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GREYWATER TREATMENT IN A SUBMERGED MEMBRANE SEQUENCING BATCH REACTOR (SM-SBR)

Greywater reuse will play an important role in the sustainable water management approach. Depending on its intended use, different treatment technologies are needed. This study focuses on the performance of a submerged membrane sequencing batch reactor (SM-SBR) for biological treatment of synthetic greywater. Two different reactors are operated: the first one, a 27 dm^3 lab-scale reactor, and the second one, a commercial 1 m³ MBR manufactured by the BUSSE company. In terms of COD removal, the first analyses show a good performance under different operating conditions, e.g. varying cycle time (t_c) and the volumetric exchange ratio (VER). The ammonium utilisation rate (AUR) and the nitrate utilisation rate (NUR) are evaluated and compared to data found in literature. The NUR shows limitations due to slow-degrading COD of synthetic greywater, resulting in the question whether an additional carbon source should be used to enhance the process.

Keywords: *SBR, membrane, synthetic greywater, low carbon, denitrification, NUR, AUR*

1. INTRODUCTION

1.1. GREYWATER REUSE FOR SUSTAINABILITY

The reuse of treated greywater helps to save valuable water resources, especially in the regions where water restrictions are a severe problem and water supplies are rapidly declining. In such circumstances, the utilisation of drinking water for, e.g., toilet flushing seems to be inappropriate. Society has to become aware of efficient and appropriate water (re)use. Examples of such a 'demand of side-water management' include the use of water-saving devices (faucets, showers, toilets etc.), segregation of wastewater streams, and the reuse of greywater [1], [20]. Greywater will become more significant in the near future for tourism facilities in the regions with water defficiency. Nevertheless "the relatively low

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values of biodegradable organic matter and the nutrient imbalance (of greywater) limit the effectiveness of biological treatment" (Jefferson, 1999: quoted in [1]).

1.2. GREYWATER-TREATMENT TECHNOLOGIES

Depending on the final use of greywater, different treatment technologies are needed, and the treated water should "fulfil four criteria: hygienic safety, aesthetics, environmental tolerance and economical feasibility" [19]. Usually simple-treatment systems for the purpose of landscape irrigation, like sand/gravel filtration or settlement and flotation are used. If a treated water is used "in-house", e.g. for toilet flushing, a disinfection step should be added to remove microbial contaminants since their contact with people greatly increases in these applications [21].

The ambitious goal to implement denitrification in the treatment of low-loaded greywater can be justified, when other process applications are considered, e.g. the treatment of eutrophic surface water or the treatment of cruise ship wastewaters, to fulfil strict standards of their disposal.

The sequencing batch reactor (SBR) has proven to be a viable alternative to continuous-flow systems in carbon and nutrient removal from domestic and industrial wastewaters. While conventional aeration and settling occur simultaneously but in a spatial sequence in the continuous-flow systems, in SBR they are carried out in the same reactor but in temporal sequence. A submerged membrane sequencing batch reactor (SM-SBR) allows us to overcome the restriction of the simple SBR process and its dependence on sludge behaviour in the process of decantation. On a small footprint a hygienically acceptable, germ-free water is produced due to implementation of microfiltration membranes in the treatment process. Therefore, the membrane-filtered water is suitable for reuse, irrigation, and direct infiltration [13], [17]. The SM-SBR can be applied in hotels located in touristic areas suffering from water defficiency in order to save valuable drinking water due to the reuse of greywater from showers and everywhere else when an effluent quality close to drinking water standards is needed. Therefore the SM-SBR can be technically viable for application to greywater reuse [2], [13], [16].

2. METHODS

2.1. EXPERIMENTAL SETUP

2.1.1. SM-SBR PLANTS

In this study, two different systems are utilised for greywater treatment. A commercially available BioMir® reactor system manufactured by BUSSE Engineering consists of two 1 m^3 tanks, the first for storage and the second for biological treatment.

In the first phase (until 300 days), a submerged plate and frame module (A3 GmbH) were introduced into the biological chamber comprising 22 elements of a total membrane area of 3.9 m^2 (see figure 1B) and permeate was removed gravitationally. In the second phase, the membrane area was doubled by adding the second module, and permeate is now removed by a suction pump.

The other system is composed of a 27 dm^3 rectangular bioreactor (figure 1A) equipped with one submerged plate and frame module (A3 GmbH) comprising twelve elements of a total membrane area of 0.38 m^2. Permeate is removed using a peristaltic pump. The reactor volume was controlled by pressure transducers and together with the information from the probes (DO, pH and oxidation reduction potential, OPR) recorded directly on a computer.
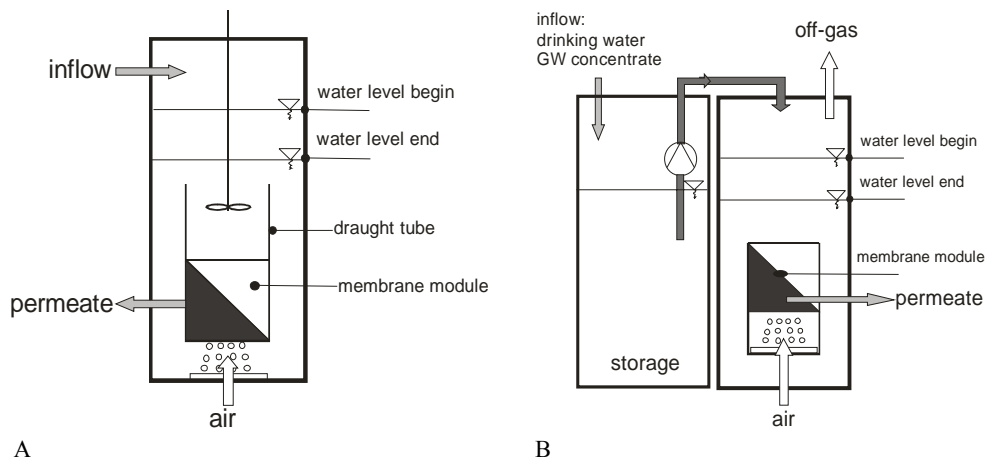

Fig. 1. Setup of both SM-SBRs (lab-scale reactor (A) and BUSSE reactor (B))

The time of solid retention can be assumed to be longer than 250 days for both systems, because since the start-up phase no biomass was taken out, except for sampling (see also table 1).

2.1.2. BATCH TEST

Batch tests were carried out in a 1 dm^3 stirred vessel to determine the ammonium utilisation rate (AUR) and the nitrate utilisation rate (NUR). In the batch test, a 90 min steering phase assures a defined starting point for all batches, a 120 min aeration phase (AUR), and 90 min sparging with nitrogen (NUR). A sample was taken every 20 min. In some batch runs, acetate was added as C-source to determine the AUR without substrate limitations.

2.2. RUN-MODE OF REACTOR

Both SM-SBRs worked under different process conditions obtained by: a) the changes in the ratio between the volume of wastewater that is filled and that discharged to the stationary volume (*VER* = volumetric exchange ratio), and (b) the changes in cycle time (t_c), including the variation of each phase involved. The cycle time was chosen according to values found in literature, where SBRs operated in cycle time varying from t_c = 3 h [22], t_c = 4 h [9], [10] and t_c = 6 h [4] to t_c = 12 h [5], [11].

In the lab-scale reactor, the time for each part of the cycle varied from 60 to 120 min for the anoxic phase and from 180 min to 270 min for the aeration phase. The filling took 2 min at the beginning of the anoxic phase. The *VER* was held in the range of 0.1 to 0.5. Experiments started from investigations of the membrane behaviour at *VER* = 0.5 and were followed by investigations at *VER* ranging from 0.3 to 0.2 to assess the treatment performance under changing conditions and to validate simulations based on microbial kinetics.

The BUSSE plant was operated at three constant settings. The first run was set at a cycle time of 12 h (240 min for the anoxic phase and 600 min for the aeration phase). Due to very long hydraulic retention time (*HRT*), the cycle time was reduced by half to t_c = 6 h (180 min for the anoxic phase and 180 min for the aeration phase) during the run 2. After introducing the second membrane module into the reactor, the cycle time was shortened in the run 3 to t_c = 4 h, reducing the anoxic phase to 90 min and the aeration phase to 150 min. The *VER* was held constant at 0.12 in all set-ups.

Table 1

Process operating conditions

Parameter	Lab-scale reactor	BUSSE reactor
Solids retention time (*SRT*), d	> 250	> 250
Hydraulic retention time (*HRT*), h	13–60	33–100
$MLSS_{reactor}^1$, mg/dm^3	1500–3500 (start-up) 1500–4500	800–1200 (start-up) 2000–3100
$MLVSS_{reactor}^2$, %MLSS	65–70	60–70
Average *F/M* ratio^3, g COD_{in} (g $MLVSS_{reactor}$ d)$^{-1}$	0.124	0.074
Volumetric exchange ratio (*VER*)	0.1–0.5	0.12
Flux, dm^3 (hm^2)$^{-1}$	7.0–13.0	8.5–12.5
TMP, bar	0.1–0.4	0.1–0.15

[1] Mixed liquid suspended solids.
[2] Mixed liquid volatile suspended solids.
[3] Food to microorganism ratio.

The biomass was fed with synthetic greywater (see table 2) adapted from literature [12], [15] on the one hand to represent the greywater of a 4-person household and on the other hand to investigate the performance of denitrification with water of low carbon concentration and to transfer the results to other application areas like treatment of surface water. The greywater masterbatch was diluted in a separated tank to a COD approaching 200 mg/dm^3, a typical value of that parameter [8]. In order to evaluate the biological performance, detailed cycle analyses were carried out. All samples were filtered through a cellulose acetate filter (pore size of 0.2 μm, Sartorius), before measuring COD, TN, NO_3–N, NO_2–N, NH_4–N and PO_3–P according to standard methods.

T a b l e 2

Recipe of synthetic greywater masterbatch and feed concentration

Ingredients	Approx. daily amount	Dilution up to 5 dm^3 with tap water	Feed concentration (mg/dm^3) or (cm^3/dm^3)
Tooth paste	1.2 g	84 g	21
Shower gel	10 cm^3	700 cm^3	0.175
Cleaner	60 cm^3	1200 cm^3	0.3
Shower oil	5 cm^3	100 cm^3	0.025
Shampoo	5 cm^3	100 cm^3	0.025
Bubble bath	25 cm^3	500 cm^3	0.125
Urea	6.0 g	120 g	30
Na_2CO_3	5.5 g	110 g	27.5
K_2HPO_4	500 mg	10 g	2.5
NH_4Cl	3.5 g	70 g	17.5
COD			200 mg O_2/dm^3

3. RESULTS AND DISCUSSION

Both reactors worked continuously for 350 days allowing investigations of COD and nitrogen removal as well as critical flux measurements of the membrane module. Especially in the lab-scale reactor, a variety of experiments were carried out, starting with critical flux investigation followed by frequent changes of *VER* and t_c. This was needed to validate mathematical models; hence the removal efficiency and the biomass concentration fluctuated widely.

3.1. DEVELOPMENT OF BIOMASS

Both reactors were inoculated with activated sludge from the WWTP Ruhleben in Berlin for start-up. The lab-scale reactor was completely filled with that sludge, whereas the BUSSE reactor was filled with 300 dm^3 of sludge and tap water to 1000 dm^3 volume. This explains a low concentration of organic compounds, measured as MLVSS, in the BUSSE reactor at the beginning (see figure 2B).

Due to foaming problems with synthetic greywater during start-up, the unquantified amounts of biomass were lost in the lab-scale reactor. The MLVSS concentration decreased from 2500 mg dm^{-3} to 2000 mg dm^{-3} (see figure 2A). To minimise foaming, the SM-SBR was operated as a continuous flow reactor for one week. Due to different experimental set-ups for investigation of critical flux of the membrane unit in the start-up phase with numerous samplings, the biomass declined further to the value of 850 mg dm^{-3}. The lab-scale reactor showed a biomass increase in the periods of the following days: 79…115, 130…230, and 295…340. The slopes in these phases are in the range of 22…27 mg MLVSS (dm^3 d)$^{-1}$, leading to the average growth rates $\mu = 0.007…0.015$ d^{-1} in these phases.

A VSS-concentration in the BUSSE reactor after 160 days of its operation was far more stable. From the 160th to the 210th days the cycle time was set at t_c = 12 h and changed to t_c = 6 h until the 300th day of operation. It has to be pointed out that although the cycle time was divided by 2, no significant growth of biomass could be seen and MLVSS reached an average value of 1300 mg dm^{-3}. This means that the cycle time was still too long and bacteria lived under maintenance conditions. The last change consisted in a further limitation of the cycle time to t_c = 4 h and resulted in an increase of MLVSS to more than 2200 mg dm^{-3}.
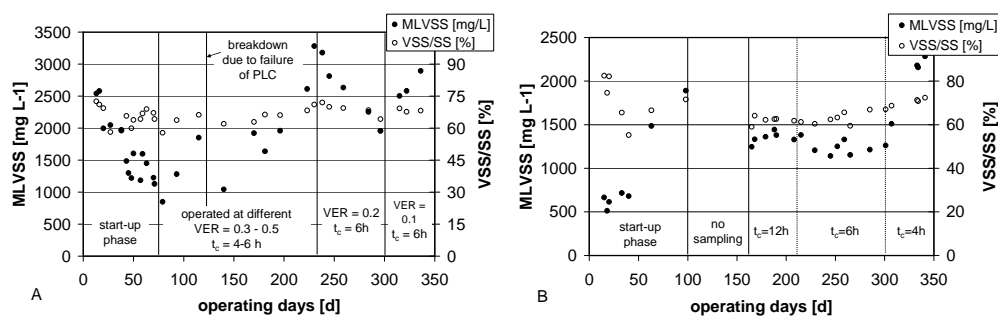

Fig. 2. Biomass growth: lab-scale reactor (A); BUSSE reactor (B)
(VSS/SS – volatile suspended solids/suspended solids, PLC – programmable logic controller)

The *F*/*M* (food to microorganism) ratio is used for the design and operation of the activated sludge process. The values found in literature range from 0.01 to 0.3 [9]. To gain a deeper insight into the process it is appropriate to find the relationship between

MLSS concentration and *F/M* ratio. Figure 3 illustrates a simplified dependence of the substrate concentration on time over one cycle. It is known from the maintenance concept that:

$$\mu = Y \frac{r_{su}}{X} - k_d ,$$ (1)

where:

μ – the specific growth rate,
X – the biomass concentration.

The specific substrate utilisation rate U can be estimated by:

$$U = \frac{r_{su}}{\overline{X}} = \frac{S_0 - S_{end}}{\tau \cdot \overline{X}} ,$$ (2)

where:

S_0 – the substrate concentration in the reactor at the start of a cycle,
S_{end} – the substrate concentration in the reactor at the end of a cycle,
τ – the time of substrate utilisation,
$\overline{X}$ – an average biomass concentration,
Y – the biomass yield,
r_{su} – the rate of substrate utilisation,
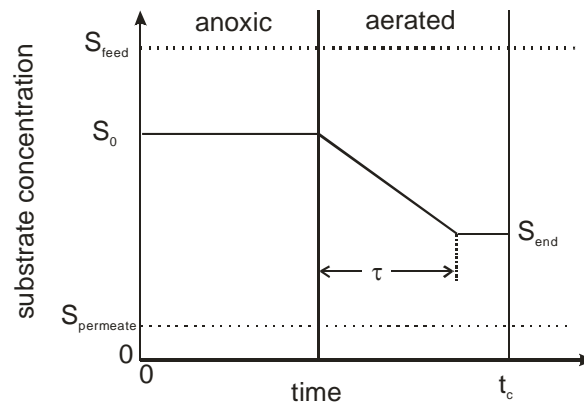k_d – the endogeneous decay coefficient.

Fig. 3. General scheme of substrate utilisation in an SBR cycle

Equation (2) does not permit us to evaluate so easily the role of the process engineering parameters. The relationship between MLSS concentration and *F/M* ratio in an SBR can be derived applying two simplifications.

Firstly, the time τ will be expressed by the cycle time t_c, where U now becomes the mean substrate utilisation rate $\overline{U}$ over the total cycle:

$$\overline{U} = \frac{S_0 - S_{end}}{t_c \overline{X}} . \tag{3}$$

Secondly, a specific removal rate can be established by multiplying the F/M ratio by the removal efficiency E. The specific removal rate can be expressed by:

$$\frac{F}{M} E = \frac{S_{feed}}{\overline{X} \cdot HRT} \cdot \frac{S_{feed} - S_{permeate}}{S_{feed}} , \tag{4a}$$

$$\frac{F}{M} E = \frac{S_{feed} - S_{permeate}}{t_c \overline{X}} \cdot VER , \tag{4b}$$

where:
    S_{feed} – the substrate concentration in the feed,
    $S_{permeate}$ – the substrate concentration in the permeate,
    HRT – a hydraulic retention time,
    VER – a volumetric exchange ratio.
Based on the a mass balance the starting concentration S_0 of a cycle can be defined by:

$$S_0 = S_{feed} \cdot VER + S_{end} , \tag{5}$$

where S_{end} is the concentration in the previous cycle. This yields:

$$\frac{F}{M} E = \frac{S_0 - S_{end}}{t_c \overline{X}} - \frac{S_{permeate}}{t_c \overline{X}} \cdot VER . \tag{6}$$

Taking into account that (compare figures 3 and 6):

$$S_0 - S_{end} \gg S_{permeate} \cdot VER , \tag{7}$$

equation (6) yields:

$$\frac{F}{M} E \approx \frac{\overline{r}_{su}}{\overline{X}} = \overline{U} . \tag{8}$$

Combining equation (8) with equation (1) that assumes an average specific growth rate $\overline{\mu}$ we arrive at:

$$\overline{\mu} = Y \frac{F}{M} E - k_d . \tag{9}$$

As μ is defined as $\mu = r_X X^{-1}$ (where r_X denotes the growth rate) we can write the final relation:

$$\overline{X} = \frac{\overline{r}_X}{Y \dfrac{F}{M} E - k_d} \tag{10}$$

with $\overline{r}_X$ being constant in one cycle.

For further calculations the following values were used: $Y_{B/S}^g$ = 0.36 mg VSS (mg COD)$^{-1}$ [14] ($Y_{B/S}^g$ denotes a true biomass yield on an available substrate), k_d = 0.006 g VSS (g VSS d)$^{-1}$ (reduced by the factor of 10^{-1} compared to [18]) and an average $\overline{r}_X$ = 24.5 mg (dm^3 d)$^{-1}$ (lab-scale reactor) or an average $\overline{r}_X$ = 5 mg (dm^3 d)$^{-1}$ (BUSSE reactor, compare figure 2).
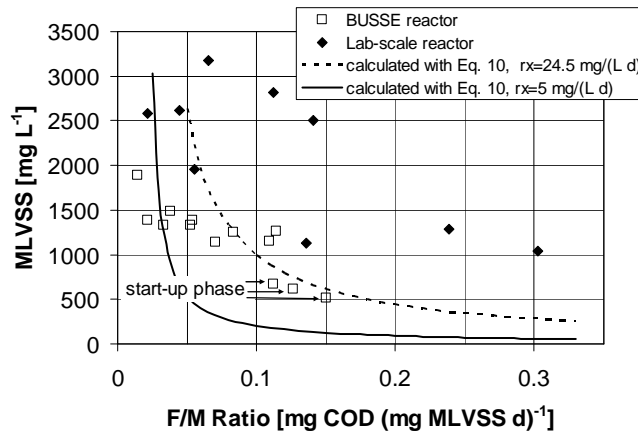

Fig. 4. Biomass versus *F*/*M* ratio

Figure 4 shows no significant trend in the lab-scale reactor. This may be explained by many variations in the operational settings. The BUSSE reactor showed a nearly constant biomass concentration at different low *F*/*M* ratios, if the three samples from the start-up phase are not taken into consideration. The curves calculated do not match the results from both reactors. This may be due to the *Y* and k_d chosen from literature. The yield might be lower because of the nature of slow-biodegrading synthetic greywater.

3.2. NUTRIENT REMOVAL

Table 3 shows the removal rates for COD, TN and NH$_4$–N in both reactors as well as the average concentrations of feed and permeate. The variation in feed concentrations can be explained by the existence of a storage tank; different in size at both plants it leads to different *HRT* during the storage and hence a biodegradation of COD at first.

The lab-scale plant achieved a COD removal of 79% on average, 37% for TN, and 99% for NH_4–N. As mentioned before due to several different cycle adjustments the performance fluctuated significantly. Especially TN removal varied widely between 20% and 80%, obtaining its best performance at a cycle time of 4.5 h (90 min anoxic phase, 180 min aeration phase) with a *VER* = 0.3.

The BUSSE plant achieved a COD removal of 89% on average, 41% for TN and 98% for NH_4–N. A low TN removal in general can be attributed to a low feed C:N ratio and a relatively high *HRT*, which resulted in a carbon deficiency throughout the cycle.

Table 3

Average concentrations in feed and permeate and removal efficiencies of both plants

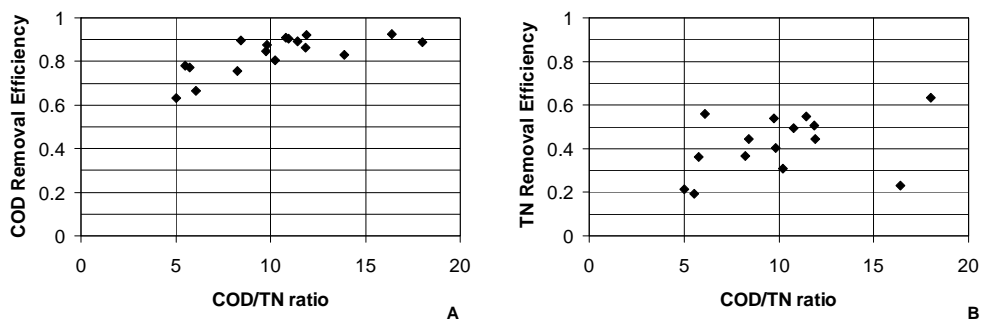| Parameters	Lab-scale plant			BUSSE plant		
	Feed (mg dm^{-3})	Permeate (mg dm^{-3})	Removal efficiency	Feed (mg dm^{-3})	Permeate (mg dm^{-3})	Removal efficiency
COD	157	31.7	0.79	192	21.0	0.89
TN	19.4	12.3	0.37	16.3	9.7	0.41
NH_4–N	15.7	0.39	0.99	11.2	0.21	0.98
NO_3–N	0.10	10.5	/	0.20	5.9	/
PO_4–P	0.46	0.23	0.50	0.53	0.15	0.71

Fig. 5. COD removal efficiency (A) and TN removal efficiency (B) versus the COD/TN ratio

Kargi and Uygur [11] designed a statistical experiment to determine the effect of nutrient composition of media on the removal efficiency of nutrient and carbon. Removal of nutrients varied depending on their availability. The COD/NH_4–N/PO_4–P ratio at the maximum efficiency was equal to 100/2/0.54 [11]. The COD/NH_4–N/PO_4–P ratio in this study ranged from 100/(3.4 to 14.5)/(0.28 to 0.53), which means too much ammonia and too little orthophosphates. The efficiency of COD removal will probably not exceed 95% due to the nature of synthetic greywater, where at least 5% of components are not degradable (compare also figure 5A). In figure 5, the results of the TN- and

COD-removal efficiency at COD/TN ratios are compared. So far it can be stated that at a COD/TN ratio between 10 and 25 a nearly constant COD removal of 90% is reached, whereas the TN removal efficiency does not allow any prediction.
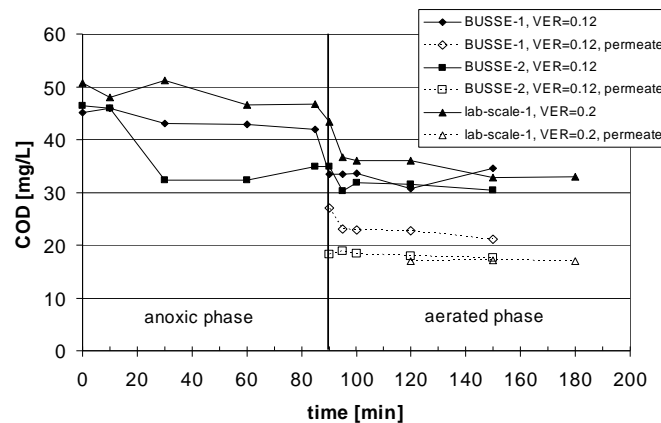

Fig. 6. Time-dependent COD concentrations

Figure 6 shows the COD removal for three different cycle analyses. The difference in the COD concentration in the reactor and in the permeate is significant. One of the explanations could be the presence of EPS (extracellular polymer substance), polysaccharide (PS), and proteins, which are known to be partly retained by the membrane or cake layer on the membrane surface. The concentration of soluble PS in the reactor determined by means of the Dubois method [7] approaches 10 mg of glucose per 1 dm^3, which is equivalent to a COD of 10.7 mg of O$_2$ per 1 dm^3. The difference in the COD concentration in the reactor and in the permeate could be explained by the COD equivalent of the soluble PS.

3.3. DETERMINATION OF NUR AND AUR

The denitrification potential is one of the key parameters in the design of activated sludge systems for biological nutrient removal. The results of the nitrate utilisation rate (NUR) can be used to assess the denitrification potential of the operating system and hence the resulting nitrogen removal efficiency. It has been observed that the carbon compounds of synthetic greywater cannot be easily used for complete denitrification. In this study, the NUR with acetate dosing was almost six times above the endogeneous NUR (see table 4). Obtaining the same ratio of endogeneous to exogeneous denitrification BUCKLEY [6] found denitrification rates for slowly biodegradable COD between 1.0 and 1.5 mg N g^{-1} MLVSS h^{-1}, whereas Kujawa and Klapwijk (quoted in

[14]) found even lower rates between 0.2 and 0.6 mg N g^{-1} MLVSS h^{-1} for endogeneous denitrification. To optimise the denitrification process in greywater treatment either an external carbon source can be necessary or the anoxic phase should be lengthened until the change in nitrate concentration approaches zero.

The ammonium utilisation rate (AUR) is measured as the nitrite and nitrate concentration obtained during aeration. In this study, nitrification rates from 2.18 to 2.62 mg NO_3–N (g MLVSS h^{-1}) were measured, exceeding the results reported in literature. The specific nitrification rates range from 0.78–1.81 mg NO_3–N (g SS h^{-1}) for synthetic wastewater (Muller, 1995, quoted in [14]) to 1.7–2.0 mg NO_3–N (g VSS h^{-1}) for municipal wastewater (Fan et al., 2000, quoted in [14]). Even the AUR during normal operation of the lab-scale plant with the average of 1.27 mg NO_3–N g^{-1} MLVSS h^{-1} is comparable to values found in literature [9].

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Average nitrate utilisation rates (NUR) and ammonium utilisation rates (AUR)

	Nitrification rate (AUR) [mg NO_3–N g^{-1} MLVSS h^{-1}]			Denitrification rate (NUR) [mg NO_3–N g^{-1} MLVSS h^{-1}]			
	Batch		Operation	Batch			Operation
	BUSSE	Lab-scale	Lab-scale SM-SBR	BUSSE	Lab-scale	BUSSE no acetate dosing	Lab-scale SM-SBR
Average	2.19	2.62	1.27	3.55	3.44	0.62	0.57
Max	2.79	3.14	1.36	5.67	4.95	0.79	0.60
Min	1.66	2.10	1.18	2.196	1.92	0.37	0.54

4. CONCLUSIONS

Synthetic greywater was treated in a submerged membrane sequencing batch reactor with anoxic and aerobic phases for denitrification and nitrification. COD removal efficiency was between 79% and 89%, but has not been optimised yet. Low loading rates resulted in small biomass concentration as well as in weak biomass growth. The nitrate utilisation rate and ammonium utilisation rate values obtained are comparable to data found in literature, but total nitrogen removal can be efficient, provided that carbon source will be added to the system.

Beiersdorf AG in donating the main ingredients for the synthetic greywater.

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