Modelling of moving bed biofilm membrane reactors (MBBMR) for on-site greywater treatment

Simon Jabornig and Wolfgang Rauch

ABSTRACT

The study evaluates with a mechanistic model the pilot plant results of a combined moving bed biofilm process and membrane filtration (MBBMR) treating single household greywater. It mainly includes the simulation of reactor hydraulics, degradation of pollutants, development of biomass and settlement of sludge. Iterative calibration was made with steady-state results of a 10-month pilot test. The model shows good predictions of readily biodegradable chemical oxygen demand and ammonium removal, as well as biomass concentration on carriers and in suspension. Also, a sensitivity analysis was made which calculates the relative significance factor of each model coefficient and by this provides comparability with other studies. Simulation data and actually measured parameters show that the suggested process was rather independent of ambient temperatures and short-term load fluctuations. Obtained datasets and model structure could be of use for future designers, as well as sellers and users of this process for on-site greywater reclamation.

Key words | greywater, MBBMR, membrane bioreactor, modelling, moving bed biofilm

Simon Jabornig (corresponding author)
Wolfgang Rauch
Unit of Environmental Engineering,
Institute for Infrastructure Engineering, University of Innsbruck,
Technikerstraße 11,
A–6020 Innsbruck,
Austria
E-mail: *simon.jabornig@gmx.net*

NOMENCLATURE

A	bioreactor tank area [m^2]
b_H	endogenous decay coefficient [1 d^{-1}]
c_A	concentration of material A in bioreactor [mg L^{-1}]
dt	numeric time step [h]
$dT_{B,conv}$	temperature change of bioreactor by natural convection [°C]
dT_B	temperature change of bioreactor [°C]
f_{ES}	soluble fraction of endogenous residues
f_{EX}	particulate fraction of endogenous residues
i_{SN}	nitrogen fraction of soluble microbial products [gN g^{-1} COD]
i_{SP}	phosphorus fraction of soluble microbial products [gP g^{-1} COD]
i_{NH}	mass of nitrogen per mass of COD in biomass [gN g^{-1} COD]
i_{PO}	mass of phosphorus per mass of COD in biomass [gN g^{-1} COD]
i_{XN}	nitrogen fraction of particulate microbial products [gN g^{-1} COD]
i_{XP}	phosphorus fraction of particulate microbial products [gP g^{-1} COD]
k_a	aeration rate of bioreactor [–]
k_h	maximum hydrolysis rate of X_S [d^{-1}]
k_{hR}	maximum hydrolysis rate of S_{RB} [d^{-1}]
k_{T0}	growth rate constant at temperature T
k_T	temperature coefficient which indicates how strongly the reaction is accelerated per °C [$°C^{-1}$]
k_{TB}	temperature coefficient which indicates how strongly the bioreactor is cooled down by ambient temperature per °C [$°C^{-1}\,h^{-1}$]
K_N	half-saturation coefficient for ammonium nitrogen
K_P	half-saturation coefficient for phosphate phosphorus
K_S	half-saturation coefficient for readily biodegradable substrate
K_X	half-saturation coefficient of hydrolyses
m_i	measured value of the output variable
p_i	predicted value of the output variable
Q_A	aeration [L air h^{-1}]

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Q_{in} inflow to bioreactor [L h^{-1}]
Q_{out} effluent of bioreactor [L h^{-1}]
R_A aeration ratio ON/OFF [–]
r_A process rate of material A in bioreactor [mg h^{-1} L^{-1}]
S_{EF} effluent COD [mg L^{-1}]
S_S readily biodegradable COD [mg L^{-1}]
S_O dissolved oxygen [mg L^{-1}]
$S_{O_2,A}$ oxygen introduced by blower [mg L^{-1}]
S_I inert soluble COD fraction in the influent [mg L^{-1}]
S_P soluble microbial products [mg L^{-1}]
S_{MB} rejected fraction of the soluble residual COD [mg L^{-1}]
S_R residual soluble COD [mg L^{-1}]
S_{RB} resulting residual soluble COD in the MBR [mg L^{-1}]
S_{NH} ammonium nitrogen [mg L^{-1}]
T_B bioreactor temperature [°C]
T_{in} inflow temperature [°C]
T_R ambient temperature [°C]
μ_H maximum growth rate of heterotrophs [h^{-1}]
V_B bioreactor volume [L]
w_S sedimentation velocity [m h^{-1}]
X_H active heterotrophic biomass in suspension and on biomass carriers [mg L^{-1}]
X_{HB} active heterotrophic biomass in biofilm [mg L^{-1}]
X_{HS} active suspended heterotrophic biomass [mg L^{-1}]
X_P particulate microbial products [mg L^{-1}]
X_S Slowly hydrolysable COD [mg L^{-1}]
Y_H heterotrophic yield coefficient [mg cell COD mg^{-1} COD]
Y_{O_2} O_2 transfer coefficient diffuser [mg O_2 L^{-1} air]

INTRODUCTION

Wastewater recycling in private households is a promising way to overcome potable water shortages in arid regions. Reused greywater (GW) could be around 50% of total wastewater and consists of used water from showers, baths, washing machines and hand basins. Although GW is less polluted than mixed sewage, it cannot be directly reused due to organic pollutants which would strongly enforce microbial growth in pipes and storages. What is more, the high concentration of suspended solids and high number of faecal coliforms (Casanova *et al.* 2001) would affect hygienic acceptance. Requested effluent quality in international reuse guidelines usually requires chemical, physical or biological treatment including disinfection. The literature reports numerous combinations of the same (Li *et al.* 2009). However, treating and recycling GW could be a first transition step to full water reuse as mixed sewage still faces acceptance problems in private household applications (Nolde 2005; Wach *et al.* 2008). A positive example for wastewater reuse in general was presented with the membrane bioreactor (MBR). This approach combines biological treatment by activated sludge process followed by microfiltration or ultrafiltration. While in the bioreactor, long sludge ages can be applied, which removes particulate and dissolved nutrients to a high extent; the physical membrane barrier additionally withdraws all other suspended solids and micro-organisms. This significantly reduces plate count and the effluent quality resembles that from disinfection. Several publications document that the process combination meets the guidelines and fulfils the expectations of users (Merz *et al.* 2007; Kraume *et al.* 2010; Jabornig & Favero 2013). However, the contradiction of increasing acceptance by enhancing treated water quality and on the other hand reducing investment and operating costs for a payback period of a reasonable period of time is still unsolved for single-household MBR applications (Jabornig 2013). One alternative which could overcome this disadvantage is a combination of moving bed biofilm process and membrane filtration (MBBMR) (Leiknes & Ødegaard 2007; Jabornig & Favero 2013). Authors report that less suspended solids in the bioreactor suspension lowered the operation costs, which is the effect of strong reductions of membrane fouling control (back-flushing, air scouring and chemical cleaning). Modelling studies of activated sludge processes (Gujer *et al.* 1999), biofilm processes (Boltz *et al.* 2011) and MBR applications for mixed sewage (Fenu *et al.* 2010) and GW (Hocaoglu *et al.* 2013) were formerly issued. In addition, the present paper provides a mechanistic model for bioreactor and hydraulic characteristics for a combined moving bed biofilm reactor with a submersed microfiltration membrane. A model for a similar process combination has not been published before. Therefore, the main task of this paper is to describe the technology with approved modelling methods in wastewater research and provide a sensitivity analysis of main parameters. Data from previous publications (Jabornig & Favero 2013) and new datasets were used to calibrate the bioreactor and membrane filtration model with special emphasis on degradation of organics, limitation of nutrients and development of suspended solids in the suspension. Further attention was drawn to typical situations and the

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environment of single-household applications and, by this, showing their impact on process stability and treated water quality.

MATERIAL AND METHODS

GW characteristics

The pilot plant was operated with 180 L d^{-1} synthetic GW freshly mixed with different shower gels, shampoos, soaps, deodorants and other typical GW ingredients according to NSF/ANSI 350 (2011) with COD:N:P (chemical oxygen demand:nitrogen:phosphorus) of 100:2.28:0.25. One day a week was assumed to be a washing day with the feeding of additional laundry detergents and softeners. Kitchen wastewater was not included throughout the tests. The daily volume represents roughly a four-person household in Central Europe (Nolde 1999). The fractions of total COD consisting of readily biodegradable (S_S), slowly hydrolysable (X_S) and inert COD (X_I) on total COD in GW was assumed to be 35, 60 and 55%, typical for GW (Hocaoglu *et al.* 2013). Feeding of GW was done in batches in the morning (40%), noon (20%) and evening (40%) in order to simulate actual inflow conditions.

Pilot plant

The pilot plant was operated for a period of 10 months and combines a moving bed biofilm process with a microfiltration membrane (C-MEM, Austria, hollow fibre, HDPE, 6 m^2, net flux during permeating phase 12.9 L m^{-2} h^{-1}) with a nominal pore size of 0.2 µm in one bioreactor. It was operated in batches (hydraulic retention time 24 h) dependent on the actual inflow of GW and flux of membrane. There were mainly two extraction times of treated effluent in the afternoon and early in the morning in order to avoid bypassing of fresh GW. Extraction of treated water (permeate) was done in two batches, each lasting a maximum of 3 h, so in total 6 h per day. Equipment selection and design of the bioreactor were made according to Jabornig & Favero (2013), which included PLC-controlled permeate and back-flushing pump (40 W, <4 L min^{-1}), non-continuous time-controlled co-current aeration device for membrane and bioreactor (29 W, 30 L min^{-1}) and cylindrical biomass carriers (HDPE, 0.95 kg m^{-3}, 320 m^2 m^{-3}) with filling rate of 30% in bottom water level. The treated effluent was pumped to a small back-flushing storage, which was followed by a treated water tank with a capacity for a whole-day permeate production (200 L). Surplus sludge and detached biomass from the carriers were extracted manually with a siphon on a monthly basis. The pilot plant was situated in a covered but non-temperature-controlled facility with ambient temperatures between 0 and 30 °C.

Model characteristics

Reactor hydraulic

Influent and effluent were considered as discontinuous and set up as follows: inflow to the bioreactor model was set to 75 L from 7 to 8 a.m., 25 L from noon to 1 p.m. and 80 L from 7 to 8 p.m. Raw water entered the system as short flushes at 10 L min^{-1}. Effluent by permeate extraction was 90 L from 4 to 7 p.m. and additional 90 L from 4 to 7 a.m. as constant output (flux) from the membranes. A steady-state (stabilized) flux in a low range (<5 L m^{-2} h^{-1}) is usually reached 1–3 weeks after start-up of membrane filtration even without or strongly reduced fouling control (Peter-Varbanets *et al.* 2010; Jabornig & Favero 2013) and was therefore used as output during permeating phase of the plant. Owing to the discontinuous hydraulic sequence the reactor was modelled as a batch reactor with variable water table. The bottom water level of the bioreactor was limited to 250 L, the top water level to 350 L. This can be described for substance A with the commonly known mass balance equation:

$$V_B \frac{dc_A}{dt} + c_A \frac{dV_B}{dt} = Q_{in} c_{A,in} + r_A V_B - Q_{out} c_A \quad (1)$$

$$\frac{dV_B}{dt} = Q_{in} - Q_{out} \quad (2)$$

The first term of expression (1) is the change of material A by time at a constant bioreactor volume V_B. The second term expresses that the mass of material A in this tank is also influenced by the variability of water volume by discontinuous influent (Q_{in}) and effluent (Q_{out}). Both terms need to be equal with incoming, evolving and outgoing loads of material A to fulfil the overall mass balance of the bioreactor. This concept came along with the assumption that change of concentration of material A in the batch reactor was dependent not only on degradation by a certain process rate r_A, but also on water level.

$$dc_A = \left(\frac{Q_{in} c_{A,in}}{V_B} + r_A - \frac{Q_{out} c_A}{V_B} \right) dt - c_A \frac{(Q_{in} - Q_{out})}{V_B} \quad (3)$$

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Temperature influence on the process rates in the batch reactor (i.e. for high inflow and low ambient temperatures) was also implemented in the model. The ambient temperature influence was assumed as exponential function. The used parameters were experimentally derived during the test period. The heat capacity of water c_p was assumed as constant in that small temperature range and was therefore excluded from equations.

$$dT_B = \frac{dT_{B,conv}V_B + (T_{in}Q_{in} - T_BQ_{out})dt}{V_B + (Q_{in} - Q_{out})dt} \quad (4)$$

$$dT_{B,conv} = T_B e^{-k_{TB}(T_B - T_R)dt} \quad (5)$$

Bioreactor

The basic elements of the mechanistic model used in this study, shown in Table 1, were adapted from Hocaoglu *et al.* (2013), who applied their ASM1-based model on GW treatment with conventional activated sludge membrane bioreactors (AS-MBR). The model includes the following constituents in the bioreactor: (1) degradation of readily biodegradable COD (S_S), (2) slowly hydrolysable COD (X_S), (3) soluble metabolic products (S_P), (4) particulate metabolic products (X_P), (5) resulting residual soluble COD in the bioreactor suspension (S_{RB}), (6) ammonium (S_{NH}), (7) phosphate (S_{PO}), (8) dissolved oxygen (S_O) and (9) heterotroph biomass (X_H). Four main processes were implemented to simulate the pilot test results. (1) Microbial growth was described through a first-order degradation reaction of S_S by heterotrophic biomass in the biofilm and suspension. A specific model on biofilm flux kinetics was not implemented. The reason was that former studies (Jabornig & Favero 2013) showed that the biofilm in this process combination is thin and therefore full penetration can be assumed. The process additionally included two switching functions for ammonium and phosphate in order to predict typical limitations of these parameters in GW. Ambient conditions of the pilot plant were considered in the process by an experimentally derived temperature-dependent growth rate.

$$\mu_H = k_{T_o} e^{(k_T(T - T_o))} \quad (6)$$

(2) Hydrolysis of X_S and (3) hydrolysis of S_{RB} were also implemented as first-order reactions dependent on the ratio of X_S/X_H and S_{RB}/X_H in the bioreactor. The additional

Table 1 | Stoichiometric matrix of GW MBR bioreactor model (adapted from Hocaoglu *et al.* (2013))

	Material i											
Process j	Soluble substrates S_S $g_{COD}\ m^{-3}$	Particulate substrates X_S $g_{COD}\ m^{-3}$	Soluble microbial products S_P $g_{COD}\ m^{-3}$	Particulate microbial products X_P $g_{COD}\ m^{-3}$	Resulting residual soluble COD in the MBR S_{RB} $g_{COD}\ m^{-3}$	Ammonium S_{NH} $g_N\ m^{-3}$	Phosphate S_{PO} $g_P\ m^{-3}$	Oxygen S_{O2} $g_{O2}\ m^{-3}$	Biomass/ biofilm X_H $g_{COD}\ m^{-3}$	Process rate ρ_j $g_{COD}\ m^{-3}\ d^{-1}$
1 Microbial growth	$-\frac{1}{Y_H}$					$-i_{NH}$	$-i_{PO}$	$-\frac{1-Y_H}{Y_H}$	1	$\rho_1 = \mu_H \frac{S_S}{K_S + S_S} \frac{S_{NH}}{K_N + S_{NH}} \frac{S_{PO}}{K_P + S_{PO}} X_H$
2 Hydrolysis of X_S	1	-1								$\rho_2 = k_h \frac{(X_S/X_H)}{K_X + (X_S/X_H)} X_H$
3 Hydrolysis of S_{RB}	1				-1					$\rho_3 = k_{hr} \frac{(S_{RB}/X_H)}{K_X + (S_{RB}/X_H)} X_H$
4 Microbial decay			f_{ES}	f_{EX}		$i_{NH} - f_{ES}i_{SN} - f_{EX}i_{XN}$	$i_{PO} - f_{ES}i_{SP} - f_{EX}i_{XP}$	$-(1 - f_{ES} - f_{EX})$	-1	$\rho_4 = b_H X_H + \frac{w_S A}{V_B} X_{HS}$

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hydrolysis step of S_{RB} was included because of the retention and accumulation of additional particulate matter through the membrane barrier (Hocaoglu *et al.* 2013). (4) Microbial decay by endogenous respiration was included as a linear function generally used in activated sludge models (ASMs). Furthermore we assumed that growing biomass on the carriers is spontaneously removed by intermediate coarse bubble aeration and settled during non-aerated phases. The process of irreversible settlement was included with a linear expression to the microbial decay. Settlement was assumed to be dependent on the settling velocity of free-floating biomass in the suspension during non-aerated phases. The settling velocity was further dependent on the aeration characteristics and was experimentally derived during a 30 min settling test. Actual biofilm mass on carriers (X_B) was included in the model with an experimentally derived biofilm load which depended only on aeration frequency (k_a) of the bioreactor. The measurements of biofilm load and aeration frequency were added to a diagram and approximated with a quadratic function $X_B(k_a)$. X_H was calculated as a sum of heterotrophic biomass in biofilm X_B and suspension X_{HS}, which were not considered with different characteristics in the model.

$$X_H = X_B(k_a) + X_{HS} \tag{7}$$

Oxygen supply to the bioreactor was simulated by a linear function dependent on the aeration characteristics of the co-currently working aeration diffusers for membrane air scouring and bioreactor aeration. The parameters for this process were also experimentally derived.

$$S_{O_2,A} = \frac{Y_{O_2} Q_A R_A}{V_B} \tag{8}$$

The overall process was simulated in MS EXCEL software on the basis of numerically solved differential equations with an output time step of 1 h for long-term development and 1 min for short-term degradation of S_S and oxygen uptake rate. Coefficients and parameters were iteratively calibrated based on experimentally derived steady-state data from the pilot tests. The procedure involved a manual calibration protocol. The model components were adapted in each iteration step in order to fit actual measured data. The starting point for each parameter was a reference value from literature (Gujer *et al.* 1999; Hocaoglu *et al.* 2013) or ASM1 default. One combined dataset of 10-month pilot testing was used for calibration. Validation of the current dataset is pending. Further experimentally gathered datasets will be required to perform this to a sufficient extent.

During calibration the accuracy of the model predictions was compared to actual measured data by means of average relative deviation (ARD) (Makinia *et al.* 2006):

$$\mathrm{ARD} = \frac{1}{N} \times \sum_{i=0}^{N} \frac{|(m_i - p_i)|}{m_i} \times 100\% \tag{9}$$

N is the number of observations, m_i the measured value and p_i the predicted value from the model. The goal of the calibration exercise was to find an ARD minimum among main output parameters S_S, S_{NH} and X_{HS}.

The continuity of the stoichiometry matrix and consistency of the model kinetic rate expressions of the matrix were checked by the methods suggested by Hauduc *et al.* (2010).

Sensitivity analysis

Sensitivity analysis was made with the relative sensitivity factor (RSF) function according to Jiang *et al.* (2005). Results were used during calibration to identify the most influential coefficients of the model. θ represents the coefficients of the stoichiometric matrix and Y the main variables of the model.

$$\mathrm{RSF} = \frac{\theta}{Y} \frac{dY}{d\theta} \tag{10}$$

The derivate was calculated with two-point finite difference formula starting from the steady-state calibrated value.

$$Y'(\theta) = \frac{Y(\theta + h) - Y(\theta - h)}{2h} \tag{11}$$

The parameters were grouped into the following partitions: $\mathrm{RSF} < 0.25$ (not influential), $0.25 < \mathrm{RSF} < 1$ (moderately influential), $1 < \mathrm{RSF} < 2$ (very influential) and $\mathrm{RSF} > 2$ (extremely influential). Only values above 0.25 were included in the manual calibration procedure.

A simple approach was preferred to recently reported global methods (Sin *et al.* 2011; Cosenza *et al.* 2014), the reasons being that the model has – in comparison – a manageable complexity, and more data from RSF analysis of similar processes was already available in literature (Fenu *et al.* 2010).

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RESULTS AND DISCUSSION

Hydraulic characteristics

The pilot plant was fed daily with 180 L fresh GW in flushes in the morning, noon and evening, simulated as short peaks. Treated water was extracted batchwise through the membrane 6 h a day at a lower and more constant level. Both short-term influent and effluent resulted in a strong change of bioreactor volume of 40% and therefore also in an observable change of concentrations of suspended and dissolved substances in the reactor after feeding. Figure 1 shows the simulation of typical inflow (Q_{in}), treated water effluent by filtration (Q_{out}) and volume of bioreactor (V_B) which was measured during the pilot tests and used for the hydraulic settings of the model. These special hydraulic characteristics and the fact that the bioreactor and membrane are situated in the same tank resulted in the need to separate inflow and effluent temporarily as best as possible. Any other configuration was shown to have higher bypass flows of non-treated GW and therefore lower treatment performance. While the influent was time-controlled and kept constant throughout the experiment, the effluent of the bioreactor through the membrane, expressed as flux in L m^{-2} h^{-1}, fell strongly during the first operating days. The results indicated that the run-in phase of the membrane can be compared with the process stabilization of the bioreactor. The effluent COD of the bioreactor process stabilized within 15–20 days. Flux during permeating phase stabilized after approximately 30 days of operation at a value of about 5 L m^{-2} h^{-1}. The stabilization of the flux was previously reported by Peter-Varbanets *et al.* (2010) and Jabornig & Podmirseg (2014) for low-pressure ultrafiltration applications. This phenomenon is probably due to a steady state of the filter cake thickness and porosity on the membrane surface. During operation, deposits and micro-organisms are adsorbed through filtration and detached again by fouling control. This steady-state process could be described in a similar way for the free moving biomass carriers used in the study.

Bioreactor characteristics

The biological degradation inside the bioreactor was modelled and calibrated with data obtained from the pilot test period (10 months). As main processes, biomass development, treated water quality and oxygen uptake were examined in more detail.

Biofilm mass on cylindrical carriers was constant once it had built up. At 50% (10 min on/10 min off) aeration rate the biofilm on the carriers was measured, with an average of 7 g m^{-2}. The amount increased to its maximum of 9.28 g m^{-2} at 9.1% (1/10) aeration rate. As the aeration blower was not flow controlled, the biofilm load on the carriers could only vary depending on aeration frequency. To fulfil the observed

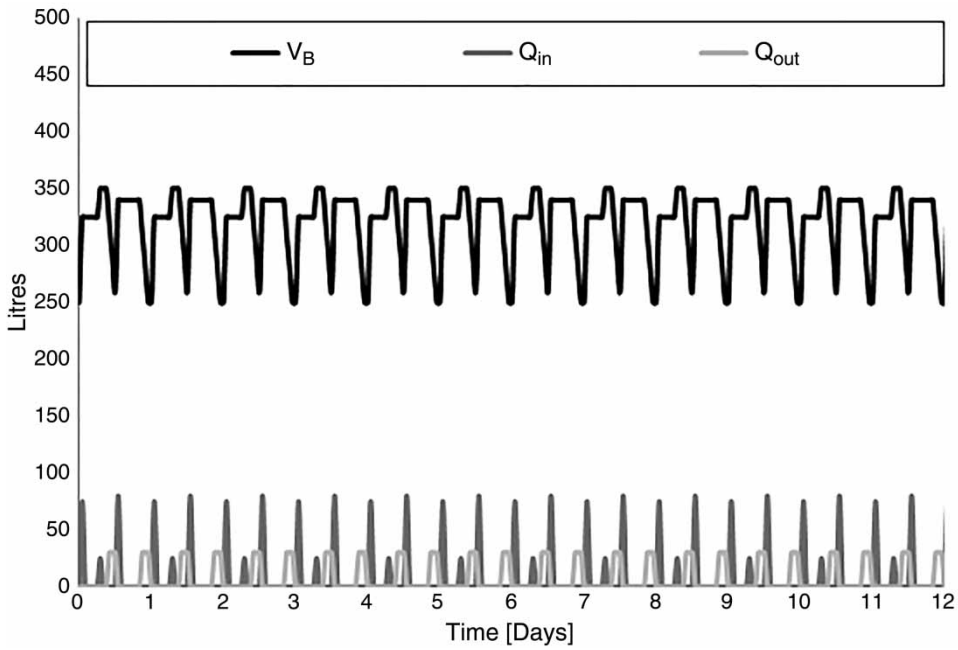

Figure 1 | Modelled hydraulic characteristics of bioreactor volume (V_B), inflow (Q_{in}) and effluent (Q_{out}).

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balance between growth and decay, all additional establishing biomass on the carriers had to be released into suspension by turbulence. Interestingly, the concentration of suspended solids in the bulk liquid of the bioreactor did not increase either, but stayed at a rather low and constant level. This equilibrium could be explained by irreversible settling of suspended biomass X_{HS} during non-aerated phases. Settled sludge did not disperse again through the aeration device implemented into the membrane module. So the actually active, total amount of heterotrophic biomass X_H in the bioreactor was also in a steady state throughout the stabilized time of pilot tests. Stabilization bioreactor characteristics established about 5 weeks after start-up of the pilot plant when the moving bed carriers were finally filled up with biofilm. Steady state of the bioreactor could be maintained until the end of the pilot tests. Figure 2 shows the modelled and measured concentration of biofilm biomass carriers and suspension (ARD 65%). The actual concentration of biomass in the model is related to the filling level. It varies when the water level changes during feeding and permeate extraction. The trend of X_{HS} compared to X_{HB} over the day was more even because of irreversible settling and thereby continuous removal of regrown biomass from the system. The settled sludge on the bottom of the tank was siphoned off at an amount of 8 L month^{-1} and was measured at a concentration of 14–16 g L^{-1}. The bottom sludge degradation characteristics were not included in the model. The reason was that mass balance between extracted bottom sludge and theoretical biomass growth without sludge removal or settlement showed a difference by about 50%. The lower amount of surplus sludge indicated that the endogenous decay rate of biomass must have increased in the bottom sludge compared to biomass in suspension and biofilm. One explanation could be a very low diffusion rate of nutrients into it. As a consequence settled biomass was not considered to be active in the process anymore.

Readily biodegradable COD (S_S) in the treated water tank during the stabilized phase of the pilot tests was in the range of 40–60 mg L^{-1}. The simulation of this parameter shows the typical variation over the day resulting from changing reactor level (Figure 2, ARD 13%). Nonetheless, the variation appeared only in the simulation because actual measurement combined the treated water from 6 h of filtration and therefore high peaks were usually omitted. The readily biodegradable COD value was comparably higher than in other GW MBR studies (Merz *et al.* 2007; Kraume *et al.* 2010; Hocaoglu *et al.* 2013), indicating slow degradation and also verifying the fitted low growth rate of biomass of the model. Oxygen as limitation could be excluded because it was available in excess even in the lowest aeration rate. The limitation of growth rate and shortened S_S removal could be further enforced by a low concentration of nutrients of ammonium and phosphate. This assumption follows the low ammonium concentrations which were measured in the treated water

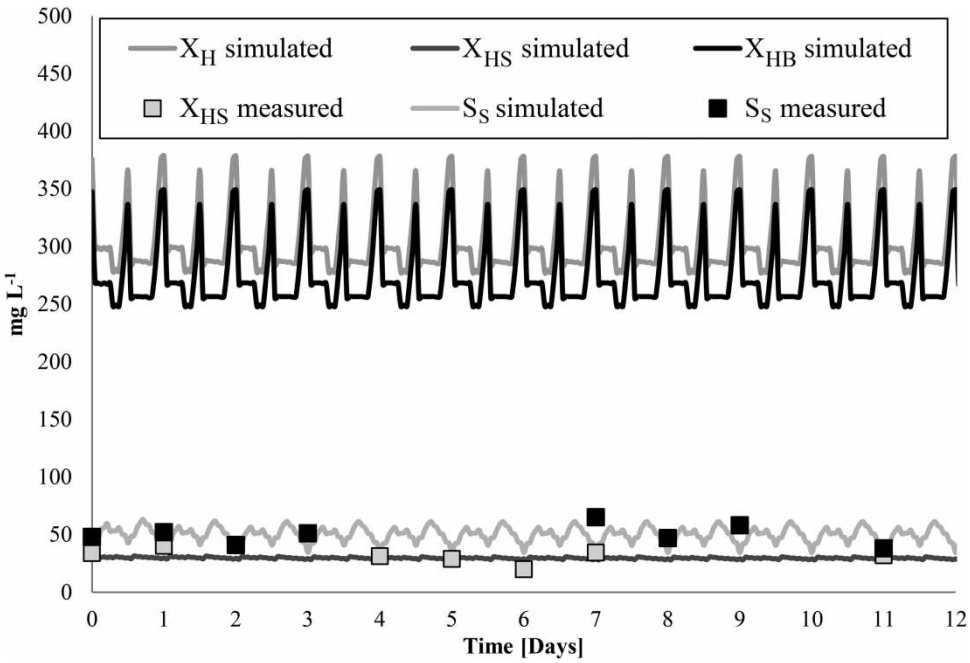

Figure 2 | Simulated and measured concentration of biomass and S_S in bioreactor at 9.1% aeration rate.

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tank to be almost exclusively lower than 0.1 mg L^{-1}. Modelled mean values of S_{NH} in the treated water show a good prediction of actual measured values (ARD 78%). However, peaks up to 0.25 mg L^{-1} right after dosing of GW were predicted but could not be seen in actual measurements. Uptake and degradation was finished before time-delayed permeate extraction started. Similar data could be measured and modelled for phosphate. In consideration of these results the fitted growth rate of heterotrophs at a value of 0.69 day^{-1} was lower than in other blackwater and GW studies. The slow growth rate of biomass was also consistent with the low oxygen consumption of the process. Figure 3 shows that right after feeding of fresh GW the model predicted a sharp oxygen decay which was enforced by degradation, dilution and temperature increase. After this fall, oxygen was consumed by the microorganisms slowly, almost linearly. Similar oxygen uptake characteristics could be confirmed by the O_2 measurements in the bioreactor of the pilot plant, respectively.

Sensitivity analysis

A sensitivity analysis was made for the main coefficients of the model, showing their influence on the overall process. Over-parameterization could be seen for K_X, k_{hR} and K_P which were not found to be influential for main output parameters S_S, X_S, S_{NH} and X_{HS} with RSF < 0.25. The parameters K_S, K_N, k_h, T, b_H and w_S were ranked moderately influential on S_S with RSF 0.25–1. Results also show the strong influence of growth rate and yield factor with RSF > 2 on the treatment performance of S_S. In contrast, coefficients for hydrolysis played only a minor role and were of a comparable low range also in literature (Jiang *et al.* 2005). Special attention should be given to the influence of ambient temperature on the process due to the outdoor installation. Congruent with the results of the pilot tests, outside temperature was of minor importance because of the slow heat transfer by radiation and convection. The actual biomass concentration in suspension proved to be further dependent on the sedimentation velocity (w_S).

CONCLUSION

The treatment combination consisting of MBBMR in one bioreactor could be successfully modelled and calibrated with obtained pilot plant results. The following conclusions could be drawn from the results:

- Flux stabilization effect of membrane led to a stable hydraulic profile and provided also balanced conditions for the biofilm treatment process.
- Low growth rate of process could be a result of limiting nutrient ratio. Dosing of nutrients would be an option but is probably not feasible in terms of operating costs.

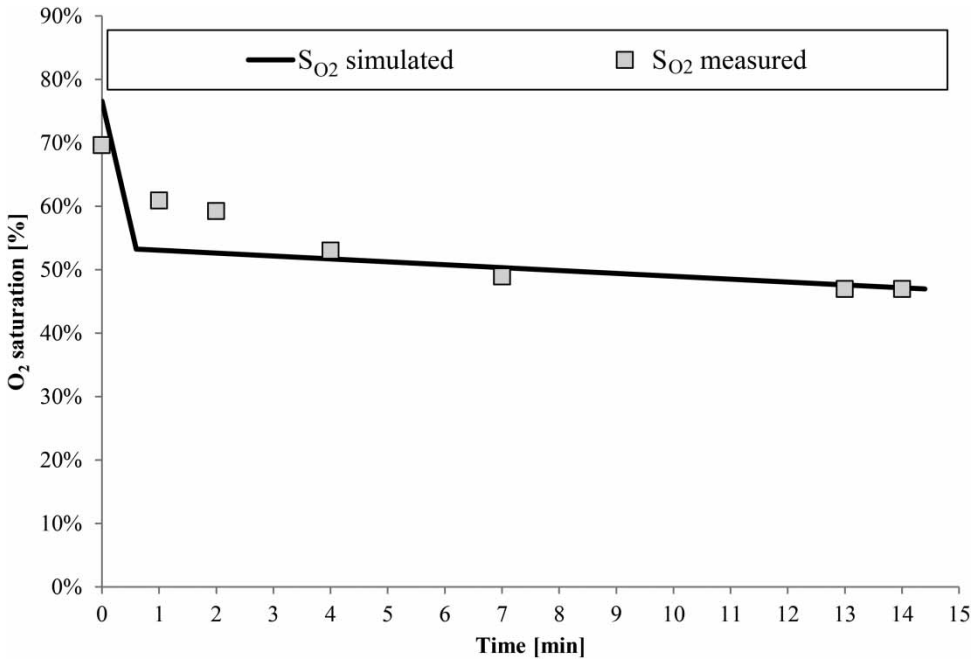

Figure 3 | Oxygen saturation decay in the bioreactor with switched-off aeration blower shortly after feeding fresh GW. The straight line shows a comparison between modelled and measured O_2 saturation.

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- Overall biomass concentration in the bioreactor was in a stable-state condition due to continuous irreversible settling of sludge. Therefore, a regular surplus sludge removal for this process was not necessary.
- Site of installation of the unit could be in a non-temperature-controlled facility, e.g. basement, because heat loss by radiation and convection through the bioreactor walls during the retention time of the process and heat input by warm GW remained balanced even at freezing level.
- Future work needs to focus on gathering a more experimentally derived dataset to perform sufficient validation of the proposed model.

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