Desalination and Water Treatment
www.deswater.com
1944-3994/1944-3986 © 2012 Desalination Publications. All rights reserved
doi: 10/5004/dwt.2012.2461
42 (2012) 80–86
April

Compact tertiary treatment based on the combination of MBBR and contained hollow fibre UF-membranes

Hallvard Ødegaard[a,*], Ulrich Mende[b], Eirik Olsen Skjerping[c], Simon Simonsen[c], Rune Strube[d], Erik Bundgaard[e]

[a]*Norwegian University of Science and Technology (NTNU), N-7491 Trondheim, Norway*
Tel. +47 73594759; Fax: +47 73591298; email: hallvard.odegaard@ntnu.no
[b]*Aquantis GmbH, Lise-Meitner-Str. 4a, 40878 Ratingen, Germany*
[c]*KrugerKaldnes AS, Hegnasletta 11, 3217 Sandefjord, Norway*
[d]*Hydrotech AB, Mejselsgatan 6, 235 32 Vellinge, Sweden*
[e]*Krüger AS, Gladsaxevej 363, 2860 Søborg, Denmark*

Received 20 October 2010; Accepted 6 July 2011

ABSTRACT

The paper describes a tertiary treatment method (intended for water reuse) based on the moving bed biofilm reactor (MBBR™), followed by a high-rate biomass separation step such as disc filter, DAF or Actiflo and followed by UF. Results from an experimental period carried out in pilot scale at the Gardermoen WWTP in Norway are described. Focus was on the use of the Hydrotech disc filter in combination with a contained, hollow fiber, outside-in ultrafiltration membrane unit. Different coagulation scenarios were tested and compared to the one based on the effluent from the full scale plant that uses coagulation/flocculation and DAF after the MBBR. It is concluded that the MBBR – high rate separation – UF process (the TERFLEX® process) offers an interesting and competitive alternative to the activated sludge based MBR. The best result when the UF membrane was used after primary biomass separation by DAF. The use of disc filter as the high rate biomass process ahead of the UF offers a very compact solution.

Keywords: Wastewater treatment; Moving bed biofilm reactor (MBBR); DAF; Microscreening; Disc filter; Ultrafiltration

1. Introduction

Membrane bioreactors (MBRs) based on activated sludge are often preferred for compact water reuse plants. There has been an increasing focus, however, on the use of biofilm reactors, especially the moving bed biofilm reactors (MBBR), in combination with membrane separation [1–5].

MBBR's have several advantages over conventional activated sludge plants [6]:

a. The treatment plant requires less space (an important cost factor).
b. The final treatment result is less influenced by biomass separation since the biomass concentration to be separated is at least 10 times lower and there is greater flexibility in choice of biomass separation method (i.e., compact settling, flotation or filtration).
c. The attached biomass becomes more specialized (higher concentration of relevant organisms) at a given point in the process train, because there is no biomass return.

*Corresponding author.

Contrary to most biofilm reactors, the MBBR utilizes the whole tank volume for biomass growth, as does also the activated sludge reactor. Contrary to the conventional activated sludge reactor, however, it does not need any sludge recycle. 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 [6].

MBR plants based on activated sludge avoid the sludge settling problems of conventional plants and high biomass concentrations (typically 8–12 g MLSS l^{-1}) can be used without compromising biomass separation significantly. Even so, the MBBR is more compact than the MBR when nitrification/nitrogen removal is to be achieved. The MBBR-based MBR is therefore an interesting concept.

The combination of MBBR and membrane (UF) separation may be carried out in different ways. The membrane unit may be placed directly after the MBBR as an immersed membrane (as in most MBRs) or as a separate, contained membrane unit. The studies on MBBR with UF separation reported so far have been based on immersed membranes [1–5]. 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 has been proposed in order to minimize this particle caused fouling [7]. Ivanovic et al demonstrated the beneficial effect of designing the membrane reactor in such a way that sludge is continuously taken out of the membrane reactor [2]. Melin et al demonstrated that coagulation of the MBBR biomass prior to an immersed UF-membrane had a beneficial effect [5]. The fouling rate 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 l m^{-2} h^{-1} without coagulation and 25 l m^{-2} h^{-1} with coagulation.

The alternative to immerged membranes would the use of contained membranes. One would, however, be skeptical to the use of contained membranes directly after the MBBR since the biomass concentration introduced to the contained hollow fiber after all will be in the range of 200–300 mg SS l^{-1}. It was, therefore, thought useful to have an intermediate biomass separation step between the MBBR and the membrane unit. Such a process solution would probably be very competitive with traditional MBRs if a high-rate biomass separation method is used to lower the MLSS concentration entering the membrane reactor. The principle of this process solution is shown in Fig. 1.

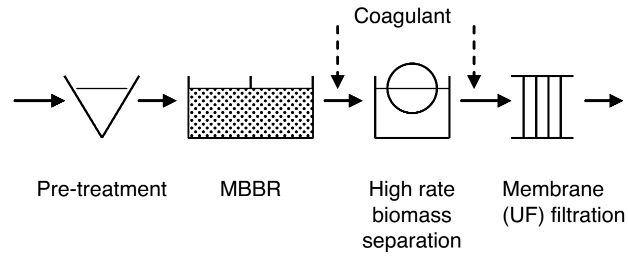

Fig. 1. The principle of the wastewater treatment process proposed.

The high-rate biomass separation step may consist of either one of the separation reactors:

- A microsieve reactor (for instance the discfilter (DF))
- A dissolved air flotation (DAF) unit
- A high-rate settling unit (for instance the Actiflo settling reactor)

This process combination (with either of the biomass separation reactors) will result in a very compact treatment solution. The four Veolia-companies (KrügerKaldnes and AnoxKaldnes, representing the MBBR technology, Hydrotech, representing the Hydrotech Disc filter technology and Aquantis, representing UF-technologies) decided, therefore, to test the proposed process at Gardermoen wastewater treatment plant outside Oslo. The focus of the test was on the combination of MBBR, Discfilter and UF (contained, outside-in, dead-end, hollow fibre). Since the plant at Gardermoen use flotation (after coagulation) for biomass separation, the tests also included the MBBR–DAF–UF combination.

2. Methods and materials

2.1. The full scale plant

The full scale MBBR plant (Gardermoen WWTP) has an average flow of around 1000 m^3 h^{-1}. Primary settling is followed by a combined pre- and post denitrification MBBR plant with a total MBBR residence time at average flow of around 6 h. The MBBR is followed by a coagulation/flocculation step (ca. 20 min HRT) with a dosage of 15 mg Al l^{-1} + 0.25 mg l^{-1} anionic polymer (as flocculant) before the DAF units at a surface load of around 5 m h^{-1} at average flow. The plant is producing a good effluent quality, typically <5 mg BOD_5 l^{-1}, <25 mg COD l^{-1}, 7 mg TN l^{-1} and <0.2 mg TP l^{-1} [8]. Fig. 2 shows the flow sheet of the Gardermoen plant as well as from where the water for the pilot was taken.

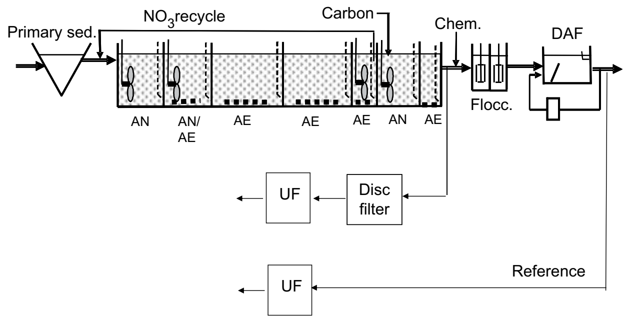

Fig. 2. The test venue – the Gardermoen WWTP.

2.2. The discfilter (DF) – unit

The pilot-plant consisted of a coagulation/flocculation unit followed by a Hydrotech Discfilter (Fig. 3(a)) and a UF unit (Fig. 3(b)). The micro-screen (disc filter) was a Hydrotech HSF 1702/1-1F. The function of the disc filter and the use of it for biomass separation after MBBR are described in Ref. [9]. The filtration cloth had a pore opening of 40 µm. The pilot disc filter had four modules, each consisting of six panels giving a total filter cloth area of 2.8 m^2. The rotational speed of of 2.8 rpm corresponds to a peripheral velocity of 0.3 m sec^{-1}. Backwash was initiated at a differential pressure of 250 mm.

Ahead of the discfilter an in-line mixer for coagulant and two stage flocculation tank (each 1,000 l) with slow mixers were placed. For Al-dosing PAC 18 (SNF, Nordic) containing 9% Al and for cationic polymer dosing mainly K5060 (Kemira) was used. Dosages ranged from 3–5 mg Al l^{-1} and 1.5–4 mg polymer l^{-1}. Dosage optimization was carried out prior to and during the pilot experiments.

2.3. The UF unit

The Aquantis AQ07 tertiary pilot (*see* Fig. 3(b)) is a contained, dead-end, outside-in, hollow-fibre ultrafiltration unit, suitable for processing water with relatively high SS. The pilot plant consists of the UF membrane module, a backwash unit, a CIP unit, a feed pump and a dosing unit. The plant was controlled by a computer equipped with a SCADA system, that monitored and managed the operation of the membrane unit and the water flows through the plant. The membrane unit, operated at constant flux, used a DOW OMEXELL™ UF hollow fibre PVDF membrane (module SPF 2860) with an asymmetric dense spongy layer and skins formed on both sides of the fibre. The nominal pore size of the membrane is 0.03 µm. The housing was 1860 mm long with a diameter of 225 mm. The fibres has an inner diameter of 0.65 mm and an outer diameter of 1.25 mm. The total area for filtration was 52 m^2.

Permeate was used to backwash at a rate of 5.2 m^3 h^{-1} at a pressure of less than 3 bar. The chemicals during chemically enhanced backwash were H_2SO_4, NaOCl and NaOH. Backwashing (without chemicals) took place every 20 min and every 12 (or 24) backwash was a chemically enhanced on (called cyclic CEB). In the cyclic CEB either NaOCl (experimental phases 1–4) or NaOH (experimental phases 5–7) was used. In addition there was a more comprehensive daily chemically enhanced maintenance backwash, performed at noon every day. Unfortunately, the acid dosing was not carried out optimally in experimental phases 1–4 and not at all during phases 5–7, due to technical failures.

2.4. Processes investigated

A number of experiments were carried out in experimental phases with different process combinations. Coagulation was implemented before or after the high-rate biomass unit. When using DAF, coagulation/flocculation was implemented before the DAF (in the full-scale plant) and no chemicals were used after the DAF and before the UF. When using DF, coagulation by the use of prepolymerized aluminium (PAC) was used alone or in combination with polymer ahead of the DF or after the DF and before the UF.

a.The disc filter plant b.The UF plant

Fig. 3. The pilot plant.

Table 1
Overview of the experimental phases and fouling rates recorded

Phase	DO[a]	Process description	Flux rate ($l\ m^{-2}\ h^{-1}$)	Turb (FNU) before UF	PF ($mbar\ min^{-1}$)	STF ($mbar\ d^{-1}$)	LTF ($mbar\ d^{-1}$)
1.0	5	MBBR–C/F–DAF (full scale)–UF	48	5	2.3	38.7	17.4
2.1	2	MBBR–C/F 3 Al[b]–DF–UF (short run)	35	30	21.0	30.6	14.9
2.2	4	MBBR–C/F 5 Al –DF–UF	35	26	14.4	56.7	13.2
3.1	3	MBBR–DF (no pre-coag)–UF (short run)	35	95	38.4	78.0	32.8
3.3	7	MBBR–DF–C 5 Al–UF	35	54	22.6	42.3	–17.8
4.2	4	MBBR–C 5Al + 1.5 cat.pol[c]–DF–UF	35	20	24.7	31.6	15.9
5.1	25	MBBR–DF–C 5 AL–UF	43	76	21.8	36.7	4.7
5.2	14	MBBR–C 3 cat.pol. –DF–C 5 Al–UF	43	17	19.9	30.10	21.8
6.1	14	MBBR–C/F–DAF (full scale plant) –UF	48	2	1.4	15.7	16.4
6.2	40	MBBR–C/F–DAF (full scale plant) –UF	45–80	14	Frequent	Flux	Changes
7.0	4	MBBR–4 cat.pol –DF–C 3 Al–UF	43	16	51.3	127.0	37.50

[a]DO – Days of operation.
[b]C/F 3 Al – coagulation/flocculation with 3 mg Al l^{-1} dose (C-coagulation only).
[c]C 5Al + 5 cat.pol. – coagulation (no flocculation tank) with 5 mg Al l^{-1} plus 5 mg cationic polymer added.

Table 1 summarizes the different process combinations as well as the flux rates of the UF and the average turbidity (over the experimental phase) in the water entering the UF.

The various process combinations studied in the various experimental phases were evaluated based on fouling rate, backwash water consumption and chemicals consumption – for coagulation as well as for membrane cleaning.

In Fig. 4 is a definition sketch related to the various fouling rates defined. Three different fouling parameters were used:

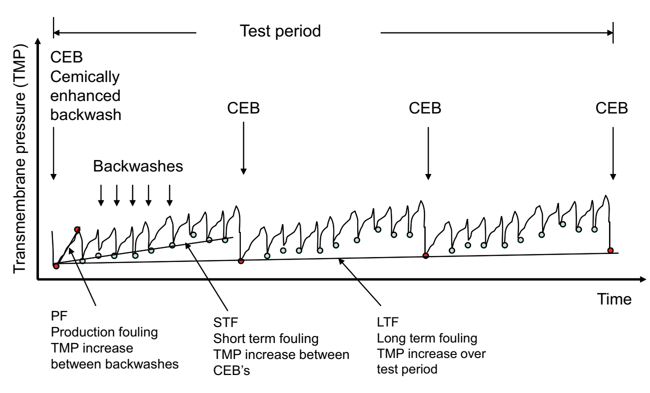

Fig. 4. Definition sketch for the different fouling rates registered (for illustration only).

a. PF – production fouling – the TMP increase during a filter cycle (between backwashes). Backwash was normally initiated after 20 min filtration time or at TMP increase of 350 mbar. PF indicates the degree of cake fouling.
b. STF – short-term fouling – the TMP increase between each chemically enhanced backwash CEB. The STF value gives information about the fouling control that can be achieved by backwashing (without chemicals).
c. LTF – long-term fouling – the TMP increase over the test period. The LTF value gives information on the

effect of fouling control by the chemically enhanced backwashing, CEB (CEB procedures – *see* Section 2.3).

3. Results and discussion

3.1. Fouling

In Table 1 are given the values of PF, STF and LTF as they were determined. One cannot draw too firm conclusions from the absolute figures since small differences in operational conditions as well as experimental phase duration can have a significant impact on the calculated number. The experimental phase describes the experiment, lasting for the given time, under which the actual process combination was operated.

We may summarize the influence of the various process combinations as follows:

a. The best results (least fouling) were achieved when using the DAF solution.(MBBR + coag/flocc + DAF + UF). At 35 l m^{-2} h^{-1} the fouling was hardly noticeable (short period – not shown in Table 2). At 48 l m^{-2} h^{-1} (phase 1 and 6.1) in the main test period it was a bit higher, but still very low. In phase 6.2 various membrane cleaning strategies and high fluxes (45–80 l m^{-2} h^{-1} were tested and the fouling rates not registered.
b. The use of disc filter without any chemical coagulation/flocculation (phase 3.1) did not seem to be any good option. Fouling rates of the UF were high and backwashing of disc filter was also high (low recovery of DF).
c. The results are not quite clear with respect to coagulation/flocculation before or after the disc filter. Fouling was less at 5 mg Al l^{-1} than at 3 mg Al l^{-1} added and suspension flocculated before the DF (phase 2.1 and 2.2), but fouling was in both cases relatively high as a consequence of inadequate biomass SS-separation in the disc filter, that normally requires use of a polymer for good separation.
d. The test period with disc filter without pre-coagulation but post-coagulation (prior to UF) (phase 3.3)

Table 2
The amount of backwash and consumption of membrane cleaning chemicals

Phase	DO	Process description	Flux (l m^{-2} h^{-1})	Cyclic CEB		Daily CEB		Total backwash of filtered water (%)
				NaOCl (37%) (ml m^{-3} m^{-2})[a]	NaOH (45%) (ml m^{-3} m^{-2})[a]	H_2SO_4 (15%) (ml m^{-3} m^{-2})[a]	NaOH(45%) (ml m^{-3} m^{-2})[a]	
1.0	5	MBBR–C/F –DAF (full scale)–UF	48	0.15	0	0.22	0.31	18.5
2.2	4	MBBR–C/F 5Al–DF–UF	35	0.22	0	0.36	0.50	25.5
3.1	3	MBBR–DF (no coag)–UF (short run)	35	0.43	0	0.39	0.53	53.6
3.3	7	MBBR–DF– C 5 Al–UF	35	0.27	0	0.33	0.46	32.0
4.2	4	MBBR–C 5Al + 1.5 cat. pol–DF–UF	35	0.27	0	0.38	0.53	35.4
5.1	25	MBBR–DF– C 5Al–UF	43	0	1.31	000[b]	0.30	22.3
5.2	14	MBBR–C 3 cat.pol.–DF– C 5 Al–UF	43	0	0.51	000	0.33	21.2
6.1	14	MBBR–C/F– DAF (full scale plant)–UF	48	0	1.53	000	0.23	18.3
7.0	4	MBBR–4 cat. pol–DF–C 3 Al–UF	43	0	0	000	0.38	50.9

[a]ml m^{-3} of filtered water per m^2 of membrane area.
[b]Out of order.

resulted in relatively high production fouling and short term fouling (because of the high turbidity before the UF) – but very low long term fouling. In fact it was negative in phase 3.3 – indicating that the CEB was able to get rid of fouling that had been established in earlier test periods. This result (low long-term fouling) was repeated when the same process was run for a relatively long period (25 d) at a higher flux (43 l m^{-2} h^{-1}) in phase 5.1.

e. Since one was afraid of polymer fouling of the membrane, the coagulation with aluminum plus cationic polymer was carried out late in the experiments (phase 4.2) in order to prevent damage to the membrane. Fouling of the membrane was higher (not dramatically higher, however) with 5 mg Al l^{-1} + 1.5 mg cationic polymer l^{-1} than with 5 mg Al l^{-1} alone – when the coagulant was added before the disc filter – even though polymer use gave better disc filter separation.
f. In order to try to optimize both DF separation and UF separation, it was decided therefore, to run two experimental phases with pre-coagulation by cationic polymer addition and post coagulation by aluminum addition (phase 5.2: low polymer and high Al dose, phase 7: high polymer and low Al dose). Even though DF-separation was improved by the polymer dose, fouling of the membrane was not much improved and was very high in the case of high polymer dose.

The period of chemical cleaning optimization (experimental period 6.2) was chosen to be carried out with the MBBR + coagulation/flocculation + DAF (full scale) + UF configuration and with an increased flux (45–80 l m^{-2} h^{-1}). It was shown that with this process configuration, the production fouling could be held within reasonable limits (<30 mbar min^{-1}) even up to a flux of 80 l m^{-2} h^{-1}. The max design flux recommended was, however, set at 50 l m^{-2} h^{-1}.

3.2. Membrane cleaning

The backwash also plays a role in deciding which process alternative that is to be favored—both the amount of backwash water used and the amount of chemicals used in the chemically enhanced backwash. When using the DF/UF alternatives, there will be backwash water both from the DF and the UF while in the DAF/UF alternative, there is only backwash from the UF. The backwash water for the DF is taken from the disc-filtered water while the backwash water for the UF is taken from the final effluent. Both these backwash waters will have to be returned to the plant inlet and therefore they play a role in the whole design picture.

The amount of backwash in the DF for this application is 2%–4%, the higher the solids load, the higher the backwash percentage.

In Table 2 is summarised the total amount of membrane backwash (in % of incoming flow or produced water flow) as well as the amounts of membrane cleaning chemicals used. It is emphasized that different cleaning procedures were tested and therefore one should be careful when comparing the numbers.

The lowest amount of backwash (around 18% of amount of filtered water) was experienced when water from the full scale plant (MBBR–coagulation/flocculation–DAF–UF) was used. The optimal strategy (from a fouling point of view) when using DF (MBBR–DF–coagulation–UF) resulted in a total backwash of around 22% of filtered water, while the use of DF without any coagulation at all resulted in a very high backwash amount (more than 50% of filtered water.

The use of two-step coagulation, cationic polymer ahead of DF and Al-coagulation after (but before UF) gave in phase 5.2 about the same backwash water amount (around 22% of filtered water) as the one with coagulation after the DF only, but less chemicals for cyclic CEB since more SS was removed in the DF. But this coagulation combination is risky because slight overdosing of polymer (as in experimental phase 7) resulted in extensive membrane fouling and consequently a high backwash need.

Also with respect to chemicals consumption for the cleaning of the membranes, the process solution based on DAF gave the most favorable results. It seems that the use of NaOCl for cyclic CEB is more favorable than the use of NaOH.

4. Conclusions

Based on these pilot tests, the following conclusions may be drawn:

1. The combination of MBBR followed by a high rate biomass separation step and a contained hollow-fiber UF step as final treatment (the TERFLEX® process), results in a very compact tertiary treatment plant, very competitive with the activated sludge MBR.
2. The results from the pilot experiments documented that a solution based on MBBR – coagulation/flocculation–DAF–UF was the one with the lower fouling, the higher possible flux, lower backwash water consumption and the lesser membrane cleaning chemicals consumption. This conclusion is expected also to be valid if DAF is replaced by Actiflo.
3. The use of a microscreen (40 µm disc filter, DF) for the primary biomass separation ahead of the UF gives the most compact solution. To be optimized it seems, however, that cationic polymer has to be used (either alone or in combination with a metal salt like aluminum) ahead of the DF. This process combination is risky,

however, because of the risk of polymer fouling of the membrane which requires thorough polymer dosing control.

4. An interesting process solution, that should be investigated further, is the one with controlled polymer dosing ahead of the DF and aluminum dosing after the DF and before the UF. The hypothesis is that the hydroxide precipitation caused by the aluminum precipitation sweeps possible polymer residuals and hence reduces the risk of membrane fouling.

References

[1] T. Leiknes and H. Ødegaard, The development of a biofilm membrane bioreactor, Desalination, 202 (2006) 135–143.

[2] L. Ivanovic, T. Leiknes and H Ødegaard, Influence of loading rates on production and characteristics of retentate from a biofilm membrane bioreactor (BF-MBR), Desalination, 199 (2006) 490–492.

[3] L. Ivanovic, T. Leiknes and H. Ødegaard, Fouling control by reduction of submicron particles in a BF-MBR with an integrated flocculation zone in the membrane reactor, Sep. Sci. Technol., 43(7) (2008) 1871–1883.

[4] E. Melin, T. Leiknes and H. Ødegaard, Kaldnes membrane bioreactor—Pilot test of MBBR in combination with Zenon Zee Weed membrane reactor. SINTEF report No STF22 F04717 (restricted) (2004).

[5] E. Melin and H. Ødegaard Pilot tests of Kaldnes membrane bioreactor with Zenon ZeeWeed membranes. Phase 2. SINTEF report No SBF IN F07303 (restricted) (2007).

[6] H. Ødegaard, Innovations in wastewater treatment: the moving bed biofilm process, Water Sci. Technol., 53(9) (2006) 17–33.

[7] R.M. Åhl, T. Leiknes and H. Ødegaard, Tracking particle size distributions in a moving bed biofilm membrane reactor for treatment of municipal wastewater, Water Sci. Technol., 53(7) (2006) 33–42.

[8] H. Ødegaard, Applications of the MBBR processes for nutrient removal. Proceedings IWA 2nd Specialized Conference in Nutrient Management in Wastewater Treatment Processes, pp. 749–760, Krakow, Poland, 6–9 September, 2009.

[9] E. Persson, M. Ljunggren, J. laCour Jansen, R. Strube and L. Jönsson, Disc filtration for separation of flocs from a moving bed bio-film reactor, Water Sci. Technol., 53(9) (2006) 139–147.