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Operation of MBR membrane modules used in a decentralised wastewater treatment plant: Field study and comparison of different cleaning strategies

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ABSTRACT

Due to their compact design and their high quality and reliable treatment, package or containerised membrane bioreactor (MBR) units are used for decentralised and semi-decentralised wastewater treatment plants. The operational availability, performance and economical viability of these MBR systems depend on the filtration performance of the membrane modules. Current chemical cleaning strategies of MBR modules, based on regular (weekly) maintenance cleanings and/or occasional (quarterly to biannual) intensive cleanings proved not to be adapted to semi-central MBR applications (100 up to 1000 p.e.): regular maintenance cleanings require automation and lead to too much care and personnel requirement. Occasional intensive cleanings increase the operational risk of membrane fouling and low cleaning recovery. In addition, semi-central MBR applications are often designed with at least two redundant filtration lines. An alternative chemical cleaning strategy was therefore proposed, implemented and assessed in a containerised MBR unit serving a population of about 250 p.e.: at a given time, only one filtration line is in operation while the other one soaks in a low-grade chemical solution. The modules are switched alternately on a monthly basis. To identify a cleaning strategy and an agent showing a good recovery, one of the modules was cleaned with H_2O_2, while the other was cleaned with NaOCl. A cleaning step with citric acid is added when necessary. These cleanings were tested over 16 months with the goal to minimise maintenance effort and chemicals used.

Keywords: MBR; Cleaning strategies; Hydrogen peroxide; Decentralised WWTPs

1. Introduction

The high-effluent quality of MBR systems combined with the small footprint and the compact design is an advantage of MBRs compared to conventional decentralised wastewater treatment plants. On the one hand these advantages can only be achieved because of the membrane filtration, but on the others hand the membrane system is a critical unit, as the whole operation relies on the functional availability of the module. Membrane fouling can be considered as major operational risk of small MBR plants. Fouling monitoring and control is therefore a key issue for designers as well as operators. The chemical cleaning of membranes is used to recover the permeability and increase the lifetime of the module. The identification of a cleaning strategy that allows easy planning of infrequent actions will help minimising contingent costs for additional travel to and labour on site.

Current chemical cleaning strategies of MBR modules, based on regular (weekly) maintenance cleanings and/or occasional (quarterly to biannual) intensive cleanings proved not to be adapted to semi-central MBR applications (100 up to 1000 p.e.): regular maintenance cleanings require automation and lead to too much care and personnel requirement. Occasional intensive cleanings increase the operational risk of membrane fouling

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and low cleaning recovery. In addition, semi-central MBR applications are often designed with at least two redundant filtration lines. The advantage of high-operational reliability due to redundancy compensates the higher investment costs. Design criteria for this demonstration plant are further presented in Gnirss et al. 2007a [1]. An alternative chemical cleaning strategy was therefore proposed, implemented and assessed in a containerised MBR unit serving a population of about 250 p.e.: at a given time, only one filtration line is in operation while the other one soaks in a low-grade chemical solution. The modules are switched alternately on a monthly basis. The novel cleaning strategy enables therefore to limit the operational risk due to fouling while renouncing to frequent and automised maintenance cleaning and optimising the staff presence on site.

Various types of cleaning agents are known to address different causes of membrane fouling. Within this study two oxidants dealing mainly with natural organic matter (NOM) are compared in terms of efficiency and costs: NaOCl and H_2O_2. The intention is to find a viable alternative to NaOCl which is not well tolerated in Germany.

2. Materials and methods

2.1. Plant description

The investigated membrane bioreactor (MBR) is operated by the *Berliner Wasserbetriebe* (BWB) in cooperation with the *Berlin Centre of competence for water* (KWB) in a peripheral area of Berlin. About 250 people living in the settlement of Margaretenhöhe are connected to the MBR which has been operated within the demonstration project *Enhanced Nutrient REmoval in Membranebioreactor* (ENREM). To achieve low effluent concentrations of carbon, nitrogen and phosphorus, this plant is operated with a process scheme that combines biological phosphorus removal with post-denitrification in order to achieve >90% TN-elimination and >99% TP-elimination. In order to warranty the maximum volume for the biological reaction before filtration, a design constraint was to keep the volume of the filtration reactor smaller than 10% of the entire mixed liquor volume (Gnirss et al., 2008 [2]). Fig. 1 shows a flow sheet of the demonstration plant.

A low pressure sewer system connects each house with the buffer tank that has a total volume of 10 m^3 and therefore gives the possibility to run the plant with a constant throughput. Because of a higher wastewater discharge during the weekend, a higher throughput is needed. Table 1 gives the ranges of the key operational parameters.

The wastewater is entirely domestic and does not contain any industrial source. Due to the newly built low-pressure sewer, the wastewater is virtually devoid of rain and stormwater.

Table 2 shows the influent and effluent concentrations of the chemical oxygen demand (COD), nitrogen

Table 1
Key operational parameters.

	Throughput in m^3/d	SRT in d	HRT in h	MLSS (in MR) in g/l	VSS (in MR) in g/l
Min.	8	18	14.6	10.2	7.4
Max.	13	45	23.9	25.4	20.0
Median	10.2	25	18.7	14.1	10.6

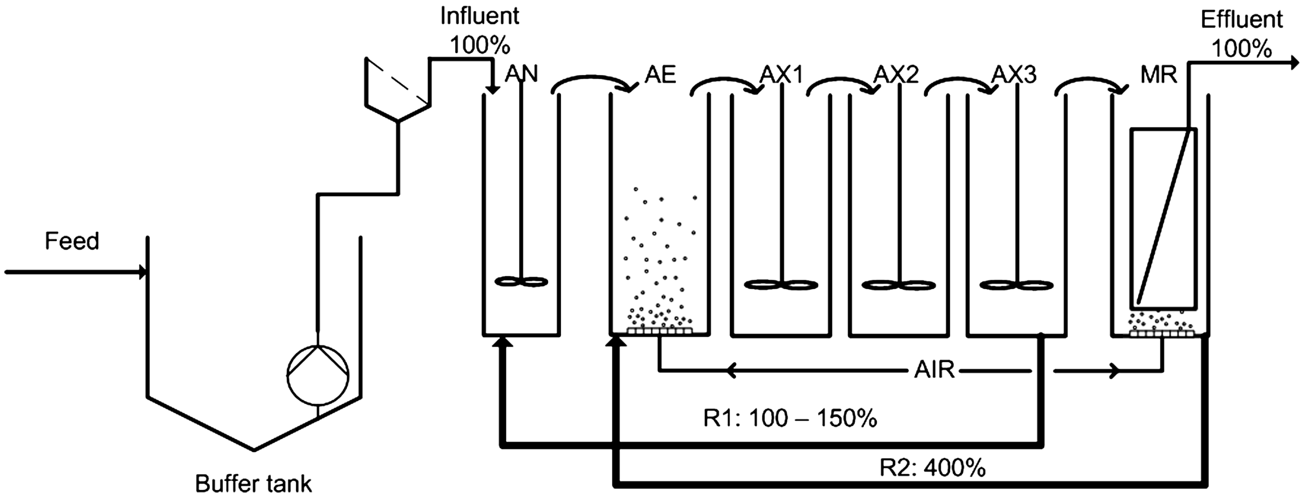

Fig. 1. Flow sheet of the MBR system.

Table 2
Influent and effluent concentration.

	COD in mg/l			Nitrogen in mg/l			Phosphorus in mg/l		
	In	Out	%	In	Out	%	In	Out	%
Min.	513	35		74,7	2.36		10.5	0.15	
Max.	2165	61		220	36.9		25.9	14.0	
Median	1150	44	96	149.5	14.4	90.3	19.4	1.4	94.1

and phosphorus as well as the achieved elimination for the period reported in this article.

2.2. Membrane modules

The plant is equipped with two parallel filtration reactors, each equipped with two filtration modules and an autonomous filtration system. The two modules, manufactured by A3 Water Solutions (Germany) are assembled one upon another which leads to a reduced footprint of the membrane reactor. The height of the two modules plus aeration system is about 2.2 m. Table 3 shows the key characteristics of the modules.

Continuous air scour during operation lies between 0.7 and 1.0 $Nm^3/(h * m^2)$. In order to reduce the operational costs, the scour was reduced to ~ 0.4 $Nm^3/(h * m^2)$ for several weeks, but increased fouling showed the necessity of higher values (see results and discussion). The instantaneous flux is about 15 $L/(m^2 * h)$ (weekdays) and 20 $L/(m^2 * h)$ (weekends). Filtration takes place using a pulse/pause regime with 16.6 minutes of filtration followed by 2.4 mins of pause during which a full relaxation is achieved.

Table 3
Key parameters of the used membrane module provided by A3 Water Solutions (2 per filtration reactor).

Reference	MX-020	
Surface area	15.9 m^2	
Material	PVDF	
Pore size	0.20 μm	
	Recommended by A3	As operated
Operational pressure difference	20–400 mbar	< 100 mbar
Instantaneous flux	15–25 L/ (m^2 * h)	15–20 L/ (m^2 * h)
pH during cleaning	2–11	2–11
pH during operation	5–9.5	7.2–7.6
Temperature range	1–50 °C	10–27 °C
Cleaning agents	Base; oxidant; tenside, acid	H_2O_2, NaOCl, citric acid
Cleaning interval	3–12 months high grade	Monthly medium grade

2.3. Cleaning strategy

As described above, two of these double deck modules are assembled within the demonstration plant in two independent filtration reactors. Only one of these modules is in operation at a time, and the other is soaking in the cleaning solution, with a switch ap prox. every month. Therefore, two different cleaning agents could be compared for the soaking solution:

1. H_2O_2 used for module 1 with a concentration of 1000 ppm
2. NaOCl used for module 2 with a concentration of 500 ppm chlorine

For both agents the following cleaning procedure was applied:

1. After withdrawal of the activated sludge the module is rinsed with permeate. Strong aeration is exercised in order to detach remaining sludge. To increase this effect and to reach surface areas assumed to be less affected by aeration e.g. corners and edges, the membrane pockets are filled with permeate up to a maximum pressure of 50 mbar. Afterwards the permeate is withdrawn and the membrane reactor is filled with tab water.
2. The chemical agent is added to the membrane reactor. When cleaning with H_2O_2 the pH was increased to approximately 11 adding sodium hydroxide.
3. Filtration and aeration is performed for 15 min to ensure a well mixed reactor.
4. The module is soaked within the cleaning agent for 1 month until the next cycle of operation is started, with disappearance of the chemical agent with time.

Occasionally (every 3-4 cleanings), an additional cleaning step with citric acid (5000 ppm, 1h) was performed before and after the main cleaning to attack the inorganic fouling. It has to be pointed out, that both chemical agents have to be of high quality, e.g. stored adequately, as a wrong storage leads to reduced concentrations.

2.4. Calculation of permeability

In order to compare the efficiency of the tested cleaning strategies the permeability was calculated and recorded online. The values for instantaneous flux and trans-membrane-pressure (TMP) are recorded as 2 h mean values. Only values recorded during filtration after flow stabilisation are used to calculate the mean values, therefore the net flux and the permeability can be calculated according to Trussell et al. (2005) [3]:

$$J = \frac{Q_{Permeate}}{A_{Membrane}} \qquad (1)$$

where J = membrane flux (L/m^2 * h); $Q_{Permeate}$ = membrane permeate flow (L/h); and $A_{Membrane}$ = membrane surface (m^2).

The permeability L can be calculated through

$$L = \frac{J}{TMP} \qquad (2)$$

where TMP = trans-membrane-pressure.(bar).

The permeability has been normalised to 20° C as described by Trussel et al. (2005) [3] using the following equation:

$$L^{20^\circ C} = L * e^{(-0{,}0239\ (T-20))} \qquad (3)$$

where T = water temperature (°C).

In addition clean water flux tests showed a pressure loss in the permeate system and before the pressure sensor due to turbulent flow, as the Reynold number (Re) lies above 7500 for the applied fluxes.

The pressure loss due to a turbulent flow should be considered for the calculation of the permeability. Equation (4) shows the theoretical equation to determine the pressure loss in cylinder pipes:

$$\Delta p = \zeta * \frac{\rho}{2} * w^2 * \frac{L}{d} \qquad (4)$$

where Δp = pressure difference (Pa); ζ = drag coefficient; ρ = density (kg/m^3); w = velocity (m/s); L = pipe length (m); d = diameter (m).

As the drag coefficient is constant for high Re values and the ratio of length to diameter does not change within the installation, the following pressure correction as a function of the flux and density can be identified for our system:

$$\Delta p_{headloss} = 2222 * \rho * J^2 \qquad (5)$$

where Δp = pressure difference (Pa); ρ = density (kg/m^3); J = membrane flux (L/m^2 h).

This equation has been determined experimentally by recording the evolution of the TMP during a clean water test for different fluxes as shown in Fig. 2.

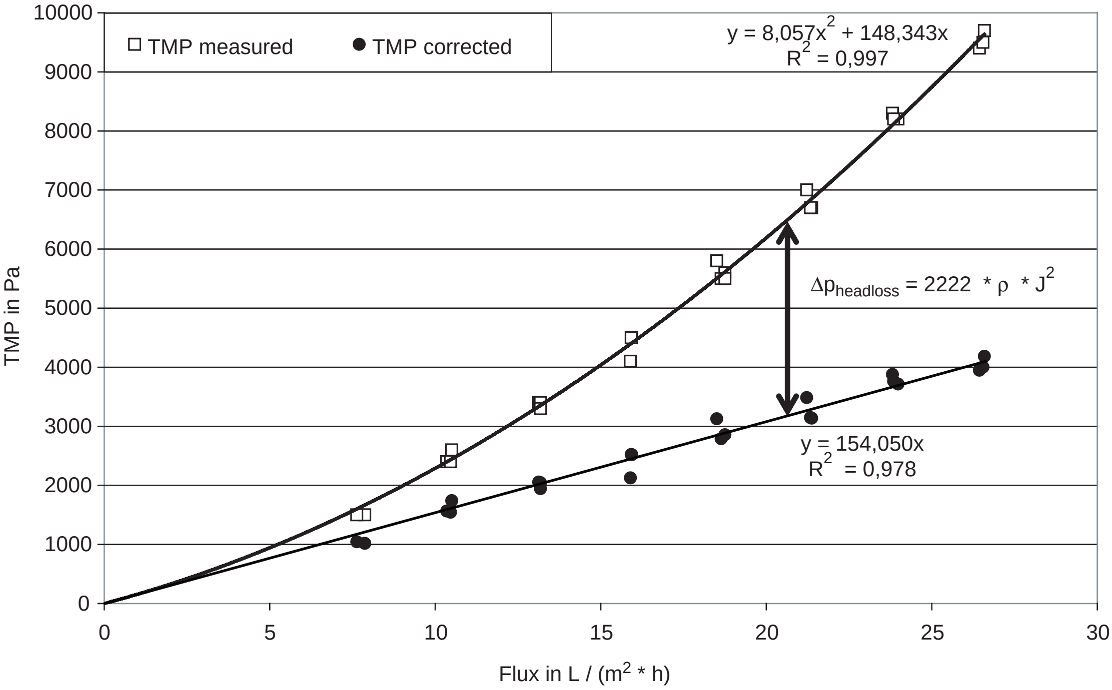

Fig. 2. Head loss determination.

Table 4
Sludge characteristics during the reported time.

	SMP Polysaccharides in mg/l	SMP Proteins in mg/l	Transparent exopolymer particles in mg/l	Bound Polysaccharides in mg/l	Bound Proteins in mg/l	COD Supernatant in mg/l
Samples	31	32	32	31	31	34
Min.	4	20	25	31	194	86.6
Max.	55	139	107	161	899	847
Median	25	38	58	124	613	267.5

The pressure difference has therefore been reduced by $\Delta p_{headloss}$ using Eq. (6) and used for the calculation with Eq. (2) of the normalised permeability, i.e. corrected for temperature and flux:

$$TMP = \Delta p_{sensor} - \Delta p_{headloss} \tag{6}$$

The recorded pressure difference caused by the head loss can be as high as 60% of the total pressure difference. This shows the necessity to take the head loss into account when calculating the permeability for systems operating in the range of a turbulent flow.

2.5. Determination of cleaning efficiency

The efficiency of each cleaning has been distinguished using the permeability values calculated during operation. As described above, each module was in operation for approximately one month. During this time a decrease of the permeability could usually be monitored and the efficiency of the cleaning can calculated using following equation:

$$R = \frac{L_i - L_{i-1}}{L_0 - L_{i-1}} * 100 \tag{7}$$

where R = recovery in percentage; L_0 = clean membrane in sludge at the beginning of operation: ~2,200 L/(m^2 h bar) for module 1 and ~ 1,600 L/(m^2 h bar) for module 2; L_i = permeability after cleaning; L_{i-1} = permeability at the end of the filtration run (before cleaning).

The average value of two days at the start and the end of the filtration run was used to calculate the recovery rate.

2.6. Additional measurements

Further measurements were carried out in order to characterise the investigated system as well as to identify influences of the biological performance and wastewater composition. A fast and easy way to determine the filterability of activated sludge is to measure the time needed to collect a specific amount of filtrate through a paper filter. In the present study, TTF was measured as the time to filter in dead end filtration 25 ml out of 250 ml activated sludge through a white ribbon filter paper (Whatman, pore size between 12 and 25 μm). The COD of the collected filtrate was measured using Hach-Lange test kits.

In situ critical flux measurements were regularly performed with the installed modules according to the flux-step method described by de la Torre et al. (2008a) [4]. Polysaccharides concentrations were analysed every week according to Dubois et al. (1956) [5]. The bound EPS were extracted using the cation ion exchange method described by Frølund et al. (1996) [6]. Proteins were analysed according to Frølund et al. (1995) [7]. Transparent exopolymer particles (TEP) were measured as described by De la Torre et al. (2008b) [8].

3. Results and discussion

3.1. Sludge characterisation

The measurements to characterise the filterability of the activated sludge were performed on a weekly basis from October 2007 up to August 2008. Table 5 shows the range of some of the measured parameters.

These relatively high values can be explained by the high influent concentrations due to the low pressure sewer system (no rain) and the limited water consumption per capita. Despite the high occurrence of the monitored potential foulants, no significant correlation with the filtration performance could be observed. As these parameters did not show a rapid change during a period of heavy fouling in November 2007, it was assumed that other components in the influent wastewater caused the decline of the permeability. Time to filter (TTF) measurements showed few peaks during the campaign, but this could not be related to fouling events either.

Therefore the filterability of the sludge was assumed to be relatively constant with the exception of a period of about five weeks in November 2007 that will be discussed

Table 5
Costs of cleaning detergents and chemicals.

	NaOCl + Citric acid		H_2O_2 + Citric acid + NaOH	
	per m^3 in €	% of operational costs[a]	per m^3 in €	% of operational costs[a]
10 m^3/d	0.047	1.16 %	0.086	2.15
15 m^3/d	0.031		0.057	

[a]operational costs w/o personnel costs: 2.80 – 4.00 €/m^3.

later. This enables to relate the cleaning effects, e.g. recovery, directly to the performed cleaning protocols.

3.2. Permeability evolution

Figure 3 shows the evolution of the permeability for both modules throughout the investigation. Module 1 was cleaned with H_2O_2 shown in blue, whereas module 2 was cleaned with NaOCl displayed in red.

The membrane modules were operated with an air scour between 0.7 and 1.0 $Nm^3/(h * m^2)$ except of the period between June and August 2007 as indicated. The reduced air scour was approximately 0.4 $Nm^3/(h * m^2)$. This reduction was expected to minimise energy costs for aeration, but the permeability of both modules decreased quickly. The permeability of module 1 decreased to an amount that was not tolerable therefore the air scour was increased again. Both modules recovered afterwards showing the strong influence of the air scour on the performance and discarding the possibility that inappropriate sludge characteristics led to the decline of permeability.

Another event of heavy fouling is related to a reduction of the sludge filterability in November 2007. While Module 2 was decommissioned during this period of fouling and the permeability could be recovered using citric acid before and after the cleaning step with a higher grade of NaOCl, Module 1 was operated within this sludge for about 4 weeks. This caused a dramatic decrease of the permeability of this module, which could not be recovered only with the soaking with the planned H_2O_2 (1000 ppm) cleaning. That is why an intensive chemical cleaning was required. Using a higher grade of NaOCl (2000 ppm at pH 11) and citric acid (2000 ppm at pH 2) while increasing the contact time to 48 h for NaOCl and 24 h for citric acid finally recovered the permeability.

The recovered permeability was even higher than the initial one. This demonstrates that the intensive cleaning is a good possibility to recover a strongly fouled module, but might change the membrane characteristic in a

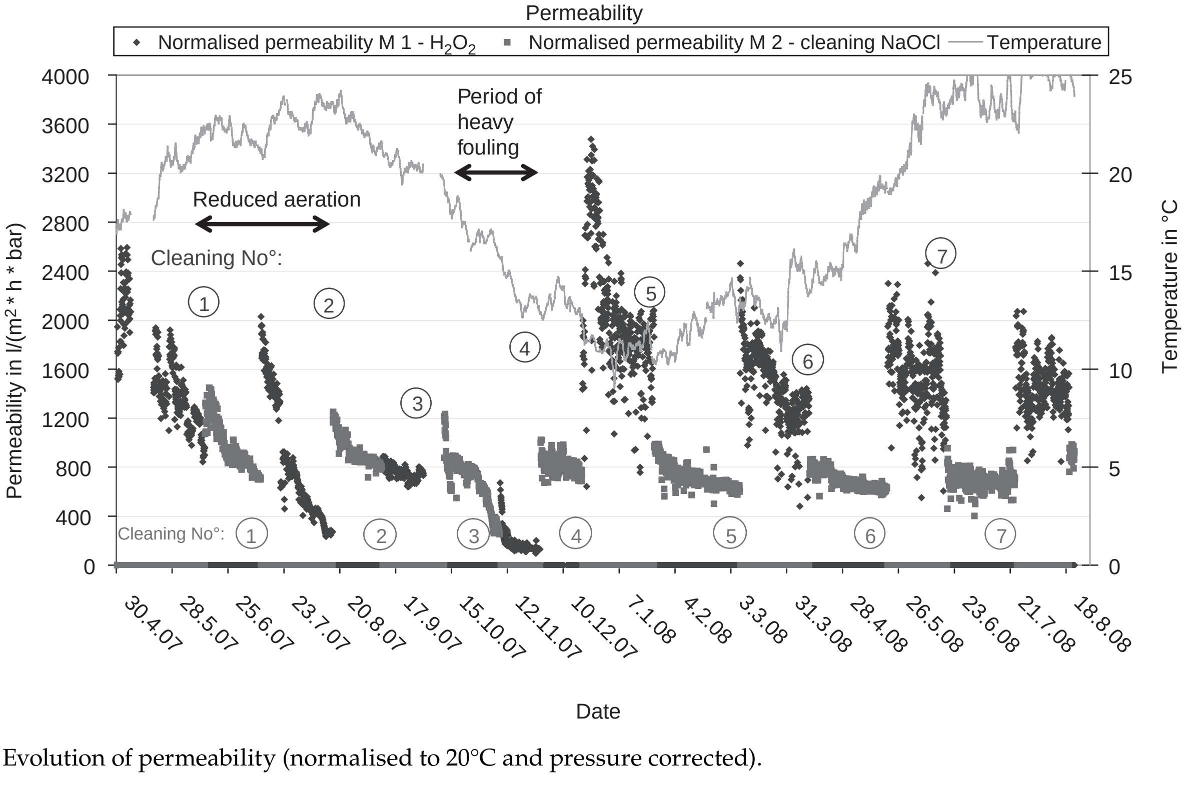

Fig. 3. Evolution of permeability (normalised to 20°C and pressure corrected).

way that could lead to a reduced lifetime of the module. However, this did not affect significantly the subsequent filtration behaviour of this module, nor its disinfection performances.

These 16 month trials demonstrated that the selected cleaning strategy was appropriate for small semi-decentralised MBR units with the filtration modules supplied by A3 Water Solutions in the investigated operation conditions: it offered relative operational security while optimising the planning of the maintenance and reducing the staff presence on site.

3.3. Permeability recovery

The permeability recovery was determined using Eq. (7). Figure 4 shows the recovery evolution for each monthly cleaning with both cleaning agents. Module 1 cleaned with H_2O_2 is displayed in dark blue and module 2 NaOCl in red. The numbers refer to the cleanings as indicated in Fig. 3. The earlier discussed events of instable operation are also reflected in the cleaning results. When module 1 was commissioned during July 2007, the permeability fell due to the reduced air scour. The cleaning was not representative of the rest of the period with moderate fouling. Similarly, the event of heavy fouling in November 2007 led to an interfact in the calculation of the recovery: Cleaning N° 3 shows a negative recovery, due to the strong fouling within the first day of operation. As two day mean values are used for calculation of the permeability, a rapid decline within the first day of operation makes it impossible to determine a cleaning effect. As described above the bad performance during this *fouling event* led to an adapted cleaning protocol and the recovery of cleaning No° 4 was over 100%.

The cleaning strategy included also the use of citric acids according to the performance of the filters (as indicated in Fig. 4 for cleanings N° 3 and 7). Cleaning No° 7 shows the importance of citric acid, to remove inorganic foulants that might have accumulated over time and contribute to irreversible fouling.

Considering the cleanings that were performed during stable operation both agents show a similar decline of the cleaning efficiency, and similar long-term accumulation of irreversible fouling, which can be partly recovered by cleanings with citric acid. According to these results, it can therefore be concluded that the cleaning efficiency of both agents NaOCl 500 ppm (no pH adjustment) and H_2O_2 1000 ppm (pH 11) show equivalent performances and will require occasional cleaning with citric acids (every 3 to 4 cleanings), and probably yearly cleaning with high grade NaOCL (for example 2000 ppm, pH 11).

As chlorine is considered in Germany and other countries to be hazardous to water bodies and decentralised waste water systems do usually not have the capacity to store toxic agents in a safe way, an alternative agent being easier and safer to handle showing the same cleaning results is an improvement in the field of decentralised waste water treatment with membranes. The results accumulated over these 16 months of operation show that H_2O_2 is a feasible option for

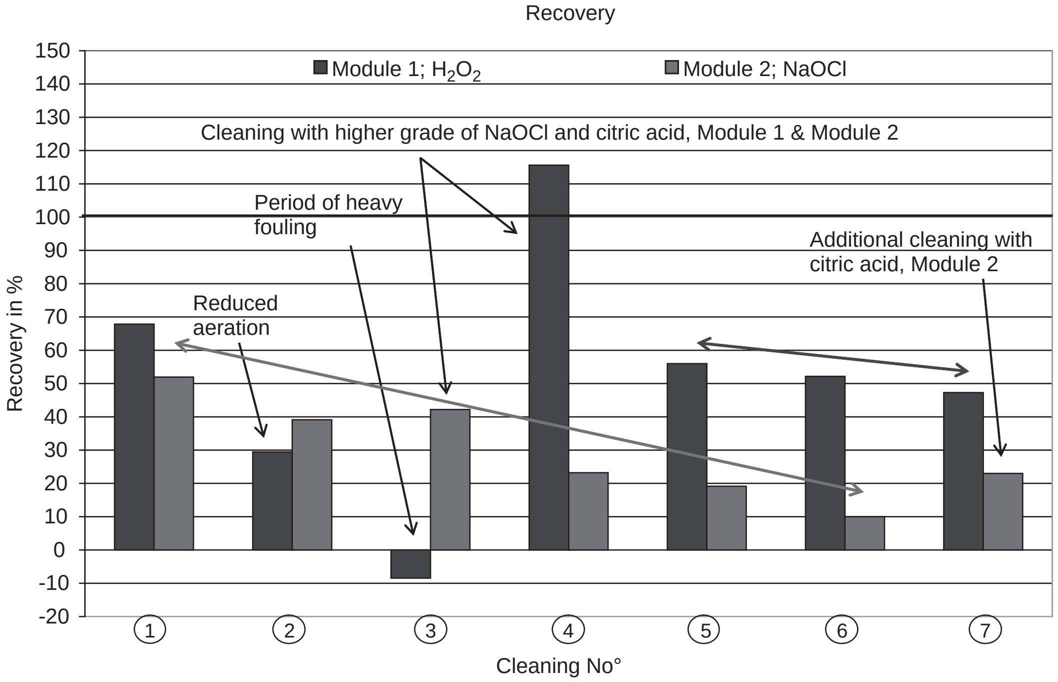

Fig. 4. Recovery of permeability after cleaning.

semi-decentralised MBR plants, and confirms the experience reported by Wedi et al. (2007) [9] on the Monheim MBR plant (9,700 p.e.; 288 m^3/h), cleaned with maintenance cleaning (H_2O_2 2000 ppm; pH 9.7) on a 2-week basis.

3.4. Costs of investigated cleaning strategies

The costs for cleaning agents do not play a major role compared to the operational costs. Table 5 shows the operational costs for both agents and the used chemicals, citric acid and NaOH, for the demonstration plant in Margaretenhöhe (Gnirss et al. 2007b [10]). H_2O_2 is more expensive than NaOCl (0.086 €/m^3 compared to 0.047/m^3 for a throughput of 10 m^3/d) but in relation to the operational costs (w/o personnel costs) the difference is about 1% additional costs between both agents. The benefits of H_2O_2, e.g. less toxic to water bodies, do compensate the marginal higher costs. These costs have been calculated with delivery prices for small amounts of the used chemicals (25L Jerricans). As storage of chemicals on site is usually difficult due to safety issues these costs are more suitable to assume the costs for cleaning agents of decentralised wastewater treatment plants.

4. Conclusions

Over 16 months two cleaning strategies using low grade agents, NaOCl and H_2O_2 respectively, were compared. The MBR demonstration plant operated in the frame of the EU project ENREM gave the opportunity to operate one membrane module at a time while the other could soak in cleaning agent. The modules in operation were switched in intervals of approximately four weeks. This set up gave the possibility to compare the efficiency of two different cleaning agents dealing with foulants caused by the same activated sludge, biological process and raw waste water.

The results of this study show, that during stable operation the cleaning efficiency of both agents were similar. It was also demonstrated, that an additional cleaning step with citric acid should be carried out 2–3 times a year in order to remove inorganic foulants, and on a yearly basis a high grade NaOCl cleaning. Throughout the reported period it was possible to plan the cleaning actions in advance which helped to manage the operation of the plant and reduced unforeseen immediate actions and operational risks.

The costs for the cleanings with H_2O_2 are twice as high compared to chlorine, but the cleaning agents represent only 1%–2 % of the total operational costs.

The investigation showed that H_2O_2 is a feasible alternative to chlorine as a low toxic cleaning agent for membranes used in wastewater treatment also for semi-decentralised applications. The proposed cleaning strategy is suitable for a number of membrane activated sludge plants and increases the flexibility to react to unforeseen disturbances, as a second filtration line is always available.

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