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FEATURE ARTICLE

A road-map for energy-neutral wastewater treatment plants of the future based on compact technologies (including MBBR)

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HIGHLIGHTS

- In the paper concepts for wastewater treatment of the future are discussed by the use of a) one flow diagram based on established, compact, proven technologies (i.e. nitrification/denitrification for N-removal in the mainstream) and b) one flow diagram based on emerging, compact technologies (i.e.de-ammonification in the main stream). The latter (b) will give an energy-neutral wastewater treatment plant, while this cannot be guaranteed for the first one (a). The example flow diagrams show plant concepts that a) minimize energy consumption by using compact biological and physical/chemical processes combined in an optimal way, for instance by using moving bed biofilm reactor (MBBR) processes for biodegradation and high-rate particle separation processes, and de-ammonification processes for N-removal and b)maximize energy (biogas) production through digestion by using wastewater treatment processes that minimize biodegradation of the sludge (prior to digestion) and pretreatment of the sludge prior to digestion by thermal hydrolysis. The treatment plant of the future should produce a water quality (for instance bathing water quality) that is sufficient for reuse of some kind (toilet flushing, urban use, irrigation etc.). The paper outlines compact water reclamation processes based on ozonation in combination with coagulation as pretreatment before ceramic membrane filtration.

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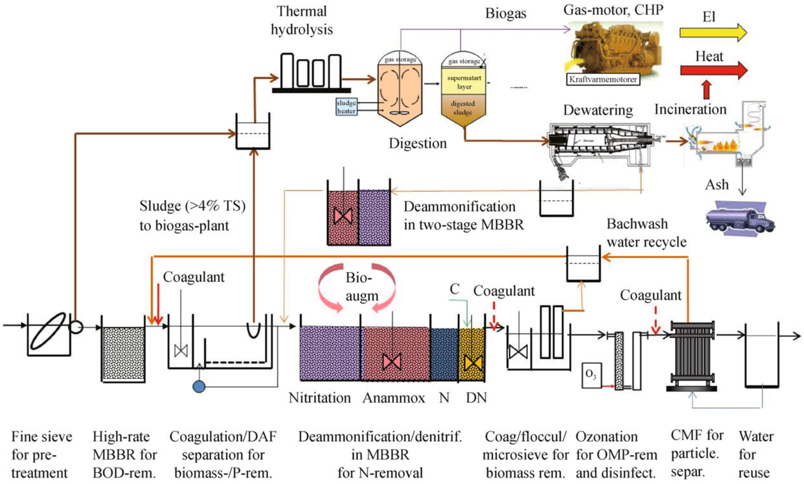
GRAPHIC ABSTRACT



ABSTRACT

In the paper concepts for domestic wastewater treatment plants of the future are discussed by the use of a) one flow diagram based on established, compact, proven technologies (i.e. nitrification/denitrification for N-removal in the mainstream) and b) one flow diagram based on emerging, compact technologies (i.e. de-ammonification in the main stream).The latter (b) will give an energy-neutral wastewater treatment plant, while this cannot be guaranteed for the first one (a). The example flow diagrams show plant concepts that a) minimize energy consumption by using compact biological and physical/chemical processes combined in an optimal way, for instance by using moving bed biofilm reactor (MBBR) processes for biodegradation and high-rate particle separation processes, and de-ammonification processes for N-removal and b)maximize energy (biogas) production through digestion by using wastewater treatment processes that minimize biodegradation of the sludge (prior to digestion) and pretreatment of the sludge prior to digestion by thermal hydrolysis. The treatment plant of the future should produce a water quality (for instance bathing water quality) that is sufficient for reuse of some kind (toilet flushing, urban use, irrigation etc.). The paper outlines compact water reclamation processes based on ozonation in combination with coagulation as pretreatment before ceramic membrane filtration.

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1 Introduction

In China the Expert Committee for China's Concept WWTPs has been established with the goal to build several concept wastewater treatment plants over the next five years. The concept wastewater treatment plants aim to achieve sustainable water quality, energy self-sufficiency, resources recovery and be environmentally friendly. It is an ambitious attempt to provide real examples for all the WWTPs to follow in future [1].

The goals of the wastewater treatment plant of the future may be summarized as follows:

- The quality of the WWTP effluent should not expose negative impact on the receiving water
- The wastewater resources should be recovered, i.e. treated water (water reuse), energy and nutrients (especially phosphorous)
- Wastewater biosolids (sludge) should be used as a resource – not a waste, and the final biosolids (sludge) production should be low
- Compact treatment processes should be used since the availability of space is increasingly more limited in urban areas and plants have to be placed under roof or under ground
- The WWTP should be energy self-sufficient and have a low carbon foot-print which means that processes that can minimize the overall energy consumption without compromising the other goals should be selected

There are, of course, many roads to the goal and it is the intention of the Expert Committee for China's Concept WWTPs that demonstration plants should be set up based on various processes – in order to evaluate their suitability [1]. In doing so, it is necessary to be open-minded with respect to process selection and not be deadlocked with the traditional wastewater treatment processes, such as the activated sludge process.

The purpose of this paper is to discuss how a concept WWTP could be built up. Two possible flow diagrams are used for demonstration; one based on proven and established compact treatment technologies (i.e. nitrification/denitrification for N-removal) and one based on emerging compact technologies (i.e. de-ammonification for nitrogen removal) (Figs. 3 and 4).

To meet the goals of the wastewater treatment plant of the future (see above), there are some very important issues that have to be taken into account:

1) Energy consumption for process energy as well as for heating/cooling and ventilation should be minimized. It is especially important to:

a. reduce the amount of air needed

b. reduce pumping (for instance recirculation pumping)

c. use compact processes in order to minimize foot-print (in plants under roof or underground)

2) Energy has to be recovered from biogas produced by anaerobic digestion through the use of combined heat and power (CHP) technologies. Biogas production should be enhanced by:

a. producing a sludge with high biogas potential (little biodegraded prior to digestion)

b. using sludge pretreatment through thermal hydrolysis process (THP)

3) Organic micro-pollutants and microbial contaminants have to be dealt with (removed)

2 Energy consumption issues

2.1 Energy consumption benchmarking

Several benchmarking investigations have been carried out throughout the world, showing great variation in energy consumption, typically ranging from 0.3 to 0.6 kWh $\cdot m^{-3}{}_{\text{treated water}}$ for a biological nutrient removal plant [2–4].

To become energy-neutral, in the order of 0.3–0.35 kWh $\cdot m^{-3}$ has to be produced. This is achieved, for instance, at the Strass WWTP in Austria, where the energy consumption is reported to be 0.211kWh$\cdot m^{-3}{}_{\text{water}}$ (0.31 kWh$\cdot m^{-3}$ if non-process demands are included) [5]. This plant is often used as a benchmark for what is achievable in a BNR plant based on activated sludge for wastewater treatment and anaerobic digestion for sludge treatment and energy recovery.

The most efficient way to reduce energy consumption is by reducing the air needed for aerobic, biological degradation in the main stream and the two most efficient ways to achieve that, is to:

a. implement de-ammonification instead of nitrification/denitrification processes for N-removal

b. separate organic matter ahead of nitrogen removal stage by using little (or no) air

2.2 Reduction of air (and hence energy) consumption

2.2.1 Implementation of de-ammonification

De-ammonification is a biological treatment process to convert ammonia to nitrogen gas without the need for carbon source (as in denitrification) and at a much lower air need than in nitrification. The principle of the process is shown in Fig. 1 [6].

De-ammonification is accomplished by two biological process steps as shown in Fig. 1. The first is termed nitritation, which is the aerobic oxidation of ammonia-N (NH_4-N) to nitrite-nitrogen (NO_2-N) by autotrophic aerobic ammonia oxidizing bacteria (AerAOB). Since stoichiometric only about half of the ammonia needs to be converted to nitrite, this is also known as partial nitritation. Nitritation is well known in wastewater treatment, as it is the initial step in biological nitrification of ammonia-N to

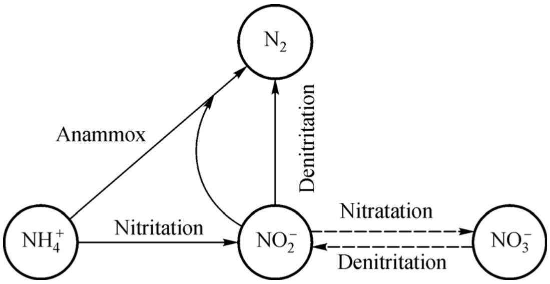


Fig. 1 Nitrogen transformation processes [6]

nitrate-N. In the second step, which is the anammox (anaerobic ammonia oxidation) reaction, NH_4-N is oxidized under anaerobic conditions by anaerobic ammonia-oxidizing bacteria (AnAOB) that can use NO_2-N as the electron acceptor. About 89% of the inorganic nitrogen (NH_4-N + NO_2-N) fed to the process ends up as N_2 gas and about 11% as NO_3-N [6].

There are several obvious advantages of the partial nitritation/anammox process, hereafter referred to as deammonification, as compared to nitrification/denitrification for N-removal; the energy for air is 60% lower, no carbon source is needed, the alkalinity consumption is 50% lower, and the sludge production is in the order of 70% lower (if organic particles are removed up-stream) [4,6]. The process is, however, not established as a proven process for nitrogen removal in the main-stream yet. Since it is easiest implemented when the temperature is high (preferably >~25°C), the ammonium concentration is high (>500 mg NH_4-N/L) and the C/N-ratio is low, the process has been successfully implemented for treatment of sludge reject water, that normally represents around 20%–25% of the nitrogen load on a typical BNR plant. Deammonification in the main-stream is still in development.

2.2.2 Separation of particulate BOD ahead of the biological, N-removing step

In traditional activated sludge BNR-plants energy is used for aeration in order to degrade BOD and ammonium at a long SRT, resulting in high energy consumption and sludge with low biogas potential. The sensible strategy for maximizing biogas production is rather to do the opposite; to capture easily biodegradable organic matter in the sludge without using much energy.

The major part of the organic matter in wastewater is present in the form of particles – a fact that was established many decades ago – as demonstrated in Table 1 [7,8]. Similar distribution of particulate organic matter has been demonstrated in Chinese wastewater [9].

Table 1shows that not only is 70%–75% of the organic matter normally present as suspended or colloidal matter, but also that the soluble matter (here defined as the one that passes a 0.08 μm filter) is biodegraded at a much higher rate than the particulate organic matter >1 μm. This is caused by the fact that the latter has to be hydrolyzed before it can pass the bacteria cell membrane and be degraded inside the cell.

Normally pre-settling, that only removes 25%–35% of the organic matter, is used. By the use of pre-coagulation that will remove suspended particles as well as colloids, 65%–75% removal of organic matter can be achieved [9,10]. By combining coagulation with a high-rate biological process (for instance high rate MBBR – see below), more than 85% of BOD may be removed [11,12].

3 Energy (biogas) production issues

There are two ways by which biogas production from sludge may be enhanced compared to traditional treatment:

a. by producing a sludge with a high biogas potential

b. by increasing biodegradability of the sludge through sludge hydrolysis pretreatment

3.1 Production of sludge with high biogas potential

As demonstrated above, a large portion (around 70%) of the BOD may be separated from the wastewater by the use of pre-coagulation. Precipitation of hydroxide may be minimized by using cationic polymer in combination with a low dose of inorganic coagulant (i.e. iron) on order to minimize the production of inorganic sludge [10].

The portion of soluble BOD may also be taken out swiftly by the use of a high rate biological step combined with the chemical step, designed in such a way that the biomass only removes the soluble BOD, while the coagulant takes care of the particulate BOD – as

Table 1 Fractionation of organic matter in wastewater – some early American studies [7,8]

size range	soluble < 0.08 μm	colloidal 0.08 – 1.0 μm	supra-colloidal 1 – 100 μm	settleable >100 μm
COD/(% of total)	25	15	26	34
BOD/(% of total)	31	14	24	31
grease/(% of TS)	12	51	24	19
protein	4	25	45	25
carbohydrates	58	7	11	24
biochemical oxidation rate/(d^{-1})	0.39	0.22	0.09	0.08

demonstrated by the use of the high rate MBBR process [11,12].

3.2 Enhanced biogas production through sludge hydrolysis pretreatment

Biogas production may be enhanced through sludge hydrolysis pre-treatment by physical, biological, chemical and thermal methods [13]. Most efficient is thermal hydrolysis. Thermal hydrolysis (treatment) involves heating/boiling of the sludge, usually to a temperature in the range of 150–170°C [13]. The yield of organic matter from hydrolysis, defined as COD_F (sample filtered through ≈ 1 μm filter) of the hydrolysate (supernatant) divided by the total COD of the sludge before treatment, is quite linear between these temperatures. At 150°C the yield can be expected to be in the range of 15% to 20%, while a yield of 28% was found when treating MBBR-sludge from a pre- and post-denitrification MBBR pilot plant at 180°C [14]. The increased content of easily biodegradable organic matter in the hydrolysate was found to be distributed as 20%–40% of the COD_F as volatile fatty acids (VFA), 30%–60% as proteins, 20%–25% as carbohydrates and 10%–20% as unknowns [14]

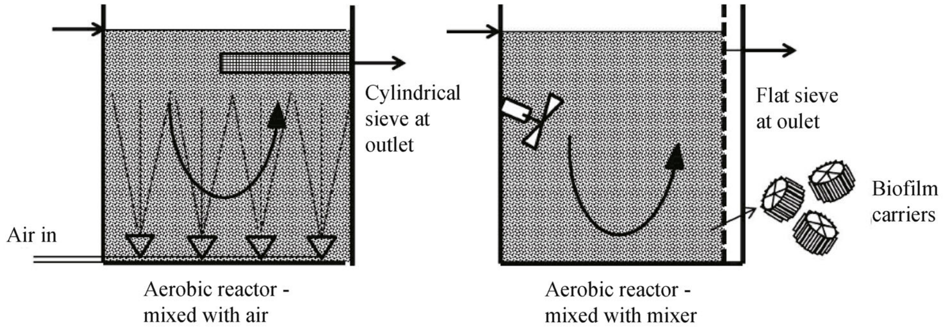
Thermal hydrolysis will results in an increased ammonium concentration in the hydrolysate and in the reject water from dewatering of the digester sludge. A linear relationship between the concentration of volatile fatty acids and ammonium in the hydrolysate can be expected: $[COD_{VFA}] = 11.0[NH_4\text{-}N]$ [14].

4 The compact moving bed biofilm reactor

The Moving Bed Biofilm Reactor (MBBR) was invented and developed in Norway around 25 years ago [15], and is now an established and proven compact technology for biological wastewater treatment [16]. It utilizes the whole tank volume for biomass growth, as does also the activated sludge reactor. Contrary to the activated sludge reactor, however, it does not need any sludge recycle, as is also the case for other biofilm reactors. This is achieved by having the biomass grow on carriers that move freely in the water volume of the reactor, kept within the reactor volume by a sieve arrangement at the reactor outlet. Since no sludge recirculation takes place, only the surplus biomass has to be separated- a considerable advantage over the activated sludge process. The reactor is used for aerobic as well as anoxic treatment (see Fig. 2).

The original biofilm carrier (Kaldnes K1) and most of the newer, commercial carriers, see examples in Fig. 2) are made of high density polyethylene (density 0.95 $g \cdot cm^{-3}$). Hence they are easily totally mixed in the reactor volume by the mixing set up by aeration (in aerobic reactors) and mixers in anoxic reactors (see Fig. 2).

One of the important advantages of the MBBR is that the filling fraction of carrier in the reactor may be subject to preferences. At the maximum recommended filling fraction for the K1 and K3 carriers (bulk specific area 500 $m^2 \cdot m^{-3}$, see Fig. 2) in aerobic reactors of 65%, the specific area in the reactor will be 325 $m^2 \cdot m^{-3}$. For the K5 and



Examples of biofilm carriers

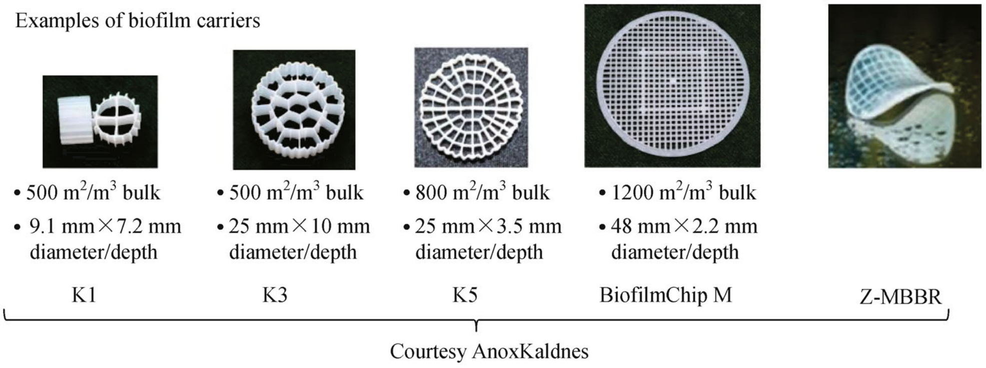


Fig. 2 Principle of the MBBR and examples of carriers

Biofilm chip carriers (see Fig. 2), the bulk specific area is higher, but the recommended maximum filling fraction is a bit lower (60% and 55%, respectively). One may, however, use as much as needed below the maximum filling fraction. This is very convenient when using the MBBR technology for upgrading of activated sludge plants.

The carriers are kept within the reactor by an outlet sieve. This may be vertically mounted, rectangular mesh sieves (mostly used in anoxic reactors), or a cylindrical bar sieve (mostly used in aerobic reactors), vertically or horizontally (most common) mounted. Even with a medium to coarse bubble diffuser system, the oxygen transfer in an MBBR is not inferior to that of a fine-bubble diffuser system in activated sludge. This is caused by the fact that the bubbles are partly caught by the moving carriers and partly broken in contact with the carriers. It has been demonstrated that the oxygen transfer increases with filling fraction up to approximately 60% where values above 15 $g\,O_2 \cdot Nm^{-3} \cdot m^{-1}{}_{\text{water depth}}$ have been measured in clean water tests with medium-sized bubble aerators [17].

The biofilm area is the key parameter in design and therefore the design rate of the process is most correctly based on effective carrier area as $gm^{-2}{}_{\text{carrier area}} \cdot d^{-1}$ [18].

5 Compact wastewater treatment plants of the future

5.1 Example flow diagrams

For the benefit of the discussion in this paper, two example flow diagrams are shown in Figs. 3 and 4. The first one (Fig. 3) is based on compact, established, proven technologies, notably:

- chemically enhanced primary treatment (CEPT) with dissolved air flotation for floc separation for BOD- and P-removal at low metal dose to ensure sufficient phosphate for downstream biological processes
- moving bed biofilm reactors for nitrification/denitrification (nitrogen removal) in combined pre-and post-denitrification mode
- micro-sand ballasted lamella separator for biomass separation and micro-sieve for suspended solids polishing
- ozonation for micro-pollutant oxidation and microbial disinfection
- microfiltration (in ceramic membranes) for final particle elimination
- thickening/thermal hydrolysis of sludge – all of which is taken out through the DAF
- anaerobic digestion for biogas production and residual sludge reduction
- de-ammonification based on MBBR-based IFAS for N-removal in sludge reject water
- dewatering and possibly incineration of residual sludge
- combined heat and power (CHP) facility for energy recovery

As mentioned above, a break-through in energy saving will take place when one is able to carry out nitrogen removal by de-ammonification also in the main-stream. This would change the proposed example diagram, primarily because a carbon source is not needed in de-ammonification and sludge production is much lower.

In Fig. 4, an example flow diagram based on de-ammonification in the main-stream is shown. Basically the flow diagram in Fig. 4 is the same as in Fig. 3, but with some important differences:

1) In the carbon removal stage (C-stage) a high rate MBBR is introduced in order to take out also the soluble organic matter – in addition to the particulate organic matter (as in Fig. 3)

2) De-ammonification (in two steps) is carried out in the main-stream (instead of nitrification/ denitrification) by the use of MBBR's and combined with a much less comprehensive polishing nitrification/denitrification step in order to be able to reach high N-removals

3) De-ammonification in the side-stream is carried out in a two-step MBBR and the main-stream de-ammonification step is bio-augmented by moving carriers from the side-stream nitritation and anammox steps respectively to the equivalent step in the main-stream

4) Micro-sieving is used (in combination with coagulation) for the biomass removal since biomass production from the bioreactor will be very low

In the following, each of the steps in these flow diagrams shall be discussed.

5.2 The carbon removal stage (C-stage)

5.2.1 The C-stage of the flow diagram based on compact, established, proven technologies (Fig. 3)

The C-stage in the flow diagram based on compact, established, proven technologies (Fig. 3) is based on coagulation, flocculation and floc separation by dissolved air flotation. Coagulation plants based on pretreatment, coagulant mixing, flocculation and floc separation are commonly used when discharging to marine waters in Norway and Table 2 shows the average results with respect to organic matter of these plants in two extensive studies [19,20]).

In the literature one may find experiences from plants based on so-called Chemically Enhanced Primary Treatment (CEPT) that do not show similarly high removals. This can be attributed to the fact that many of the CEPT plants are normally designed (and previously operated) as primary settling tanks and later upgraded to CEPT simply by adding a coagulant. The Norwegian plants are optimized for coagulation with proper mixing, coagulant control, flocculation and properly designed settling or

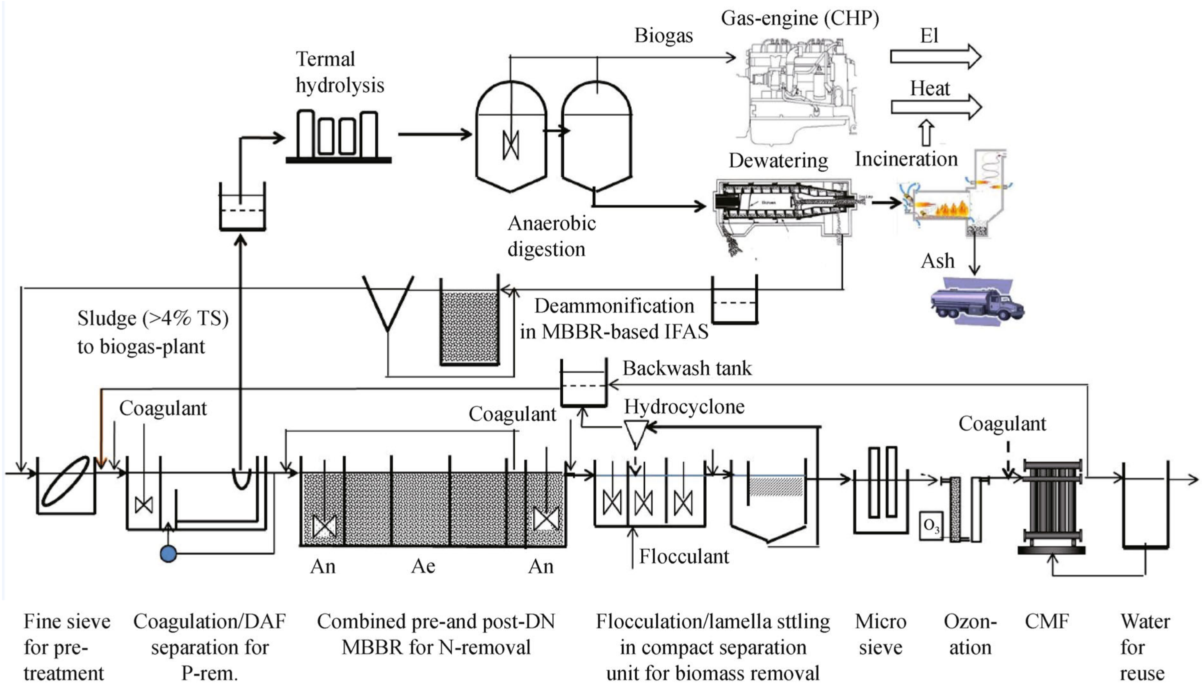


Fig. 3 Example flow diagram based on compact, established and proven technologies

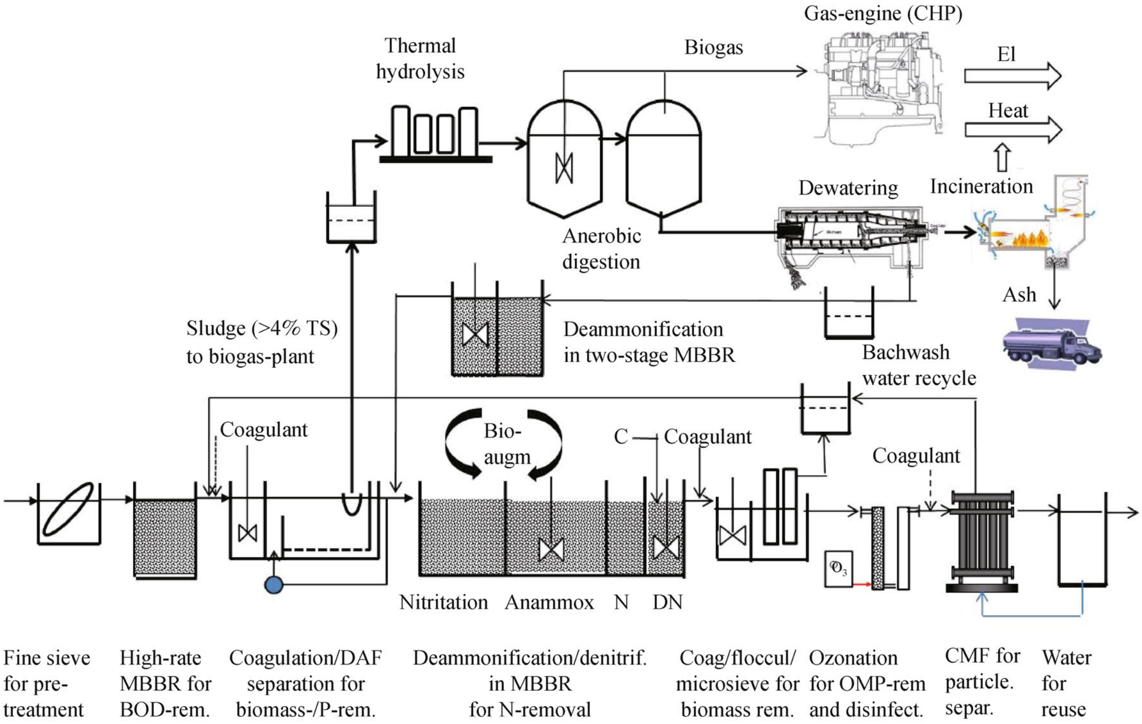


Fig. 4 Example flow diagram based on de-ammonification for nitrogen removal in the main-stream

flotation units for floc separation. If this is done properly, 70% removal of organic matter is achievable in most places–also in China – as the data in [9] indicates.

It is important to minimize inorganic coagulant (Fe or Al) dose in order to minimize the precipitation of metal hydroxide that does not contribute to biogas production. Therefore a combination of cationic polymer plus a small dose of iron may be chosen [10].

Table 2 Removal of organic matter in Norwegian coagulation plants (average values)

	N	*n*	COD_{in} mg·L^{-1}	COD_{out} mg·L^{-1}	COD_{rem} %	BODin mg·L^{-1}	BOD_{out} mg·L^{-1}	$BOD_{rem.}$ %	Ref.
1990 investigation	87	531_{COD} 183_{BOD}	418	99	73,4	167	27	80,9	Ødegaard[19]
2002 investigation	88	778_{COD} 787_{BOD}	366	90	75,5	135	33	75,7	Nedland [20]

Note: *N* - Number of plants, *n* – number of samples

The residual, soluble organic matter is easily biodegradable and is utilized in the pre-denitrification step of the following combined pre- and post-denitrification MBBR. The NO_3-N removal capacity of this soluble organic matter will be in the order of 15–20 mg NO_3-N·L^{-1} in normal wastewater.

Dissolved air flotation (DAF) is an energy consuming process (0.01–0.03 kWh·$m^{-3}{}_{treated\ water}$). A less energy consuming compact floc separation process (lamella settler, floc blanket settler, micro-sand ballasted lamella settler etc.) maybe used instead. DAF was chosen here because it concentrates the recycled and dilute backwash water from separation processes later in the process train. This back wash water has a certain coagulant content that may reduce coagulant dose ahead of the DAF and concentration of the sludge taken out of the DAF will be high (>4% DS).

5.2.2 The C-stage of the flow diagram based on de-ammonification in the main stream (Fig. 4)

Since de-ammonification does not require a carbon source, as do nitrification/denitrification, it will be beneficial to remove both particulate and soluble organic matter ahead of de-ammonification and this may efficiently be carried out by a high rate carbon removal stage (C-stage). For the C-stage, high rate activated sludge (HRAS) have been used [4]. However, an activated sludge step requires a spacious settling tank. An alternative is to use a high rate MBBR-process [11,12] based on:

- a highly loaded MBBR (typically < 30 min HRT) followed directly by
- coagulation by cationic polymer plus a small dose of inorganic coagulant (like iron)
- floc and biomass separation through compact flocculation/dissolved air flotation (DAF)

The MBBR should be so designed with high organic surface area loading rate, so that the biomass will only degrade the soluble, most easily biodegradable organic matter, while the particulate, more heavily biodegradable organic matter will pass the MBBR more or less unchanged and be picked up by the coagulant. The soluble organic matter is easily biodegradable and is quickly converted to biomass with a high biogas potential, as long as the loading is high. Particulate matter entering the MBBR and biomass produced by biodegradation of easily degradable soluble organic matter in the MBBR is separated from the water by the compact flocculator/DAF reactor in the example flow diagram in Fig. 4.

To avoid hydrolysis of particulate organic matter, the biodegradable, filtered organic load has to be higher than around 20–25 gBFCOD·$m^{-2}{}_{carrier\ surface}$·$d^{-1}$ (which in a normal wastewater is equivalent to 15–25 $gSBOD_5$·m^{-2}·d^{-1}), as demonstrated in Fig. 5 [21].

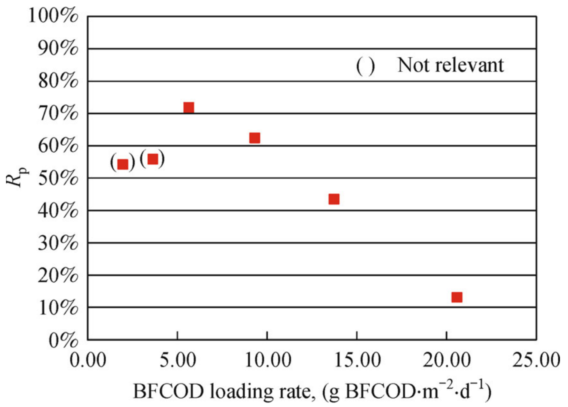


Fig. 5 Extent of hydrolysis (R_p, %) of particulate COD being hydrolyzed to soluble COD as a function of biodegradable, filtered COD (BFCOD) load [21]

The reaction rate at this high load will be at its maximum (and 0 order) of around 30 gBFCOD·m^{-2}·d^{-1}(20 $gFBOD_5$·m^{-2}·d^{-1}), as demonstrated in Fig. 6 [12]resulting in an MBBR effluent concentration of biodegradable filtered COD (BFCOD)>150–200 g·m^{-3}, equivalent to 100–150 $gFBOD_5$·m^{-3}. Below this concentration the reaction order will be close to half order, but even down to 100 gBSCOD·m^{-3} (or 70 gFBOD·m^{-3}), the removal rate will be as high as 20 gBFCOD·m^{-2}·d^{-1}(15 $gSBOD_5$·m^{-2}·d^{-1}) corresponding to around 60 g COD_{total}·m^{-2}·d^{-1} (40 $gBOD_{total}$·m^{-2}·d^{-1}). At this high load only a negligible fraction of the particulate COD will be converted to soluble COD.

A pilot plant study [12] demonstrated that secondary treatment standard (25 mg BOD in effluent) could be reached at a total HRT from inlet to outlet of around 1.0 h when operating on a quite concentrated municipal wastewater at relatively cold temperatures (10–15°C). Based on this study, the following design criteria may be proposed for the high-rate MBBR process:

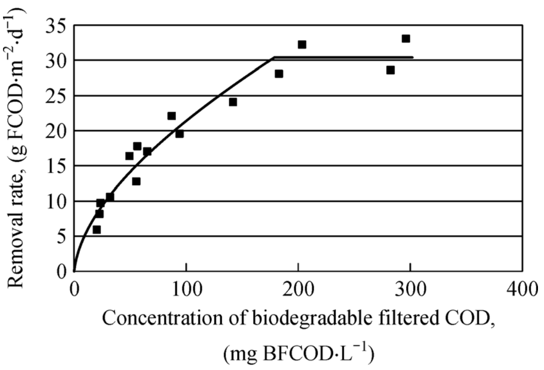


Fig. 6 The removal rate of filtered(1 μm filtered) COD versus the concentration of filtered biodegradable COD [12]

–Load: 20–25gCOD$_{filt}$ · m^{-2} · d^{-1}(15–20gBOD$_{filt}$ · m^{-2} · d^{-1}), 65–85gCOD$_{tot}$ · m^{-2} · d^{-1}(45–60 gBOD$_{tot}$ · m^{-2} · d^{-1})

– Sludge production: 0.5 gDS · gCOD$_{rem}$$^{-1}$ in MBBR step, 1.0 gDS · gSS$_{rem}$$^{-1}$ in biomass separation step

– Coagulant dosage: 5 g$_{cationic\ polymer}$ · kgSS^{-1} + 35 gFe · kgSS^{-1}

5.3 The nitrogen removal stage

5.3.1 Nitrogen removal in the main-stream of the flow diagram based on compact, established, proven, technologies (Fig. 3)

In the flow diagram based on established, proven compact technologies (Fig. 3), nitrogen removal is based on a combined pre- and post-denitrification process in compact moving bed biofilm reactors. Combined-denitrification MBBR plants may typically have a division between aerobic and anoxic compartments of the MBBR as indicated in Fig. 7.

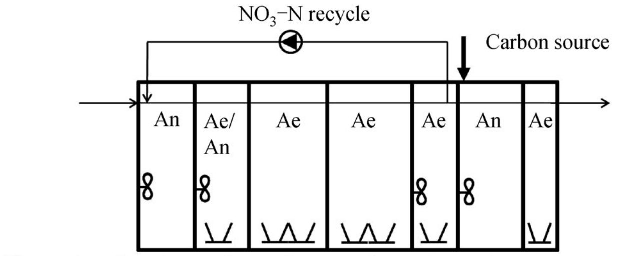


Fig. 7 Typical build-up of a combined pre- and post-denitrification MBBR [22]

The first compartment (comp. 1) is anoxic (pre-denitrification) and receives recycled nitrate from compartment 5. The second compartment (comp. 2) may be operated both anoxic (no air) and aerobic (with air). Then follow two aerobic reactors (comps. 3 and 4), used for residual organic matter removal and nitrification. Compartment 5 is aerobic, but may be operated with or without aeration, in the latter case functioning as a deoxygenation compartment. The sixth compartment, to which carbon is added, is anoxic and used for post denitrification. Finally there is optionally a small aerobic compartment for oxygenation as well as for removal of any residual biodegradable organic matter.

This bioreactor scheme offers great flexibility of operation and may give very high nitrogen removal efficiencies (Tot $N < 3$ mg · L^{-1}) [22]. The use of external carbon source is minimized. During summer, for instance, less nitrification volume is needed because of higher temperatures. In this situation compartment 2 may be used for denitrification (stirred and not aerated). More nitrate may be returned to the pre-denitrification tank and the need for added carbon source in the post-DN tank is lower.

In summer time, therefore, more of the nitrogen can be removed by pre-denitrification. During cold weather (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-denitrification tank in order to achieve the treatment goal. In compartment 5 the ammonia concentration will normally be low (< 3 mg · L^{-1}) and oxygen will no longer be rate limiting for the nitrification rate, but rather the ammonia concentration. Hence this compartment is used for de-oxygenation (no aeration), minimizing 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 (compartment 4) since there is a linear relationship between nitrification rate and oxygen concentration [23] [24]). The extent of denitrification in the post denitrification step may be controlled by the external carbon source addition [25].

The experiences with combined pre- and post-denitrification plants in Norway are very good. Surveys have demonstrated that < 3 mg Tot N can be achieved at total HRT of the bioreactor down to 3–5 h (depending on pre-treatment) at an energy consumption of around 0.25 kWh · m^{-3} and external carbon consumption of typically 1.6 g COD · m^{-3}, i.e. with more than half of the denitrification taking place in the pre-denitrification step [26].

5.3.2 Nitrogen removal by de-ammonification in the side-stream (reject water) of the flow diagram based on established, proven, compact technologies (Fig. 4)

The MBBR has demonstrated itself to be a robust and compact biofilm technology for de-ammonification in sludge reject water [27,28,29]).

There are presently 2 commercial processes available for MBBR-based de-ammonification: DeAmmon® - developed by Purac/Läckeby AB (Kalmar, Sweden) in collaboration with the University of Hannover (Hannover, Germany) and the Ruhr River Association (Ruhrverband; Essen, Germany) and ANITA™Mox- developed by AnoxKaldnes AB (Lund, Sweden) – a Veolia subsidiary.

The first full scale MBBR demonstration plant for treatment of reject water was built in Hattingen in Germany [27] as a multi-stage process. The demonstration project proved the MBBR de-ammonification process here to be cost-effective in comparison to conventional N-removal [28].

ANITA™ Mox is a 1-stage MBBR de-ammonification process [29]. Partial nitrification to nitrite and autotrophic N-removal (i.e. anammox) occur simultaneously within the biofilm, where aerobic and anoxic zones results from oxygen mass transfer limitation under limited dissolved oxygen (DO) conditions (see Fig. 8). The ammonium oxidizing bacteria (AOB) as well as the anammox 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 [29] [30].

AOB oxidize NH_4 to NO_2 in the aerobic zone of the biofilm (i.e. outer part) while anammox bacteria located in the anoxic zone of the biofilm (i.e. inner part) consume NO_2 produced by AOB together with the excess NH_4 (see Fig. 8).

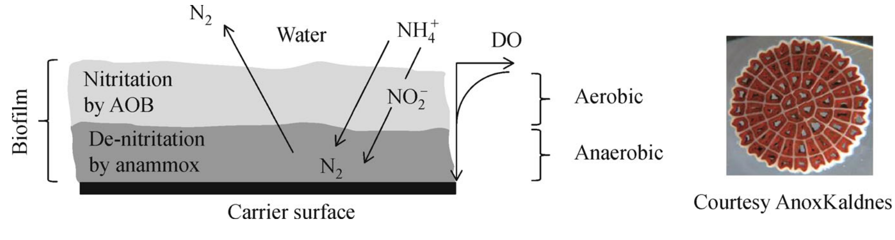
An advanced DO control system is used to prevent further oxidation of NO_2 into NO_3 by the nitrite oxidizing bacteria (NOB) in the aerobic zone of the biofilm maximizing the amount of nitrite available for the anammox bacteria. The DO set-point is automatically adjusted based on online inlet and outlet concentrations of NH_4 and NO_3 to control the NO_3 production below 11% of NH_4 removed (i.e. stoichiometric NO_3 production by anammox) while keeping high NH_4 oxidation performance in the reactor. This real-time DO control strategy reduces the need of mechanical mixer in the MBBR due to the continuous aeration pattern [29].

To shorten the start-up phase, new ANITA™Mox 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. To meet the request for seeding carriers, the first ANITA™Mox full-scale plant built in 2010 at Sjölunda WWTP, Sweden, is used as a nursery for anammox bacteria growing on suspended carriers, to be sold to customers [29].

When comparing 2-stage MBBR and 1-stage MBBR on sludge reject water, excellent nitritation in the 2-stage system was found, but also that it was more difficult to control the anammox stage [31].Nitrite is a substrate for anammox bacteria but nitrite accumulation can also inhibit their activity [32,33].One argument in favor of a 1-stage process is, therefore, that nitrite production inhibition of anammox bacteria at high nitrite concentration is less problematic. Inhibition was experienced in a 2-stage process pilot-plant, while nitrites produced by AOB were consumed immediately by anammox in the 1-stage process [34]. To prevent anammox inhibition in a 2-stage process the NO_2-N/ NH_4-N ratio in the reactor effluent (influent to anammox reactor) should be kept as close as possible to 1.32 [34].

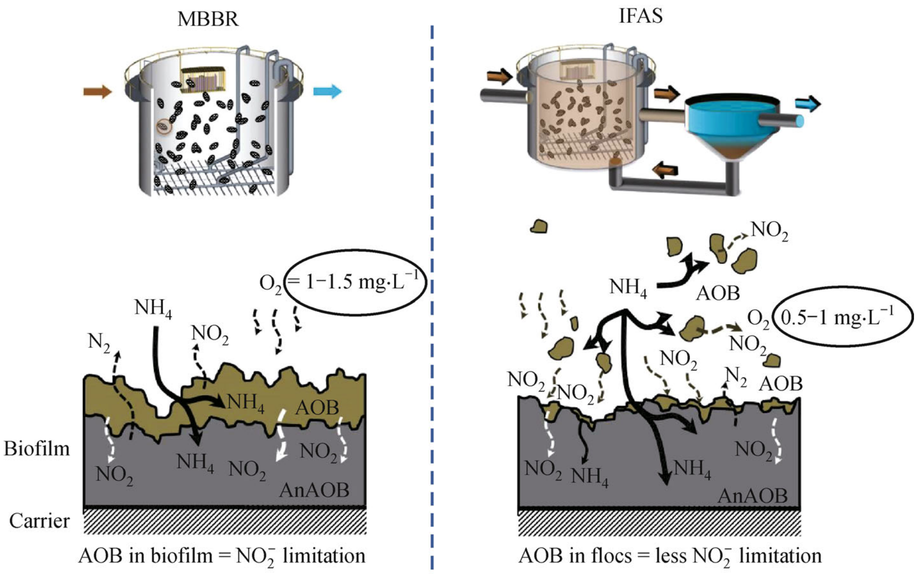
In an attempt to overcome the inhibition challenge and to improve the ANITA™Mox process performances under different operating conditions, it was hypothesized [35] that substrate transport could be enhanced by combining suspended cultures and fixed biomass into one system, so-called Integrated Fixed-Film Activated Sludge (IFAS).

In laboratory-scale on anaerobic sludge digester reject water, it was demonstrated that the nitrogen removal capacity was almost 4 times higher when comparing the pure 1-stage MBBR version with the IFAS 1-stage version [35]. With reference to Fig. 9 the authors offered the following explanation: “In an IFAS system the NO_2 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, NH_4) 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



Courtesy AnoxKaldnes

Fig. 8 Schematic of 1-stage nitritation/anammox biological processes occurring inside a carrier’s biofilm and anAnox™K5 carrier colonized with anammox bacteria



Courtesy Veolia

Fig. 9 MBBR versus MBBR-based IFAS for de-ammonification [35]

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 Fig. 10. The larger AOB population in IFAS configuration improves the overall flux of NO_2 produced for AnAOB but also the residual concentration in the bulk improving the diffusion of the NO_2 through the basal layer of the biofilm where AnAOB are located and therefore increasing their active fraction in the biofilm" [36]. In the experiments the global amount of AnAOB in the IFAS and the MBBR configuration was found to be similar, but the improvement of nitrites flux and production in the IFAS mode lead to a N-removal rate near the maximum values obtained during fully anoxic AnAOB activity batch test with non-limiting substrate levels (i.e. NH_4 and NO_2) [35].

Full-scale experiences with the ANITA^{TM}Mox process, both in pure biofilm as well as in IFAS mode, from the demonstration plant at Sjölunda WWTP in Sweden, that confirm the findings in the laboratory scale test, are described in [30] and [36].

5.3.3 Nitrogen removal in the flow diagram based on de-ammonification in the main-stream (Fig. 4)

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 by which to sustain de-ammonification in the colder and more dilute mainstream wastewater are yet to be developed. The challenges of mastering mainstream de-ammonification have been summarized as follows [37]:

- 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 mainstream wastewater.
- 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 IFAS system instead of pure MBBR may also be a possible way for NOB out-competition – as discussed above.
- An effective retention of the anammox biomass in a reactor is required. This is because inflow nitrogen concentration in mainstream wastewater is low (25–50 g NH_4^+-N· 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 the MBBR is especially suitable.
- Nitrogen transformation rates are about 70%–80% lower for mainstream wastewater with 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 rate can be expected.

At this time most developers and researchers work with MBBR-based IFAS systems for main-stream de-ammonification [38,39]. The Veolia companies, for instance, have worked with the two strategies [38], illustrated in Fig. 10.

The main move here is to utilize bio-augmentation to the less robust main-stream process from the more robust side-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 (Fig. 10(a)). This action will allow for partial regeneration of the anammox biofilm due to the more favorable conditions in the side-stream reactor (higher T°C, higher NH_4 level, lower COD level). The carrier transfer flow rate (Fig. 10(a)) will have to be controlled in order to maximize the bio-augmentation effect to the mainstream IFAS ANITA^{TM}Mox reactor [38].

Alternatively, another approach has been proposed [38]. This alternates the feeding between the mainstream COD-treated effluent and the side-stream reject water to a multi-celled IFAS ANITA^{TM}Mox system. As shown in Fig. 10 (b), 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 mainstream 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 mainstream IFAS ANITA^{TM}Mox studies were presented [38] - one pilot-scale study in Paris treating chemically enhanced primary effluent and one full-scale prototype (reactor volume 50 m^3) study at ANITA^{TM}Mox plant in Malmö, Sweden treating effluent from a high rate activated sludge (HRAS) step in the full-scale Sjölunda plant. In the prototype plant the bio-augmentation strategy using alternating feed between side-stream and mainstream effluent was tested. From these studies it was concluded that the IFAS ANITA^{TM}Mox process was successfully applied as a mainstream process [38]:

- The removal rates obtained in the mainstream IFAS ANITA Mox pilot-plant were on average 1.4 $gN \cdot m^{-2} \cdot d^{-1}$ during the summer (at 23°C) and 0.5–0.8 $gN \cdot m^{-2} \cdot d^{-1}$ 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 0.1–0.3 $kgN \cdot m^{-3} \cdot d^{-1}$(0.23–0.68$gN \cdot m^{-2} \cdot d^{-1}$) at 13–18°C. This N-removal performance is higher than what is commonly achieved in conventional BNR activated sludge systems.

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 optimize process conditions for exactly this stage. The single-stage MBBR seems to be inferior to the single-stage MBBR-based IFAS system for de-ammonification, but a two-stage MBBR-process, where nitritation and anammox is separated in two different stages, may be competitive in a main-stream process if an extended BOD-removal in the C-stage is secured and bio-augmentation from the side-stream to the main-stream is implemented.

A two-stage system with one or more continuously aerated MBBR(s) for nitritation and an anoxic MBBR for anammox was tested out in laboratory-scale [40]. 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 μm by the use of a special new saddle-shaped carrier (AnoxKaldnes Z-200 carriers, see Fig. 2), hence exposing a large fraction of the biomass to 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 NOB from the biofilm would hence increase. The thin biofilm would also ensure high oxygen availability, which was expected to improve nitritation rates. The feed was switched periodically from low-strength, low-temperature mainstream wastewater to reject water at high temperatures and concentrations (in agreement with the strategy shown in Fig. 10(b)). This sudden exposure to high substrate

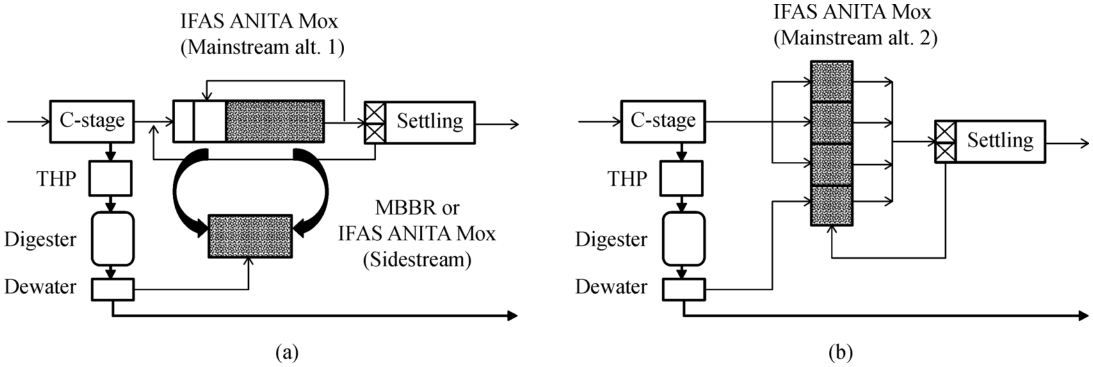


Fig. 10 Schematic of mainstream ANITA^{TM}Mox WWTP (a) with carrier recycling concept betweenside-stream and main-stream ANITA^{TM}Mox reactors and (b) with alternating feed concept betweenside-stream and main-stream water (after [38])

concentrations and temperatures was expected to inhibit NOB growth in the thin biofilm, and possibly also to boost AOB activity.

The laboratory-study indeed demonstrated 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 favor AOB activity and suppress NOB growth was successful [40]. In the nitritation stage, ammonia removal rates ranged between 0.25 and 0.50 $kgNH_4\text{-}N \cdot m^{-3} \cdot d^{-1}$ (1.25–2.5 $g \cdot m^{-2} \cdot d^{-1}$), with a nitrite accumulation ratio above 75%, and the effluent was fed to the second-stage anammox reactor, with nitrogen removal rates reaching 0.20 $kgNH_4\text{-}N \cdot m^{-3} \cdot d^{-1}$(1.0 $g \cdot m^{-2} \cdot d^{-1}$). At an average influent concentration of 51.6±5.2 $gNH_4\text{-}N \cdot m^{-3}$effluent concentrations below 15 $g \cdot TotN \cdot m^{-3}$ could be achieved. Further studies are underway to optimize the scheme and evaluate the fate of the NOB, but once fully applied this concept is expected to enable new solutions for mainstream anammox applications [40].

Based on all these findings, the strategy chosen for N-removal in the flow diagram based on de-ammonification in the main stream (Fig. 4) is:

- Good removal (>85%) of BOD up-front of the de-ammonification reactor (by the high rate MBBR process) in order to suppress heterotrophs in the de-ammonification processes
- De-ammonification in the main stream in two-stage MBBR (thin biofilm carrier) in order to be able to optimize the two parallel processes:
  - o Nitritation at proposed process conditions: DO = 1.5–2.0 $g \cdot m^{-3}$, $NH_4\text{-}N_{out}$: 4–5 $g \cdot m^{-3}$ (to maximize AerAOB, suppress NOB)
  - o Anammox at proposed process conditions: DO < 0.1 $g \cdot m^{-3}$, $NO_3\text{-}N_{out}$: 3–4 $g \cdot m^{-3}$ (to make AnAOB overrule NOB for NO_2)
- Use of de-ammonification also in side-stream (as MBBR-based IFAS) and bio-augmentation of the less robust main-stream process from the more robust side-stream process, by moving carriers back and forth between the two by the use of an air-lift pump (or similar).
- Use of a two-step nitrification/denitrification MBBR directly after the main-stream de-ammonification MBBR when high nitrogen removals are required (< 3 $gTotN \cdot m^{-3}$ in effluent). Post denitrification (with carbon source addition) will have to be used since there is very little $sCOD_{biodegradable}$ left in the recycled side-stream after anaerobic digestion and dewatering, and very little $sCOD_{biodegradable}$ left in the main-stream after the nitrification step.

5.3.4 Micro-pollutant removal in the MBBR steps

Organic micro-pollutants (OMP) will be removed, more or less, in all the treatment steps of the proposed flow diagrams; adsorbed to particles in the primary as well as the secondary and tertiary particle removal steps, biodegraded in the biological (MBBR) step and oxidized in the ozonation step.

Of particular interest is the potential OMP-removal in the MBBR-steps. Through batch experiments, the removal of seven active pharmaceutical substances (ibuprofen, ketoprofen, naproxen, diclofenac, clofibric acid, mefenamic acid, and gemfibrozil) was assessed, by the biomass on suspended carriers and in suspended sludge from several full-scalewastewater treatment plants (including activated sludge, MBBR-based IFAS and MBBR plants) [41]. A distinct difference between nitrifying activated sludge biomass and biomass on suspended biofilm carriersin removal of several pharmaceuticals was demonstrated. The biomass on biofilm 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, clofibric 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 [41].

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 [42].

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 activated sludge. The results from the model and plant measurements showed that the removal efficiency of the process could be predicted with acceptable accuracy (~25%) for most of the modeled micro-pollutants. The model estimations indicated that the attached growth in hybrid (IFAS) processes can contribute significantly to removal of individual pharmaceutical compounds [42].

5.4 The biomass separation step

When using MBBR one is free to use any biomass separation method (settling, lamella settling, dissolved air flotation, micro-sieving, sand filtration, membrane filtration etc.) since the SS to be separated is low (< 200 $gSS \cdot m^{-3}$) – while with activated sludge one has to use very space demanding settling or expensive membrane filtration because of the high MLSS (2000–8000 $gSS \cdot m^{-3}$). The options for biomass separation are described in [43].

5.4.1 Biomass separation in the flow diagram based on nitrification/denitrification in the main stream (Fig. 3)

In the flow diagram for the compact, proven, well established technologies (Fig. 3) one has chosen to apply the micro-sand ballasted lamella separator that has been used with success in several MBBR plants in Norway. Coagulants are used and the waste sludge flow, separated from the micro-sand in the hydro-cyclone, has a low SS-concentration and is returned to the CEPT stage and separated in the DAF-unit. Hence all the sludge is taken out here at a relatively high DS-concentration (>4% DS).

5.4.2 Biomass separation in flow diagram based on de-ammonification in the main stream (Fig. 4)

In the process of the flow diagram based on de-ammonification in the main stream, sludge production in the de-ammonification step is very low (< 30–40 gDS $\cdot m^{-3}$) and micro-sieves are chosen for biomass separation – possibly after polymer coagulation. The backwash water is returned to the high rate MBBR/coagulation stage (the C-stage) for separation of biomass together with the incoming SS. The residual coagulation capacity of the backwash water may be sufficient for acceptable coagulation by cationic polymer addition alone.

5.5 The water reclamation stage

A water reclamation step is included in both flow diagrams consisting of:

- Ozonation (for organic micro-pollutants removal and disinfection)
- Coagulation (for enhancement of colloidal matter separation – including microorganisms)
- Ceramic membrane filtration (for final colloidal matter separation)

The extent of treatment in the water reclamation step is, of course, dependent on what the water is to be reused for. In the flow diagrams presented here, bathing water quality is aimed at, for the reuse of water for agricultural, industrial and public reuse (such as toilet flushing, garden irrigation, car and street washing etc.).

5.5.1 Micro-pollutant removal in the water reclamation stage

Organic micro-pollutants (OMP) will be removed, more or less, in all the treatment steps of the proposed flow diagrams; adsorbed to particles in the primary as well as the secondary and tertiary particle removal steps, biodegraded in the biological (MBBR) step and oxidized in the ozonation step.

A thorough study of organic micro-pollutants (OMP) removal was carried out at EAWAG in Switzerland where the removal of selected OMP's was analyzed by the use of three methods; ozonation, powdered activated carbon adsorption and nanofiltration [44]. As demonstrated in Fig. 11 all these methods were able to remove OMP's more or less.

Taking efficiency and cost into consideration, it was decided to recommend ozonation as the method to use. Figure 12 shows full-scale results from the ozonation plant at the city of Regensdorf, Switzerland [45] [46]. As demonstrated a dose of 0.6–0.8 $gO_3 \cdot gDOC^{-1}$ was sufficient to significantly reduce (80%–100%) most of the selected micropollutants. At a typical concentration of 5 $gDOC \cdot m^{-3}$ in the tertiary treated water and a medium dosage of 0.6 $gO_3 \cdot gDOC^{-1}$, the electrical energy consumption would be 0.035 $kWh \cdot m^{-3}$ for the ozonation step, corresponding to 12% of the energy consumption at a typical medium-sized nutrient removal plant [46].

The EU-bathing water directive has a standard of 500 $cfuE.coli \cdot 100mL^{-1}$ and this level or better was achieved at a dosage >0.5 $gO_3 \cdot gDOC^{-1}$ at Regensdorf WWTP, which means that bathing water quality will be reached at the dosage needed for OMP removal [45].

5.5.2 Final particle separation by ceramic membrane filtration (CMF)

In the proposed flow diagrams ceramic microfiltration is used for final particle separation. The main reason behind

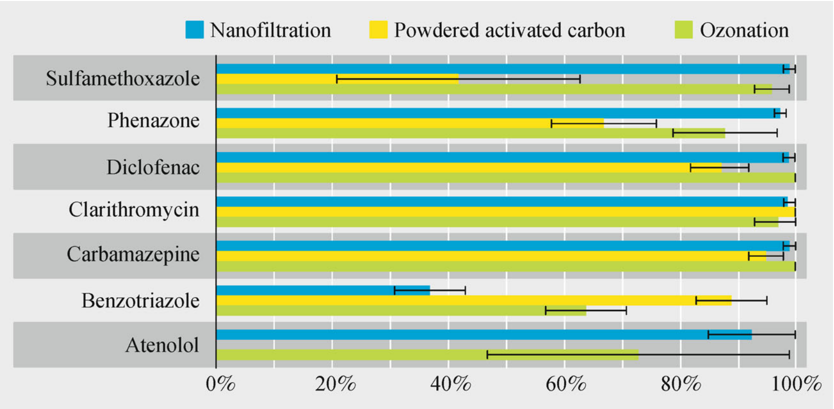


Fig. 11 Removal of organic micro-pollutants by three different methods in the EAWAG study [44]

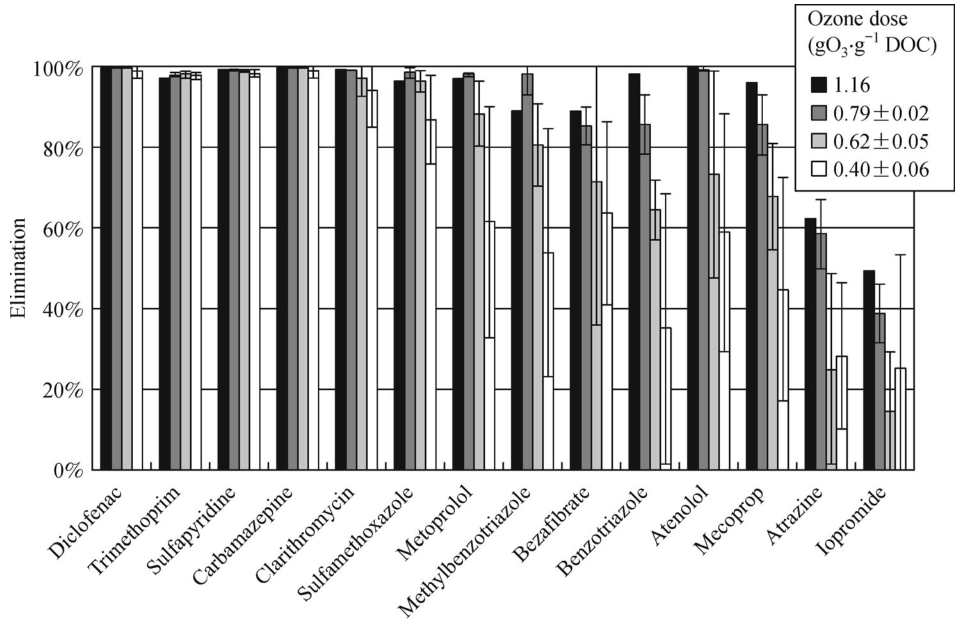
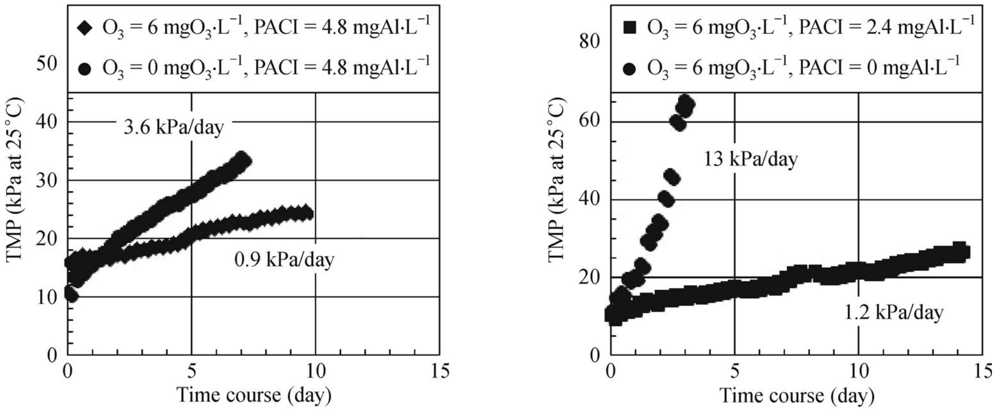


Fig. 12 Treatment results for the full-scale ozonation plant of the city of Regensdorf, Switzerland [46]

this choice is that ceramic membranes can accept pre-ozonated water. The residual ozone, which is already there for OMP- removal, will contribute to reduce fouling. There are several other good reasons to choose ceramic membranes such as; high permeability (flux>100 LMH), high water recovery (>98%), high mechanical strength, high stability for chemicals so that online CIP (cleaning in place) can be performed, well defined pore size distribution, accepts turbidity variation well and has relatively low operation cost. The invest cost is high but since the life-time of ceramic membranes can be expected to be much higher than for polymeric membranes the life-time cost is not higher than that of polymeric membranes [47].

Studies of the combined ozonation/CMF system at Shibaura water reclamation plant in Tokyo demonstrated (Fig. 13(a)) that at a constant coagulant dose, the TMP increase was much lower when ozone was applied – indicating the positive effect of pre-ozonation on fouling [48]. In Fig. 13(b) it is demonstrated that the use of coagulant also had a marked positive effect on TMP development at a given ozone dose –because coagulation of fine colloids reduces fouling [48].

To study the potential for the use of ozone as an anti-fouling agent in the membrane plant, the residual ozone concentration (in the water entering the ceramic membranes) was raised from zero to 0.12 $g \cdot m^{-3}$ over a period. This resulted in an immediate reduction (ca 20%) in TMP, demonstrating in full-scale the positive effect of residual ozone in fouling prevention [48]. At Shibaura WWTP, that has been in stable operation for more than 5 years, the cost of water reclamation in ozonation/ceramic membrane filtration has been demonstrated to be 0.49 US$\cdot m^{-3}$ [48]



a. The influence of ozone dose on TMP

b. The influence of coagulant dose on TMP

Fig. 13 The influence of ozone dose and coagulant dose on TMP-development in CMF [48]

5.6 The sludge treatment stage

5.6.1 Thermal sludge hydrolysis (THP) combined with anaerobic digestion for biogas production

Thermal hydrolysis of sludge is an efficient way to enhance biogas production. In one of the commercial processes the sludge is thickened/pre-dewatered to 15%–18% DS, then fed to the so-called "pulper tank" in the thermal hydrolysis (THP) pre-treatment system. Thereafter the sludge is transferred to the THP reactor where it is batch-wise pressure cooked at 150–170°C and 6–8 bar pressure for 20–30 min, before flashed into the flash tank, with temperature dropping to around 102–107°C. The sludge is heat-exchanged to 38–42°C before it is fed to the anaerobic digester at 8%–12% DS. The typical digester HRT is 15 days, resulting in a VS-reduction of 50%–65% while this would be 30%–45% without thermal hydrolysis. The high temperature secures complete elimination of pathogens and THP pretreatment improves dewaterability of digested sludge (30%–40% sludge cake DS depending on dewatering equipment) [49].

The biogas is utilized for combined heat and power (CHP) production or/and biogas boiler, with waste heat from CHP for boilers as well. In the Davyhulme WWTP (UK) the digester load could be increased by a factor of 2.4 (to about 4 $\mathrm{kgVS \cdot m^{-3} \cdot d^{-1}}$) and the CHP out-put increased by around 30% after THP was introduced. So altogether the same digester volume gives around 3 times more biogas after THP installation [50].

6 Conclusions

In this paper two possible concepts for a future compact wastewater treatment plant have been discussed:

1. One based on established, proven treatment technologies (i.e. nitrification/denitrification for N-removal in the main-stream) (Fig. 3).
2. One based on emerging technologies (i.e. de-ammonification in the main-stream) (Fig. 4).

The latter will give an energy-neutral wastewater treatment plant, while this cannot be guaranteed for the first one. The example flow diagrams show plant concepts that:

- Minimize energy consumption by:

o using compact biological and physical/chemical processes combined in an optimal way, for instance by using moving bed biofilm reactor (MBBR) processes for biodegradation and high-rate particle separation processes

o de-ammonification processes for N-removal in side-stream as well as in main-stream

- Maximize energy (biogas) production through digestion by using:

o wastewater treatment processes that minimize biodegradation prior to digestion

o pretreatment of the sludge prior to digestion by thermal hydrolysis

The treatment plant of the future should produce a water quality (for instance bathing water quality) that is sufficient for reuse of some kind (toilet flushing, urban use, irrigation etc.). The paper outlines compact water reclamation processes based on ozonation in combination with coagulation as pretreatment before ceramic membrane filtration.

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Short biography—Hallvard Ødegaard

Professor Hallvard Ødegaard received his MSc in Civil Engineering at Norwegian Institute of Technology(NTH) in 1969 and his PhD (Dr.ing.) of Environmental Engineering at the same university in 1975.

Since 1977 he has been professor of Water and Wastewater Treatment at Department of Hydraulic and Environmental Engineering, Norwegian University of Science and Technology (NTNU) in Trondheim, Norway. In 2011 he retired from this position and is now Professor Emeritus at NTNU.

He is also doing consulting work through his consulting company Scandinavian Environmental Technology AS. For instance he was independent expert reviewer for vetting the assessments and recommendations about sewage and sludge treatment technologies and processes for the relocated ShaTin Sewage Treatment Works in Hong Kong.

Prof. Ødegaard was visiting professor at EAWAG, Switzerland in 1991, at Hokkaido University, Sapporo, Japan in 1999 and at CSIRO, Molecular Science, Melbourne; Australia in 2000.

He has written more than 530 papers out of which more than 280 refereed papers published internationally.

Prof. Ødegaard has 40 years of experience in work on treatment of drinking water, wastewater and industrial process water. His specialities are : Particle separation (including membrane) processes; Biofilm processes; Disinfection processes; Removal of humic substance in drinking water treatment; Nutrient removal in water and wastewater treatment.

He is the inventor of the moving bed biofilm reactor (MBBR) and holds 5 international patents.

Professor has been appointed IWA Distinguished Fellow and has received the highest honour in Norway: Knight of 1. Class of The Royal Norwegian St. Olav Order