*, 2010).

Figure 3.27 shows the effluent SS versus influent SS when no coagulant was dosed, while Figure 3.28 shows the effluent SS as a function of polymer dose when a cationic polymer was dosed – as the only coagulant. When comparing the two

figures, it is demonstrated that the effluent SS could be significantly reduced by the use of polymer coagulation and flocculation ahead of the disc filter (Ødegaard *et al.*, 2010).

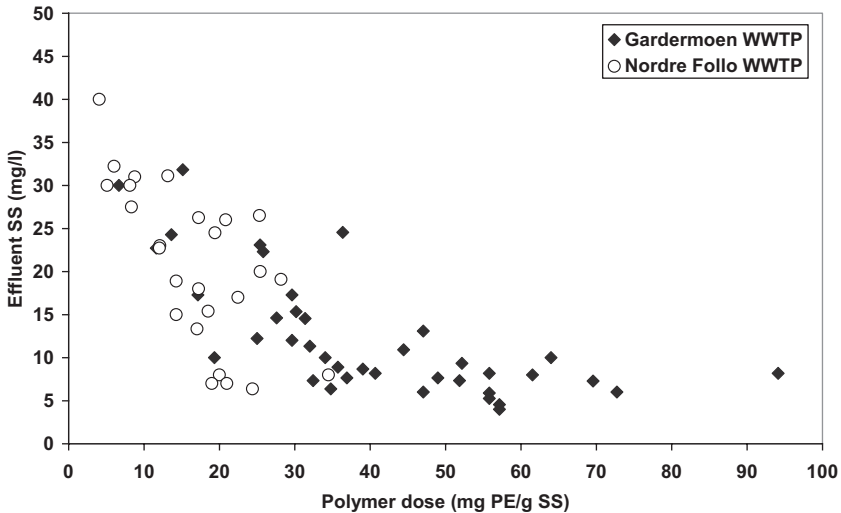


Figure 3.28 Effluent SS after disc filtration of MBBR effluent when polymer was dosed as a function of polymer dosing (Ødegaard *et al.* 2010).

Effluent-SS well below $10 \text{ mg} \cdot \text{l}^{-1}$ could be achieved when using cationic polymer at a relative high dose – above $35 \text{ mg polymer/g SS}$, which corresponds to $5\text{--}7 \text{ mg cationic polymer} \cdot \text{l}^{-1}$ at a typical MBBR effluent-SS of $150\text{--}200 \text{ mg SS} \cdot \text{l}^{-1}$. When using even higher polymer doses, the hydraulic capacity could be further increased, but the effluent SS remained relatively constant.

Disc filtration is used directly after MBBR post-denitrification in several plants, for instance for Sjölanda WWTP as well as Rya WWTP in Sweden (Nunes *et al.*, 2013; Wilén *et al.*, 2016). At Rya WWTP Hydrotech™ disc filters in two separate lines, are used to separate suspended solids from a) an effluent from secondary settlers of the non-nitrifying activated sludge system and b) biomass from the post-denitrifying MBBRs directly. The disc filter plant consists of 16 disc filters in each line corresponding to a total filtration area (area of filter cloth) of 3584 m^2 and with $15 \mu\text{m}$ cloth pore size (Nunes *et al.*, 2013). As demonstrated in Figure 3.29, the hydraulic capacity of the disc filters varies between 5 and $10 \text{ m} \cdot \text{h}^{-1}$, when the influent-SS varies between 10 and $40 \text{ gSS} \cdot \text{m}^{-3}$. It is $7.5 \text{ m} \cdot \text{h}^{-1}$ at an influent concentration of around $20 \text{ mgSS} \cdot \text{l}^{-1}$ both from the post denitrifying MBBR and from activated sludge settlers – corresponding to an average particle load of $152 \text{ gSS} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$.

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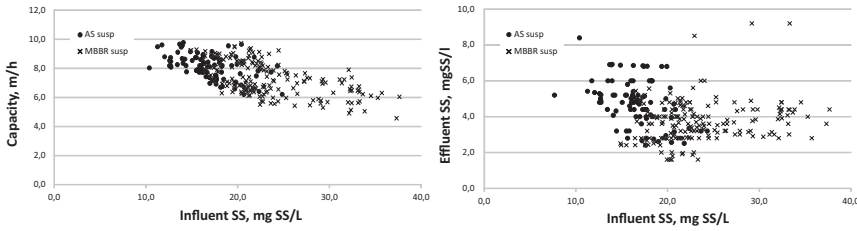


Figure 3.29 The hydraulic capacity (filtration rate) of the disc filters versus influent-SS (left) and effluent-SS versus influent-SS (right) at Rya WWTP (Nunes *et al.*, 2013).

Figure 3.29 demonstrates that the residual suspended solids concentrations were slightly lower in the effluent treating the post-denitrification MBBR biomass than that from the activated sludge settlers and well below $5 \text{ gSS} \cdot \text{m}^{-3}$. The tot. P concentration was $<0.2 \text{ gP} \cdot \text{m}^{-3}$ for approximately 60% of the days (Wilén *et al.*, 2016).

In Table 3.3 are given typical design values when using Hydrotech™ disc filter for separating biomass after an MBBR (Ødegaard *et al.*, 2010). Pre-coagulation/flocculation is normally required when separating directly after an MBBR, while it is normally not needed when used as a polishing step after the earlier coagulation/flocculation/main separation step.

Table 3.3 Hydraulic capacity and energy and backwash water consumption based on the Hydrotech™ disc filter (Ødegaard *et al.*, 2010).

Influent SS-concentration	Filtration Rate ($\text{m} \cdot \text{h}^{-1}$)		Footprint Requirement ¹ ($\text{m}^3 \cdot \text{m}^{-2} \cdot \text{h}^{-1}$)		Energy Consumption ($\text{Wh} \cdot \text{m}^{-3}$)	
	18 μm	40 μm	18 μm	40 μm	18 μm	40 μm
$<40 \text{ gSS} \cdot \text{m}^{-3}$ ²	8		75		12	
$100\text{--}200 \text{ gSS} \cdot \text{m}^{-3}$ ³		6–8		60–75		16

¹Surface loading based on required ground area, not including sidewalks for maintenance etc.

²i.e. when used directly after separate MBBR for post-denitrification or for polishing after main separation step

³i.e. when used directly after MBBR without previous separation and including coagulation/flocculation before disc filter with $7\text{--}9 \text{ mgAl} \cdot \text{l}^{-1}$ and 3–4 ppm cationic polymer dosage.

As the table demonstrates, the footprint is significantly lower than for methods like sand filtration and DAF. Energy consumption is low and principally related to backwashing of the filter (pressurizing backwash water to approximately 7 bar). Nunes *et al.* (2013) reported $0.013 \text{ kWh} \cdot \text{m}^{-3}$ at the Rya plant. The amount

of backwash water is approximately 2–4% of the total flow. Regular and frequent cleaning is of utmost importance to keep backwashing at a minimum and maintain the design capacity (Wilén *et al.*, 2016).

3.7.3 Biomass separation in IFAS systems

Several authors (Fouad & Barghava, 2005; Blank *et al.*, 2007; Briggs, 2009; Kim *et al.*, 2010; Wilson *et al.*, 2012) have all reported better AS settling properties after introducing IFAS into an existing AS plant. Figure 3.30 (Briggs, 2009) and Figure 3.31 (Wilson *et al.*, 2012) both show improvements of Sludge Volume Index (SVI) after introduction of IFAS.

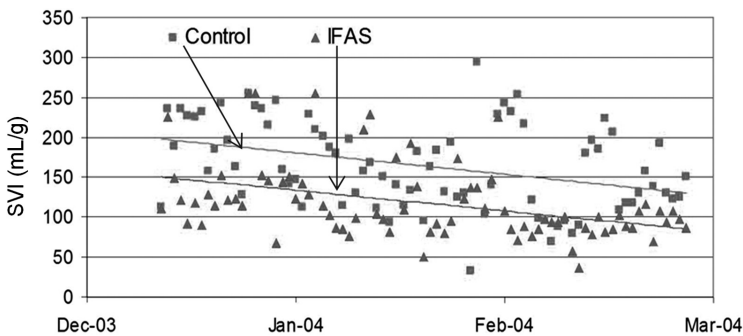


Figure 3.30 Comparison of SVI in two parallel lines, at Lakeview WWTP, Ontario, Canada (Briggs, 2009).

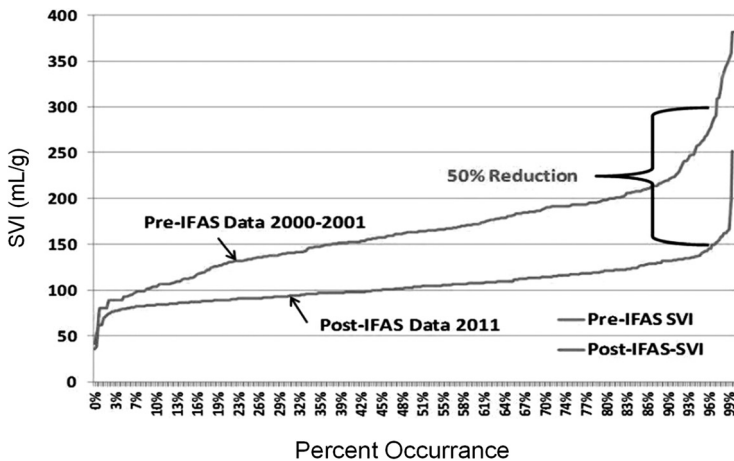


Figure 3.31 Pre- and post IFAS SVIs at Field's Point at WWTP, USA (Wilson *et al.*, 2012).

3.8 MBBR-BASED MEMBRANE BIOREACTOR (MBR) SYSTEMS

When very strict effluent standards are to be met (typically in re-use situations) membrane bioreactors (MBRs) are being favoured. Traditional MBRs are based on activated sludge with biomass separation by ultrafiltration (UF) membranes (typical pore size: 40 μm) designed for a reasonably high MLSS concentration (7–10 $\text{g MLSS} \cdot \text{m}^{-3}$) and a membrane flux of around 20 $\text{l} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$. This makes the MBR far more compact than a traditional activated sludge plant and a competitor to MBBR with respect to compactness.

The main draw-back of the traditional MBR process is the high energy consumption caused by the extra air that is used for keeping the membranes from fouling and clogging too fast. The higher energy also results in a comparatively high operation cost. Another draw-back is the need for membrane replacement because of membrane ageing and permanent fouling.

An interesting alternative is an MBBR-based MBR system that may be placed after a pure MBBR or in an IFAS-based MBBR, see Figure 3.32. The nomenclature has been confusing in literature. MBBR-MBR is often used for an MBBR-based IFAS system, but there are clear differences between an MBR system based on a pure MBBR and one based on IFAS, see Figure 3.32.

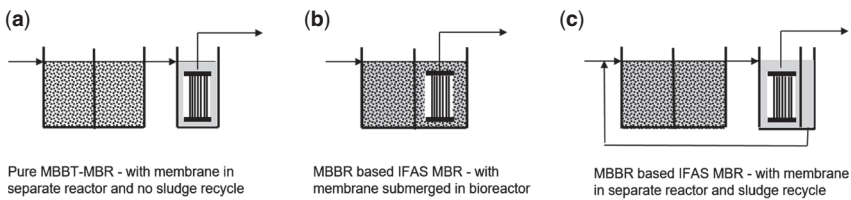


Figure 3.32 The various MBBR-based MBR systems (Ødegaard, 2017).

The system in Figure 3.32a does not have recycle from the membrane reactor and hence the primary bioreactor is a pure biofilm reactor, while the primary bioreactor the system in Figure 3.32c is a hybrid activated sludge/biofilm (IFAS) bioreactor, as is also the one in Figure 3.32b since the MLSS stays in the reactor. In this chapter, there is clear differentiation between those systems based on pure biofilm in the primary bioreactor (here called MBBR-MBR) and those based on a hybrid (IFAS) primary bioreactor (here called IFAS-MBR).

3.8.1 Pure MBBR + membrane (MBBR-MBR)

The combination of pure MBBR and membrane separation may be implemented in different ways. The membrane unit may be placed directly after the MBBR

as an immersed membrane (Figure 3.33a and 3.33b) or as a separate, contained membrane unit (Figure 3.33c and 3.33d). All these systems have been tested.

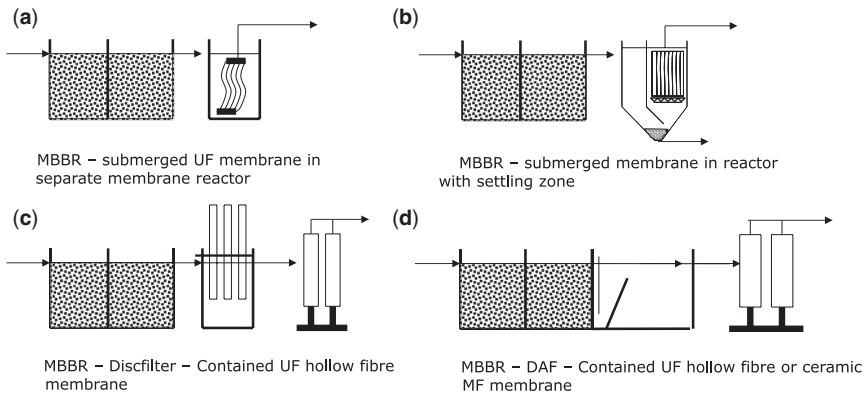


Figure 3.33 Examples of MBBR-membrane configurations.

Most of the studies on MBBR-MBR reported so far, have been based on immersed UF-membranes (Leiknes & Ødegaard, 2006; Ivanovic *et al.*, 2006; 2008; Ivanovic & Leiknes, 2012). These studies have shown that the concentration of biomass as well as the particle size distribution of the MLSS adjacent to the membrane, is influencing the fouling of the membrane and therefore several measures have been proposed in order to minimize this particle-caused fouling.

Ivanovic *et al.* (2006) demonstrated the beneficial effect of designing the membrane reactor in such a way that bio-sludge is flocculated in the reactor and continuously taken out. Melin and Ødegaard, (2007) demonstrated that coagulation of the MBBR biomass prior to separation in an immersed UF-membrane had a beneficial effect. The fouling rate in these experiments was found to be linearly proportional to the flux and it was shown that a sustainable flux (close to zero fouling rate) could be obtained at a flux of around $20 \text{ l} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$ without coagulation and around $25 \text{ l} \cdot \text{m}^{-2} \cdot \text{h}^{-1}$ with coagulation (Melin & Ødegaard, 2007).

When applying membranes directly after the MBBR, contained membrane modules are not useful since the biomass concentration to be separated would be $200\text{--}300 \text{ mgSS} \cdot \text{l}^{-1}$. Ødegaard *et al.* (2012) investigated, therefore, the use of an intermediate, high-rate biomass separation step between the MBBR and the membrane unit in order to lower the MLSS concentration entering the membrane reactor. The principle of this process solution is shown in Figure 3.33 where the high-rate biomass separation step consists of a disc filter (DF) unit (Figure 3.33c) or a dissolved air flotation unit (DAF) (Figure 3.33d). It could alternatively be a micro-sand ballasted lamella settling unit (ACTIFLO™).

Ødegaard *et al.* (2012) reported on pilot-scale experiments with real municipal wastewater from the Gardermoen WWTP in Norway. The full-scale plant is a combined pre- and post-denitrification MBBR-plant with coagulation/flocculation and DAF separation. The water that was to be separated, could be taken (a) directly after the full-scale MBBR and passed through coagulation/flocculation + disc filter + UF membrane pilot module (referred to below as MBBR + DF + UF), or (b) from the effluent of the full-scale plant (with DAF separation) and passed through the UF pilot module (referred to below as MBBR + DAF + UF) – without any coagulant added directly ahead of the membrane pilot module.

Both solutions would be very compact in practice, but the use of a 40 µm disc filter (DF) for the primary biomass separation ahead of the UF would give the smallest footprint. In order for the DF alternative to be optimized, a cationic polymer had to be used, however, (either alone or in combination with a metal salt like aluminium) ahead of the disc filter (DF).

This process combination is quite risky, therefore, because of the possibility of polymer fouling of the membrane. An interesting alternative, based on DF, would be one with controlled polymer dosing ahead of the DF and aluminium dosing after the DF and before the UF. The hypothesis was that the hydroxide precipitation caused by the aluminium precipitation would sweep possible polymer residuals and hence reduce the risk of membrane fouling (Ødegaard *et al.*, 2012).

It was found, however, that the solution based on MBBR – coagulation/flocculation – DAF – UF was the one with the lowest fouling, the highest possible operating flux, the lowest backwash water consumption and the lowest membrane cleaning chemicals consumption.

Even though a fair amount of R&D has been executed and a number of plants have been built (e.g. Darling Walk project in Sydney (Shaw, 2017), plants for pure MBBR with biomass separation (MBBR-MBR) are not yet common.

3.8.2 MBBR based hybrid MBR (IFAS MBR)

Most studies in lab- and pilot-plants on MBBR-based MBR systems have been carried out in MBBR-based hybrid MBRs (IFAS MBR). A comprehensive review is given in Leyva-Díaz *et al.* (2016).

Many research studies have been performed at lab-scale in a system such as the one in Figure 3.32b (Duan *et al.*, 2015; Leyva-Díaz, 2016; Tang *et al.*, 2016). Most new full-scale MBR plants today use, however, separate membrane tanks with sludge recycle, which is the IFAS-solution as in Figure 3.32c. The one in Figure 3.33b is restricted to small plants. One benefit that has been reported for such a system (Figure 3.33b), is reduced fouling as a consequence of reduced foulant deposition as well as abrasion of the cake layer on the membrane – caused by the carriers (Wei *et al.*, 2006; Yang *et al.*, 2009).

Most of the studies that compare the pure MBBR-MBR and the IFAS-MBR go in the favour of IFAS-MBR (Leyva-Díaz *et al.*, 2016) and most of the studies

AQ7 that compare the traditional MBR and the IFAS-MBR go in the favour of IFAS-MBR (Duan *et al.*, 2015a,b). The IFAS-MBR is claimed to have the advantage of operating at higher fluxes, being more compact, having better energetic efficiencies and better membrane fouling control than the traditional MBR. Hence the IFAS-MBR should be the most interesting system, taking advantage of the best qualities of suspended biomass, attached biomass and membrane separation systems. In many studies, however, the SRT_{MLSS} and HRT in the IFAS-MBR has been about the same as in conventional MBRs (i.e. SRT_{MLSS} : 15–30 d, HRT: 10–24 h). In this author's opinion, this does not utilize the full potential of IFAS-MBR in the practical scale where size and cost matters. The benefit of IFAS is that nitrification can be obtained at less than half of the SRT_{MLSS} than what is required in traditional activated sludge plants (Ødegaard *et al.*, 2014).

Moreover, at a lower SRT_{MLSS} more carbon will be available for denitrification through hydrolysis of the MLSS and several studies (Rusten *et al.*, 2003; Onnis-Hayden *et al.*, 2011) have shown that the specific denitrification rate in IFAS systems is around twice as high as in conventional activated sludge systems. Hence there is the potential for the IFAS MBR system as compared to the conventional MBR system, to have a much smaller bioreactor since the nitrification as well as the denitrification zone may both be smaller.

3.9 A COMPARISON BETWEEN MBBR-, MBR- AND IFAS MBR SYSTEMS

In order to evaluate the practical full-scale potentials of the MBBR-based MBR processes for advanced wastewater treatment, a comparative analysis between conventional MBBR/DAF-, conventional MBR- and IFAS MBR processes was carried out (Ødegaard, 2017). The study is based on data for an existing, large wastewater treatment plant that is to be upgraded within very limited space, and with very strict effluent standard ($TotN < 4 \text{ g} \cdot \text{m}^{-3}$, $TotP < 0.3 \text{ g} \cdot \text{m}^{-3}$) with an average flow of $190,000 \text{ m}^3 \cdot \text{d}^{-1}$, a peak flow of $15,833 \text{ m}^3 \cdot \text{h}^{-1}$ and the wastewater characteristics given in Table 3.4.

Three basic alternative processes were analysed (with a number of sub-alternatives (see Figure 3.34):

- MBBR with subsequent advanced particle separation based on DAF + Disc filter (DF)
- MBR (activated sludge-based MBR with UF membranes for biomass separation)
- IFAS-MBR (IFAS with carriers in aerobic reactors only and with UF membranes for biomass separation)

The MBBR alternative is based on combined pre-and post-denitrification and chemical P-removal. The MBR alternatives (MBR and IFAS-MBR) are also based on pre- and post-denitrification (with and without external carbon source), and

with one sub-alternative based on biological P-removal (with external carbon source). The three basic processes were analysed with three different pretreatment alternatives: (a) no primary (fine screen only), (b) primary treatment (based on settling), and (c) chemically enhanced primary treatment (CEPT, based on coagulation/flocculation/settling). The processes are compared with respect to bioreactor volume, construction volume and construction footprint at two different climate situations/wastewater temperatures (design temperature 10°C and 18°C).

Table 3.4 Composition of the wastewater for which the process design will be made.

Parameter	Raw Water	After Primary	After CEPT	Effluent Standard 95%-ile	Average Used for Design
Biochemical Oxygen Demand (BOD ₅) (mg/l)	200	130	70	<10	<5
Chemical Oxygen Demand (COD) (mg/l)	500	360	150	–	–
Total Suspended Solids (TSS) (mg/l)	250	135	30	<10	<5
Ammonia Nitrogen (NH ₃ -N) (mg/l)	30	28	28	<6	<3
Total Kjeldahl Nitrogen (TKN) (mg/l)	50	40	35	<8	<4
Ortho-Phosphate (OP) (mg/l)	3.0	2.7	1.0	–	–
Total Phosphorus (TP) (mg/l)	6.5	4.6	2.0	<0.5	<0.25

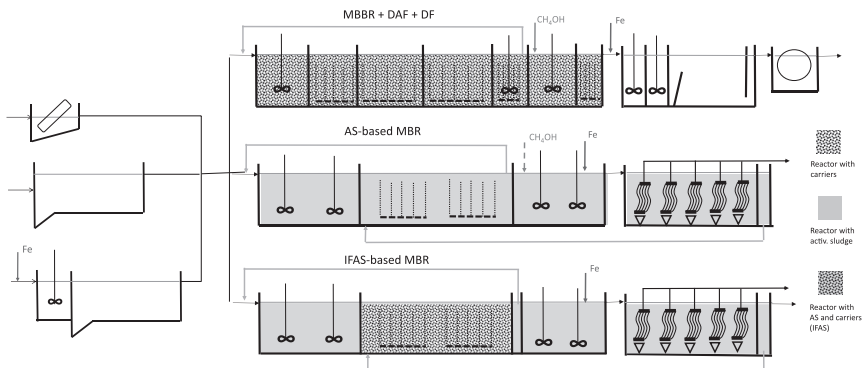


Figure 3.34 The basic wastewater treatment processes analysed and compared.

The MBBR + DAF + DF alternative is based on a pre-denitrification step in two stages, a BOD-removal step, a nitrification step in two stages plus a de-oxygenation step, a post-denitrification step in two stages and a re-aeration step. A carrier with a bulk specific surface of $800 \text{ m}^3 \cdot \text{m}^{-2}$ is used with filling fractions of 60% in aerobic and 50% in anoxic reactors respectively. The reactor volume design rates are at 10°C (18°C in brackets) – average if two stages; pre-DN: 0.6 (0.95) $\text{g NO}_x\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, BOD-removal: 3.6 (6.25) $\text{g BOD}_5 \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, nitrification: 0.5 (1.0) $\text{g NH}_4\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, O_2 -depletion: 0.225 (0.45) $\text{g NH}_4\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, post-DN: 0.6 (0.95) $\text{g NH}_4\text{-N} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$, re-aeration: $3 \text{ g COD} \cdot \text{m}^{-2} \cdot \text{d}^{-1}$. The design DO is $2 \text{ g O}_2 \cdot \text{m}^{-3}$ in the BOD-removal reactor and $5 \text{ g O}_2 \cdot \text{m}^{-3}$ and $4 \text{ g O}_2 \cdot \text{m}^{-3}$ in the two nitrification stages respectively, and $2 \text{ g O}_2 \cdot \text{m}^{-3}$ in the de-oxygenation tank. The recycle from the aerobic (deoxygenation) step to pre-denitrification is 300%. External carbon in the form of methanol is added to the post-denitrification step. Iron chloride is added to the effluent of the bioreactor before flocculation (in two stages) and biomass/floc separation in the DAF and subsequent micro-screen (disc filter).

Because of the relatively low C/N-ratio, an external carbon source has to be used in the MBBR-alternative. In the MBR- and IFAS MBR alternatives, hydrolysis will be more extensive and it may be possible to meet the required standard without an external carbon source if an extra post-anoxic volume for endogenous denitrification is included.

In the traditional MBR alternative two sub-alternatives are analysed, therefore, with and without external carbon source added. The MBR bioreactor is based on a pre-anoxic step (pre-denitrification), an aerobic step (residual BOD-removal and nitrification) and a post-anoxic step (post-denitrification). The pre-denitrification step was designed for a denitrification rate of 2.2 and $1.3 \text{ gNO}_x\text{-N} \cdot \text{kgMLSS}^{-1} \cdot \text{h}^{-1}$ at 18°C and 10°C respectively. The post-denitrification step was designed for a high denitrification rate with external carbon source (methanol) added (6.25 and $3.6 \text{ gNO}_x\text{-N} \cdot \text{kgMLSS}^{-1} \cdot \text{h}^{-1}$ at 18°C and 10°C respectively) and for a low (endogenous) denitrification rate when carbon source was not added (0.75 and $0.45 \text{ gNO}_x\text{-N} \cdot \text{kgMLSS}^{-1} \cdot \text{h}^{-1}$ at 18°C and 10°C respectively). Iron chloride was added to the post-anoxic step. The nitrification step was designed based on sludge age according to accepted design rules (such as DWA A-131 and DWA – M 227) ($\text{SRT} = 3.8 \text{ d}$ at 18°C and process factor = 1.5).

When dimensioning the bioreactor in MBR systems, one may include (or not) the biomass in the membrane tank when designing for the nitrification tank volume. Membrane fouling is, however, very dependent upon the extent of organic matter degradation in the water that is to be membrane separated, which again is dependent upon the SRT_{MLSS} . Because the SRT_{MLSS} is lower in the IFAS MBR, and in order to be able to compare on equal basis, the biomass of the membrane tank is not included in the sizing of the nitrification tank. It is included, however, in the calculation of the aerobic SRT_{MLSS} . The design MLSS in the aerobic part of the

activated sludge reactor was set at $8 \text{ kgMLSS} \cdot \text{m}^{-3}$, resulting in $6.4 \text{ kgMLSS} \cdot \text{m}^{-3}$ in the pre-denitrification reactor, and it was set at $10 \text{ kgMLSS} \cdot \text{m}^{-3}$ in the membrane reactor. The design DO in the aerobic tank of the bioreactor was set at $2 \text{ gO}_2 \cdot \text{m}^{-3}$.

The IFAS-MBR alternative was designed in principle exactly in the same way as the MBR alternative, except for the fact that carriers were used in the aerobic (nitrification) stage of the bioreactor. A carrier with a bulk specific surface of $800 \text{ m}^3 \cdot \text{m}^{-2}$ was used, at a filling fraction 50%. No carriers were used in anoxic reactors. As for the conventional MBR system, the biomass in the membrane reactor was also not included when sizing the aerobic (IFAS) stage of the bioreactor. As mentioned above, the pre-denitrification rate in IFAS systems is higher than that in traditional CAS systems and was designed at a pre-denitrification rate of 4.0 and 2.3 $\text{gNO}_x\text{-N} \cdot \text{kgMLSS}^{-1} \cdot \text{h}^{-1}$ at 18°C and 10°C respectively while the post-denitrification step was designed for the same rates as the conventional MBR system (see above).

In most of the alternatives, P-removal is carried out with iron precipitation. The use of bio-P was also evaluated and is discussed briefly below. It became clear, however, that the carbon available in the raw water was not sufficient for both denitrification and bio-P removal. So either the system has to use precipitating chemicals anyway or extra carbon source has to be dosed. Methanol is not useful for bio-P bacteria and sodium acetate was used.

The membrane step consists of a submerged membrane unit (hollow fibre) and a subsequent de-oxygenation stage on the sludge return. The volumes of the membrane tank and the flocculation/DAF tanks are very close to equal. The disc filter station (in the MBBR alternative) is placed above ground and no volume is calculated, but rather the surface space required. The MBR alternatives need a fine sieve between the bioreactor and the membrane reactor to prevent hair etc. clogging the membrane. The DF in the MBBR alternative and the fine sieve in the MBR alternatives have about the same footprint.

Below some parameters related to the size and cost are analysed. One has chosen to highlight bioreactor volume and construction volume as well as construction area (footprint). In Table 3.5 the required process volumes for the most probable process alternatives are compared for three alternatives of primary treatment at 18°C and 10°C . The volumes are given relative to the volume for the MBR without external carbon source alternative.

Table 3.5 shows that the bioreactor volume needed for the MBBR alternative (with external carbon source) is about 9% smaller than the MBR alternative without carbon source, but about 9% larger if carbon source is used also in the MBR alternative. The bio-P MBR alternative requires considerably more bioreactor volume – 24% more than the MBR alternative. The bio-P alternative is without methanol addition for denitrification, but needs acetate for bio-P. Actually this alternative cannot guarantee that the effluent tot P-standard is met.

Quite often there is a discussion regarding which alternative is the more compact one. This was investigated by analysing the construction volume as well as the

construction footprint. These two parameters are the most decisive ones for the investment cost.

Table 3.5 Bioreactor volumes at different primary treatments relative to the one based on MBR without C-source and primary settling at 18°C and 10°C. Biomass in membrane tank is not included as active. In brackets: volume at 10°C relative to that at 18°C for the same process.

Treatment System	18°C			10°C		
	Primary	No Primary	CEPT ⁶	Primary	No Primary	CEPT ⁶
MBBR + DAF + DF ¹	0.91	1.14	0.82	0.83 (1.71)	1.03	0.72
MBR						
No – C-source ²	1.00	1.19	–	1.00 (1.89)	1.30	–
C-source ³	0.81	–	0.71	0.86 (1.99)	–	0.60
Bio-P ⁴	1.24	–	–	1.23 (1.88)	–	–
IFAS-MBR ⁵	0.62	0.71	0.39	0.59 (1.85)	0.68	0.37

¹With Fe-addition for P-precipitation and with external carbon source (methanol).

²With Fe-addition for P-precipitation and without external carbon source.

³With Fe-addition for P-precipitation and with external carbon source (methanol).

⁴With bio-P removal (with acetate as external carbon source), no Fe-addition or methanol addition.

⁵With Fe-addition for P-precipitation and without external carbon source.

⁶P-precipitation carried out in CEPT. All alternatives with CEPT have external carbon source addition.

The construction volume was calculated by increasing the process volumes by 5% as well as adding 1 m from water level to top of basin (free-board). The construction area was increased by 10% of the area calculated as volume divided by depth, to allow for gangways etc. The volume of the primary was included. Traditional settling tanks with a peak flow overflow rate of 4.8 m · h⁻¹ were used for primary settling and the same tanks were used in chemically enhanced primary settling – but now with lamellas in the tank securing an overflow rate based on projected area of >0.8 m · h⁻¹ at peak flow. This means that the footprint surface area will be the same in the two alternatives. The water depth of the primary (as well as flocculation in CEPT) was 4 m (3 m under lamella in CEPT). The depths of the MBBR- and MBR tanks were 8 m and the DAF (including flocculation)- and membrane tanks were 3.5 m.

Table 3.6 shows the construction volumes and construction footprint at different primary treatments relative to the one based on MBR without C-source.

Table 3.6 shows that the construction volumes needed, as well as the construction footprint for the whole plant, of the MBBR alternative and the MBR alternative,

are quite similar. The operation cost is estimated, however, to be about 20% lower in the MBBR alternative than in the MBR alternative even if external carbon is not used in the latter. It is interesting to note, though, that the IFAS MBR alternative is clearly the process that requires the smallest construction volume as well as the smallest footprint. To this author's knowledge, the IFAS-MBR is not used in full-scale plants yet. This analysis proves, however, that it has the potential of becoming a very competitive alternative.

Table 3.6 Construction volumes and construction footprint at different primary treatments relative to the one based on MBR without C-source (18°C). Conditions are as described in the footnotes of Table 3.5.

Treatment System	Construction Volume Required (m ³) ¹			Construction Footprint (m ²) ¹		
	Primary	No Primary	CEPT	Primary	No Primary	CEPT
MBBR + DAF + DF	0.94	0.88	0.92	0.96	1.06	0.92
MBR ¹						
No – C-source	1.00	0.92	–	1.00	1.09	–
C-source	0.87	–	0.88	0.91	–	0.92
Bio-P	1.16	–	–	1.11	–	–
IFAS-MBR ¹	0.74	0.59	0.62	0.83	0.87	0.71

¹Biomass in membrane tank not included as contributing.

3.10 SUMMARY AND CONCLUSIONS

Since its introduction around 25 years ago, the moving bed biofilm reactor (MBBR) has established itself as a well-proven, robust and compact reactor for wastewater treatment. The efficiency of the reactor has been demonstrated in many process combinations, both for BOD removal and nutrient removal and for small as well as large plants. Even though it has been focused on municipal wastewater applications in this chapter, the reactor has been used extensively also for industrial wastewater treatment, particularly in the food-, the pulp and paper- and the aquaculture industries.

The reason for the popularity is the simplicity and the compactness of the MBBR and the solid scientific basis on which the MBBR-based process was developed. Even though the MBBR was invented and developed as a pure biofilm reactor, there is now an increasing use of the reactor principle also in hybrid systems, i.e. IFAS processes – especially for upgrading and capacity increase of existing activated sludge plants.

The primary advantage of the MBBR as compared to activated sludge reactors is its compactness and the lack of need for sludge recirculation. The advantage over

other biofilm processes is its flexibility. One may use almost any reactor shape and one may choose different operating loadings in a given reactor volume, simply by choice of carrier filling fraction.

The MBBR has favourably been used for nitrification as well as for denitrification. High nitrification and denitrification rates have been demonstrated even at low temperatures. When using the MBBR process for nitrogen removal in carbon-deficient waters, a combined pre- and post-denitrification process is recommended because this process combination is superior when it comes to process control and performance.

The MBBR processes are normally combined with chemical P-removal and this process combination may result in extremely compact plants when high-rate biomass/floc separation processes are used. In this paper results and experiences are reported from the upcoming use of micro-sand ballasted lamella sedimentation (ACTIFLO™), flotation (DAF), micro-screening (disc filtration) as well as membrane (UF) filtration. The possibility of using these compact separation processes gives MBBR processes great flexibility. It is a considerable advantage over activated sludge processes, where one has to rely on sedimentation or membrane filtration (in MBRs) for biomass separation.

MBBR-based processes (MBBR and IFAS) have successfully been used in practice for de-ammonification in side-stream water (sludge reject) and lately de-ammonification has also been successfully demonstrated in main-stream. MBBR-based IFAS systems are most favourable in one-stage systems, but pure MBBR systems may be shown to be more favourable in two-stage systems.

MBBR-based processes (MBBR and IFAS) seem to be more efficient for the removal of organic micro-pollutants (OMP) than activated sludge processes.

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