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LOW PRESSURE OPERATED MEMBRANE BIOREACTORS FOR DECENTRALIZED GREY WATER RECYCLING

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KURZFASSUNG

Die Produktion und Verfügbarkeit von Lebensmitteln auf der Erde sind, wie allgemein bekannt, äußert ungleichmäßig verteilt. Das lebensnotwenigste Grundnahrungsmittel, nämlich Wasser, stellt dabei leider keine Ausnahme dar. Während es in vielen Gebieten auf der Erde saisonal zu Wasserknappheit kommt bzw. es überhaupt nur wenig nutzbares Trinkwasser gibt, steigen in den gut versorgten Industriestaaten kontinuierlich die Kosten für Aufbereitung, Logistik und Abwasserbehandlung. Die Lösung der dahinterstehenden Problematiken und Zusammenhänge, muss ein zentraler Fokus der Politik und Siedlungswasserwirtschaft im kommenden Jahrhundert sein.

Diese Arbeit darf hierzu als ein kleiner Mosaikstein zu diesen Lösungsstrategien angesehen werden. Der zentrale Inhalt ist die mehrfache Nutzung von Wasser, also Wasserrecycling. Im Speziellen wird die Anwendung der Aufbereitung von Grauwasser (Abwasser aus Dusche, Waschbecken, Waschmaschine) zu Brauchwasser für die neuerliche Nutzung im Haushalt untersucht und diskutiert. Es zeigte sich, dass mit dem derzeitigen Stand der Technik in der Aufbereitung (Jahr 2013) sich diese Anlagen noch nicht rechnen und dadurch generell noch sehr selten eingesetzt werden. Dabei war nicht nur die Ablaufqualität und Akzeptanz das Hindernis für die Anwendung, sondern vor allem die relativ hohen Betriebskosten und Prozessunsicherheit. Daher wurden auf Grundlage der bewährten Membranbioreaktor Technologie zwei Alternativprozesse für die Anwendung im häuslichen Bereich entwickelt, untersucht und dokumentiert. Der erste vorgestellte Prozess beinhaltet die Kombination von Biofilm Aufwuchskörpern mit einer nachgeschalteten Membranfiltration. Der zweite Prozess nutzt keine Aufwuchskörper, jedoch eine Hohlfasermembran mit hoher spezifischer Oberfläche als kombinierte Aufwuchsfläche für Biofilm und Membranfiltration. Eine zusätzliche Belebtschlammstufe war in beiden Fällen nicht notwendig. Das Ziel war es jeweils, die Qualität hinsichtlich internationaler Richtlinien für Brauchwasser (z.B. NSF oder fbr) zu erreichen. Bei der Untersuchung und Optimierung dieser Prozesse wurde das Hauptaugenmerk auf die Reduktion der Investition, Betriebskosten und Wartung gelegt. Dadurch konnte in beiden Fällen bei gleichbleibender hoher Ablaufqualität der Strombedarf für diese Kleinstanlagen unter 1,5 kWh/m³ gereinigtem Wasser gesenkt werden. Neben der genauen Prozessdokumentation wurden die Betriebsdaten und Abbaukurven der ersten Anlage dazu genutzt, ein allgemeines mathematisch-mechanistisches Modell zu erstellen.

Dadurch ist es auch in Zukunft möglich, allgemeine Ansätze für die Dimensionierung abzuleiten und Vorhersagen von Ablaufdaten bei diesen Kleinanlagen zu treffen.

Die Ergebnisse der Untersuchungen sind vielversprechend und bieten die Grundlage für weitere Forschungen und Optimierungen. Die beschriebenen Prozesse haben das Potenzial, weite Verbreitung zu finden und dadurch ihren Teil zur Ressourcenschonung und Versorgungssicherheit beizutragen.

Schlagworte

Grauwasser, Membranbioreaktor, Biofilm, Modellierung, Wirtschaftlichkeit, Betriebskosten

ABSTRACT

It is common knowledge that the production and availability of food in the world is most unevenly distributed. Unfortunately, this holds also true for the most important foodstuff i.e. water. While there is seasonal water scarcity in many regions of the world and only little potable water available respectively, the costs for water treatment, logistics and wastewater treatment continue to increase in the industrial nations with good water supply. Hence, the focus of politics and water management in the next century must be on the solution of the underlying problems and their interdependencies.

In this context this thesis may be regarded as a small contribution to these solution strategies by focussing on multiple use of water i.e. water recycling. In detail, the processes for the production of service water from grey water (shower, washbasin, washing machine) for reuse in households are analysed and discussed. The analyses have shown that state-of-the-art processes (year 2013) are not profitable yet and therefore in general rarely applied. The reason for this does not only lie in effluent quality and acceptance, but much more in relatively high operation costs and process uncertainties. Therefore, based on the proven membrane bioreactor technology, two alternative processes for the application in households were developed, analysed, and documented. The first process described is a combination of biofilm carrier material with downstream membrane filtration. The second does not use any carrier material, but a hollow fibre membrane with high specific surface as combined growth area for biofilm and membrane filtration. An additional activated sludge stage was not necessary in either case. In both cases, the aim was to reach the quality requested in international guidelines for service water e.g. NSF or fbr. In the analyses and optimization of these processes the focus was on the reduction of investment, operation and maintenance costs. In this respect, it was possible to reduce the power consumption of both small plants to less than 1.5 kWh/m^3 treated water at constantly high effluent quality.

In addition to the detailed process documentation the operating data and degradation curves of the first plant described were used to elaborate a general mathematical-mechanistic model, with which it will be possible to identify general approaches for dimensioning and predict the effluent data of these small plants in the future.

The results of the analyses are promising and may serve as the basis for further research work and optimization. The processes described have the potential to become widely disseminated and by this contribute to resource conservation and supply security.

Keywords

Grey water, membrane bioreactor, biofilm process, modelling, feasibility, operating costs

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ABBREVIATIONS AND SYMBOLS

A	bioreactor tank area [m^2]
AS-MBR	Activated sludge membrane bioreactor
BF-MBR	Biofilm membrane bioreactor
b_H	Endogenous decay coefficient [1 d^{-1}]
BOD_5	Biological oxygen demand
c_A	concentration of material A in bioreactor [mg L^{-1}]
COD	Chemical oxygen demand
dt	numeric time step [h]
$dT_{B,conv}$	temperature change of bioreactor by natural convection [°C]
dT_B	temperature change of bioreactor [°C]
EPS	Extracellular polymeric substances
f_{ES}	Soluble fraction of endogenous residues
f_{EX}	Particulate fraction of endogenous residues
F/M	Food/Microorganism (Biomass)
GW	Grey water
HRT	Hydraulic retention time
i_{SN}	Nitrogen fraction of soluble microbial products [gN g^{-1} COD]
i_{SP}	Phosphorous 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 phosphorous per mass of COD in biomass [gN g^{-1} COD]
i_{XN}	Nitrogen fraction of particulate microbial products [gN g^{-1} COD]
i_{XP}	Phosphorous fraction of particulate microbial products [gP g^{-1} COD]
k_a	Aeration rate of bioreactor [-]
k_h	Maximum hydrolysis rate of XS [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 phosphorous

K_S	Half saturation coefficient for readily biodegradable substrate
K_X	Half saturation coefficient of hydrolyses
LMH	Flux in L/m^2h
MBBMR	Moving bed biofilm membrane reactor
MBR	Membrane bioreactor
m_i	Measured value of the output variable
MLSS	Mixed liquor suspended solids
NH_4-N	Ammonia
NO_3-N	Nitrate
ORL	Organic loading rate
PH_4-P	Phosphate
p_i	Predicted value of the output variable
PSD	Particle size distribution
Q_A	Aeration [L air h^{-1}]
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}]
RBC	Rotating biological contactor
S_{EF}	Effluent COD [mg L^{-1}]
S_I	Inert soluble COD fraction in the influent [mg L^{-1}]
S_{MB}	Rejected fraction of the soluble residual COD [mg L^{-1}]
SMP	Soluble microbial products
S_O	Dissolved oxygen [mg L^{-1}]
S_P	Soluble microbial products [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}]
SRT	Sludge retention time (sludge age)
S_S	Readily biodegradable COD [mg L^{-1}]
T	Temperature
T_B	Bioreactor temperature [°C]
TEP	Transparent exopolymer particles
T_{in}	Inflow temperature [°C]

TN	Total nitrogen
TP	Total phosphorous
T_R	Ambient temperature [°C]
TSS	Total suspended solids
μ_H	Maximum growth rate of heterotrophs [h^{-1}]
V_B	Bioreactor volume [L]
w_S	sedimentation velocity [m h^{-1}]
WW	Waste water
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_{O2}	O_2 transfer coefficient diffuser [mg O2 L^{-1} air]

INTRODUCTION

1. INTRODUCTION

1.1. BACKGROUND

Water, principally the most important foodstuff for life, is still not available in sufficient quantities for all of the more than seven billion people on Earth. Water availability and demand are rather imbalanced around the globe. There are poor regions with latent water shortages, emerging economies with strongly increasing water demand and finally a more and more declining number of industrialized countries, which seem to have a deceptive spill-over. One assumption is most probably valid for all of them, namely that prices for drinking water, supply and wastewater treatment will rise even further.

One way of breaking the vicious circle of latent water shortages or rising prices could be water recycling. After all, wastewater is only used water, which could theoretically be reprocessed to any kind of quality necessary for industry, farming and even potable purposes. Technologies are available, but their dissemination is still in its infancy. The focus of this thesis is on a small, but in total also significant part of wastewater i.e. grey water (GW).

1.2. GREY WATER

GW is household wastewater, which consists of used water from washing basins, bathtubs, showers and washing machines. Sometimes kitchen wastewater is also included. Toilet flushing water (black water) is definitely excluded. Huge parts of daily water consumption of roughly 50 – 80 % fall into this definition. Therefore, the water recycling and saving potential is of special interest and has recently come into the focus of industry. Compared to rain water harvesting technologies there is also one major advantage: GW production and recycling are independent of weather and season. The production and reuse possibilities are usually in direct connection with consumer habits and activity. Long storage of treated water is not necessary. Finally, this will also avoid storage problems which occur especially in hot climates. Rain water and particularly not sufficient treated GW is often subject to further degradation processes and an ideal breeding ground for mosquitoes (Christova-Boal *et al.*, 1996). As a consequence, fast reuse and short intermediate storage is preferable in all reuse applications.

Table 1: Break-down of household water quantities for different countries (extended from Boyjoo *et al.*, 2013)

Domestic water use (%)	Denmark[a]	Australia[b]	UK[c]	India[d]	US[e]	Oman[f]
Bathroom	36	26	28	39	18	47
Laundry	14	15	12	20	22	7
Kitchen (incl. drinking)	21	5	19	23	17	37
Toilet flushing	23	34	35	15	27	4
Other (garden, cleaning)	6	20	6	3	16	5

a: Revitt *et al.*, 2011
b: Christova-Boal *et al.*, 1996
c: Pidou *et al.*, 2007
d: Mandal *et al.*, 2011
e: Sheikh, 2010
f: Prathapar *et al.*, 2005

Despite of the same GW sources, the characteristics and flows vary strongly due to individual personal habits and personal activity. There is also a clear regional and cultural impact (**Table 1 and Table 2**). Load and concentration of pollutants are also dependent on household income and population structure. Worthy of a mention is that the Northern European countries, like Germany, Sweden, Norway or Denmark produce less GW than Southern European countries because of a special water saving mentality (Boyjoo *et al.*, 2013) which is probably also a matter of local water costs. The strong variation of characteristics reports especially Eriksson *et al.* (2002), who found that there could be more than 900 different trace substances in GW. Finally, these substances could affect treatment and further reuse. In addition, bacterial contamination washed off from human skin cannot be neglected. Coliforms could be measured even with values up to 7.0×10^8 per 100 ml (Eriksson *et al.*, 2002). In Australia, a study confirmed that 21% of GW sample from 93 typical households contained pathogens (O'Toole *et al.*, 2012). An outline of general GW qualities measured in different countries from a literature review is listed in **Table 2**. Treatability and especially biological degradability need to be considered if treatment will be with activated sludge or biofilm processes. Based on the strong impact of personal habits, there is no common conclusion in literature so far. There is the tendency that biodegradability in terms of BOD/COD ratio related to nitrogen (N) or phosphorous (P) nutrients is limited when only GW from hand basins and bathrooms are combined (Jefferson *et al.*, 2002; Chaillou *et al.*, 2011) while it is in an acceptable range if also kitchen and washing machine effluents are connected (Nolde, 1999; Bullermann et al., 2002).

Table 2: GW quality parameters of different publications and guidelines

	Germany[a]	**UK**[b]	**USA**[c]	**Australia**[d]	**India**[e]
pH [-]	7.9 / - / 7.5	7.47	6.8	7.5	7.7
BOD_5 [mg/l]	111 / 188 / 360	155	162	160	170
COD [mg/l]	225 / 340 / 535	451	366	-	-
SS [mg/l]	40 / - / -	100	162	115	190
TN [mg/l]	10 / - / 13	8.73	-	12	12
TP [mg/l]	1.5 / - / 5.4	0.35	-	8	12
Turbidity [NTU]		100.6	-	100	161
Total coliforms [CFU/100 ml]	10^5 / 10^6 / 10^6	7.4×10^3	2.4×10^6	-	-
Faecal coliforms [CFU/100 ml]	10^4 / 10^5 / 10^6	2.0×10^3	1.4×10^6	10^6	-

a: Nolde, 1999, Bullermann *et al.* 2001 - Hand basin, shower, baths / washing machine / kitchen
b: Jefferson *et al.*, 2004 - Hand basin, shower, baths
c: Brandes, 1978 – mixed GW
d: Jeppersen & Solley, 1994,– mixed GW
e: National Environmental Engineering Research Institute, 2007 – mixed GW

1.3. GREY WATER REUSE

There is a wide range where treated GW can be reasonably reused. The saving potential lies between 30 – 120 L/day and person depending on household income and region (Morel & Diener, 2006). Typical applications are agricultural irrigation, washing, cooling and gardening but there are also more sensitive possibilities in-house i.e. toilet flushing or washing machine. This is why the requirements on the treated water diversify. The higher the possibility of direct human contact, the more sophisticated the treatment method and the better the treated water quality should be. A general international guideline covering all reuse applications is pending. As a result, many countries have developed their own guidelines which show partially big differences. Some standards of countries and organizations for different applications are listed in **Table 3**.

Table 3: GW guidelines of selected organizations and countries

	Germany[a]	UK[b]	USA[c]	Australia[d]	India[e]	EU[f]	WHO[g]
pH	-	5 – 9.5	6 – 9	6.5 – 8.5	-	6.5 – 9.5	-
BOD_5 or BOD_7 [mg/l]	< 5	-	< 10	< 10	-	< 5	-
COD [mg/l]	-	-	-	-	-	-	-
SS [mg/l]	-	visually clear, free from floating debris	-	< 10	<30	-	-
TN [mg/l]	-	-	-	-	-	NH_4-N 0.5, NO_3-N 50, NO_2-N 0.5	-
TP [mg/l]	-	-			-	-	-
Turbidity [NTU]	-	< 10	< 2	< 2 (5 max.)	-	-	-
Oxygen saturation	> 50 %	-		-	-	-	-
Total coliforms [CFU]	< 100/ml	10/100ml		-	-	< 100/ml	-
Faecal coliforms [CFU]	< 10/ml	-		-	-	< 10/ml	-
E.coli [CFU]		n.d.	<14/100 ml (any) 0 (90%)	<1/100ml	<10/100ml	-	< 10^3/100ml
Salmonella [CFU]	n.d.	-	-	-	-	-	-

	Germany[a]	UK[b]	USA[c]	Australia[d]	India[e]	EU[f]	WHO[g]
Nematodes [CFU]	n.d.	-	-	-	-	-	< 1/1000 ml
Pseudom. aerug. [CFU]	<1/ml	-	-	-	-	< 1/ml	-
Coliphage [PFU]	-	-	-	< 1 /100ml	-	-	-
Clostridia [CFU]	-		-	< 1 /100 ml	-	-	-
Intestinal (enterococci)	-	10/100 ml	-	-	-	-	-
Legionella pneumophila	-	n.d.	-	-	-	-	-
Colour	-	Colourless	-	-	-	-	-
Final disinfection	-	Cl	Cl	Cl, UV	-	-	-

a: fBr Hinweisblatt H 201, 2005 - for non-restricted non-potable urban reuse
b: Environment Agency, 2011, based on BS 8525- for non-restricted non-potable urban reuse
c: US EPA Guidelines for Water Reuse, 2004 - for non-restricted non-potable urban reuse
d: Guidelines for the Non-potable Uses of Recycled Water in Western Australia, 2011 - for non-restricted non-potable urban reuse
e: National Environmental Engineering Research Institute, 2007 - Non restricted agricultural reuse
f: EU-bathing water directive 2006/7/EG - for non-restricted non-potable urban reuse
g: WHO, 2006 - Non restricted agricultural reuse
n.d.: not detectable

There is i.e. the WHO (2006) guideline referring to agricultural purposes for irrigation only. Germany is one of the market leaders of grey and rain water reuse systems but still with pending legal guidelines. Anyhow, the most popular private recommendation in German speaking countries is from the German Association for Rainwater Harvesting and Water Utilisation (2005) and suggests rather low limits in terms of biodegradable pollutants. The American, Australian and British standards distinguish also between different risks levels depending on human contact and further recommend final disinfection. Thus, all mentioned guidelines are a logic consequence of the already mentioned high regrowth potential of GW and nuisance problems during storage and reuse. While in German speaking countries in-house usage of chlorine is frowned upon and therefore in the treatment process the basis of regrowth (BOD_5) shall be retained, in English speaking countries final disinfection by chlorine or UV is accepted and therefore higher substrate concentration in the effluent are allowed. Advanced biological treatment with rotating biological contactor (RBC) showed that one to three magnitudes of indicator bacteria (faecal coliforms, heterotrophic plate count) and specific pathogens (*Pseudomonas aeruginosa sp.*, *Staphylococcus aureus sp.*) will be reduced. Anyhow, final disinfection was still mandatory to achieve unrestricted urban reuse guidelines. Chlorine in a concentration of 0.5 – 1 mg/L prevented regrowth but only partially inactivated existing microorganism. Regrowth of the different species were also not affected equally by UV radiation. UV seems to be only efficient in inactivating the microorganism (Friedler *et al.*, 2013) but cannot be used as net protection in case of long retention time in pipes and storages. In summary, GW is a highly variable and complex waste water and need specially adapted and flexible treatment processes in order fulfil all requirement.

1.4. GREY WATER TREATMENT

The treatment method of GW is always a matter of the reuse application. The easiest way would be of course direct reuse. Direct recycling is common for agricultural purposes or soil watering but even for that coarse filtration would be economically in order to avoid clogging of irrigation system. In literature a comprehensive list of physical, chemical, biological treatment or a combination of different processes was reported (**Table 4**). Most of the presented technologies already achieve the lower criteria of agriculture reuse. In terms of unrestricted non-potable urban reuse additionally high limits for microbiological contamination need to be considered. In this regard, especially rotating biological contactor (RBC) with final disinfection or membrane bioreactors (MBR) seems to be a technically acceptable solution. Nonetheless, in consideration of investment and amortisation of on-site GW treatment, payback times are particularly for small applications still too long and often not feasible (Nolde, 2005; Friedler & Hadari, 2006; Humeau *et al.* 2011).

With the membrane bioreactor technology a process was found which combines high effluent quality with process stability (Lesjean & Gnirss, 2006; Merz *et al.*, 2007; Kraume *et al.*, 2009, Huelgas & Funamizu, 2010; Jong *et al.*, 2010) but still has significant disadvantages concerning high energy consumption, maintenance by chemical cleaning and also long payback times (Friedler & Hadari, 2006; Humeau *et al.*, 2011; Kraume *et al.*, 2010 and Merz *et al.*, 2007).

1.5. MBR TECHNOLOGY

In a common definition the membrane bioreactor (MBR) technology combines biological waste water treatment with micro- or ultrafiltration in one process. Generally, the biological treatment is carried out as an activated sludge bioreactor. The membrane filtration separates activated sludge from treated water and is available in mainly two configurations. In the submersed configuration the membrane unit is situated inside the bioreactor and permeate is produced by slight vacuum or gravity. In the external configuration the membranes are placed in pressurized tubes outside the bioreactor tank and are operated in cross-flow mode. In both configurations the conventional final clarifiers can be omitted. Without final clarifiers the loading rate of the bioreactor can be increased and the footprint of the plant will be reduced. The physical barrier of membrane filtration, which is usually in a range of 0.01 – 1 µm, safely removes biomass as well as all particulate and colloidal matter. The independent control of sludge retention time (SRT) and hydraulic retention time (HRT) leads to an improved biodegradation. Independent of the actual condition of activated sludge, the membranes will retain complete biomass and thereby provide a certain disinfection of the effluent. In summary, the process usually provides a better effluent quality than conventional waste water treatment in terms of particulate matter and microbiological contamination (Judd & Judd, 2010). So far the major drawbacks and disadvantages are still the comparably high investment and operating costs through excessive fouling control, necessity of frequent membrane monitoring and maintenance, less efficient oxygen transfer due to higher MLSS concentration, frequent membrane changes caused by irreversible fouling and physical damages (Judd & Judd, 2010; Melin *et al.*, 2006).

Table 4: Published technologies for GW treatment reworked from Li *et al.* (2009)

Reference	Process	Non restricted agricultural reuse (WHO 2006)	Non restricted non-potable urban reuse (EU directive 2006/7/EG):
Gerba *et al.* (1995)	Cartridge filter	F	F
Birks (1998)	UF membrane	M	F
Nolde (1999)	Sedimentation + RBC + UV desinfection	M	F
Nolde (1999)	Fuidized-bed reactor + UV desinfection	M	F
Ward (2000)	Sand filter + Membrane + Disinfection	M	M
Brewer *et al.* (2000)	Filtration+Disinfection	M	F
CHMC (2002)	Screening + Sedimentation + Multi-media filter + Ozonation	M	F
Hills *et al.* (2003)	Coarse filtration + Disfinfection	M	F
March *et al.* (2004)	Screening + Sedimentation + Disinfection	M	F
Itayama *et al.* (2004)	Soil filter	M	F
Ramon *et al.* (2004)	UF membranes 30 – 400 kDa	M	M
Sostar-Turk *et al.* (2005)	UF / NF / RO	M	M (F for UF)
Sostar-Turk et al. (2005)	Coagulation + Sand filter + GAC	M	M
Friedler *et al.* (2005)	Screen + RBC + sand filtration + chlorination	M	M
Lin *et al.* (2005)	Electro-coagulation + disinfection	F	F
Liu *et al.* (2005)	MBR	M	M

Reference	Process	Non restricted agricultural reuse (WHO 2006)	Non restricted non-potable urban reuse (EU directive 2006/7/EG):
Lesjean & Gnirss (2006)	MBR	M	M
Prathapar *et al.* (2006)	Filtration + activated carbon + sand filter + disinfection	M	F
Merz *et al.* (2007)	MBR	M	M
Elmitwalli *et al.* (2007)	UASB	M	F
Gross *et al.* (2007)	Constructed wet lands	F	F
Kim *et al.* (2007)	Fibre filter media	F	F
Kim *et al.* (2007)	Metal membrane	M	F
Kim *et al.* (2007)	A2O MF + Ozone	M	M
Scheumann & Kraume (2007)	Submersed Membrane Sequencing Batch Reactor (SM-SBR, laboratory)	M	M
Hernandez *et al.* (2008)	SBR	M	F
Pidou *et al.* (2008)	Coagulation with aluminium salts	M	F
Pidou *et al.* (2008)	Magnetic ion exchange resin	M	F
Kraume *et al.* (2009)	SM-SBR (long-time study)	M	M
Li *et al.* (2010)	UF Membrane	M	M
Friedler & Gilboa (2010)	RBC + UV	M	M
Friedler & Gilboa (2010)	MBR + UV	M	M
Jong *et al.* (2010)	MBR	M	M
Huelgas & Funamizu (2010)	MBR (high load)	M	M
Buntner et al. (2011)	Three stage MBR	M	M
Bani-Melhem & Smith (2012)	Electrocoagulation + MBR	M	F
Jabornig & Favero (2013)	Moving bed biofilm	M	M

Reference	Process	Non restricted agricultural reuse (WHO 2006)	Non restricted non-potable urban reuse (EU directive 2006/7/EG):
	membrane reactor (MBBMR)		
Jabornig & Podmirseg (2014)	Fixed fibre biofilm membrane reactor	M	M

F…Fail guidelines
M…Meet guidelines

Fouling in filtration processes is inevitable and can be defined as a loss of permeability with time. Convective transport processes are the main principle of membrane filtration. The higher the permeate production through the membrane (= flux), the lower will be the back-transport of particles from the membrane surface into the suspension. This results in an accumulation of particles near the membrane and therefore the probability of contact to the surface. Below a so called critical flux, there is equilibrium of accumulation and back-transport. Flux decline by time theoretically does not occur. Above the critical flux fouling is observed (Field *et al.*, 1995). Thus, fouling includes restrictions, occlusions or blocking of pores but also the complete coverage of the membrane surface with deposits. The deposits can be inorganic or organic origin, reversible or irreversible. A reduction in output can also be caused by clogging or sludgeing of the complete membrane module. Different fouling control strategies have been developed in order to maintain an economic operation of MBR plants. They include periodical back-flushing, continuous or intermediate air scouring, relaxation phases without filtration and more or less strong, but frequent chemical cleanings to remove organic and inorganic foulants from the membrane.

Since the beginning of MBR technology in the 70s' roughly 30% of the published articles about MBR technology were about fouling and fouling control strategies. It shows the importance but also the still unsolved correlations. Main related parameters which have shown to be influential on fouling rate are (i) bioreactor design (MLSS concentration, sludge condition and extracellular polymeric substances), (ii) different applications (almost any kind of waste water) and (iii) membrane module design and operating philosophy (plate membrane, hollow fibre, different materials and configurations). The current status is far away a standardisation and is often contradictory (Drews, 2010). Suppliers and designers of

MBR plants usually provide their processes with general fouling control approaches, i.e. air scouring and chemical cleaning. Sophisticated and not practicable on-line monitoring systems of fouling parameters are not offered in a wide range so far. As a matter of fact, main energy demand for municipal MBR plants goes into air scouring of membrane modules. Basically, air scouring removes deposits mechanically from the surface by shear stress which is created through bubbles sweeping along the membrane. Recent research particularly focuses on strategies how to reduce this energy demand during operation. Optimized introduction of air to the membrane module, bubble size, intensity and frequency of aeration has been investigated so far (Verrecht, 2010). Despite some achievements in cyclic aeration during filtration (Verrecht, 2010), air scouring is still responsible for almost 35-50 % of energy demand of MBR plants (Judd & Judd, 2010).

Nonetheless, MBR technology has become an accepted process in waste water treatment whenever high quality effluent and area restrictions take effect. In Europe, there are already several hundred installations in middle and large scale (> 500 p.e.) and even some thousands in small and micro scale (Lesjean & Huisjes, 2007). With further falling membrane prices and optimization of energy demand, the outlook is even brighter and may represents the next evolution step in waste water treatment.

1.6. CHALLENGES OF GREY WATER RECYCLING

The market potential of water recycling and the actually existing market of GW recycling technologies could not be more different. Various treatment methods are offered on the market for GW treatment. Depending on the technical concept, they achieve WHO reuse guidelines for agricultural irrigation up to unrestricted non-potable urban reuse (e.g. toilet flushing). On the other hand, the number of sold units is still very low. One of the market leader in GW treatment, namely Weise Water Systems, reports only from roughly 1,000 units for a capacity < 1 m^3/day and 300 unit for a capacity < 10 m^3/day from the year 2001 on.

Hygienic aspects play a very important role in the GW reuse discussions. In fact, GW is human waste water with a relatively high amount of microbial activity and also may contain pathogens. The reuse applications and therefore the potential risks for humans to be exposed to GW dictate the appropriate treatment method. Reuse applications with direct contact to humans, i.e. garden spraying or car washing, have higher exposure risks and therefore usually require advanced treatment methods including disinfection. Although it is published that pathogens may appear in untreated and treated GW, a 60 year study in California revealed that there haven't been any case where GW is linked to diseases such as Cholera, Hepatitis A or Salmonellosis (Sheikh, 2010).

The reliability on the GW treatment unit in terms of output quantity and quality will be also a major factor for the acceptance. But how to design a common treatment plant when raw GW characteristics have a strong variability due to personal habits? Thus, the treatment system is more or less dependent on the good will of the consumer. For example if a biological treatment process is applied, the use of hazardous cleaning agents need be implicitly avoided. Donner *et al.* (2010) proposed improved eco-labelling, green procurement, information campaigns, substance substitution, and regulatory controls for household chemicals when GW recycling is applied. In order to prevent electro-mechanical failures in on-site MBR plants, Friedler *et al.* (2008) proposed on-line warnings, on-line measurements of main process parameters and regular maintenance of equipment which of course go along with increased costs.

Odour and nuisance problems are often due to remaining biological contaminants after treatment of GW and are also linked to anaerobic degradation processes in pipes and storage. These problems may only be eliminated through advanced chemical or biological treatment and low retention time. The regrowth of microorganism in the reuse water pipes can be limited but not completely eliminated by disinfection chemicals. Membrane filtration is an elegant way to retain bacteria from the reuse system as it acts as physical barrier. Nonetheless even in these systems regrowth has been measured because of small membrane ruptures or droplet transfer from raw water tanks or reactors to open clean water tanks (Merz *et al.*, 2007).

GW reclamation or rain water usage is seen as green technology and is therefore applied even if it doesn't pay off in the end. Therefore the extensive use of chemicals and high energy demand wouldn't be accepted in terms of environmental aspects (Nolde, 2005). However, a general statement is also difficult in this aspect. While in Germany, Austria and Switzerland the use of chlorine for final disinfection is commonly not accepted by the population, people in other countries would not use treated grey water without the typical chlorine smell.

Besides ecological reasons for the decision to install a GW treatment system, the economic aspects are very important. In the end GW treatment systems for single households or larger facilities need to pay-off. This can be achieved either due to lower investment and operating costs, high fresh water and sewage disposal costs or due to subsidies by the state (Friedler & Hadari, 2006). Currently, there is only a decision between low-tech equipment and rather effluent results or an expensive and more reliable technical solution with high operating and maintenance costs. A comparison of existing on-site in-door reed bed and MBR plants for GW treatment in Austria showed that the reed beds are in terms of investment costs and energy demand unbeatable. In terms of quality of the effluent they are fragile to shock loads and do require further disinfection to achieve high reuse standards. In contrast, the advanced biological SBR and especially the MBR systems have high energy consumption with up to 9.2 kWh/m³ but they are very reliable in terms of operation and quality of the reclaimed water. Periodic controls by the owner of the plants require both systems (BMFLUW, 2009).

Wide dissemination can only go hand in hand with technical robustness and economical sustainability. Generally, four main requirements have to be met: (i) hygienic safety, (ii) aesthetic aspects, (iii) environmental tolerance and (iv) economic aspects (Nolde, 2005). The difficulties of course lie in the realisation of these rather simple statements into a technical and economical feasible solution. This challenge was the driving force for this thesis.

1.7. COSTS OF GREY WATER REUSE

In literature many different system were tested due to their technical feasibility but only few systems were described in terms of investment and amortisation. In recent studies RBC treatment plants would be only feasible if they are used for at least 100 people (**Table 5**). In the same study the on-site MBR systems wouldn't be cost-effective within a suitable time of less than 15 years (Nolde, 2005; Friedler & Hadari, 2006). The strong influence of fresh water fees and sewage disposal costs is clearly visible. In case of subsidies the payback time would decrease, respectively (Friedler & Hadari, 2006). Humeau *et al.* (2011) showed that with current state-of-the-art MBR technology the costs per m³ treated grey water for less than 50 persons are even higher than the fresh water costs in France (3.01 €/m³) and only slightly lower than in high price countries like Germany (5.16 €/m³) and Denmark (6.18 €/m³). In summary, the investment in GW treatment in small scale applications, like single households, is not advisable with current state of the art processes.

Table 5: Amortization and costs of GW treatment

Reference	Process	Range	Freshwater and Sewage costs	GW costs / Payback time
Lazarova *et al.* (2003)	on site MBR	500 persons	-	1.70 €/m³ -
Nolde (2005)	RBC + UV	Hotel 400 beds	4.00 €/m³	- 6.5 years
Friedler & Hadari (2005)	RBC	5 storeys (28 flats)	1.46 $/m³	- < 15 years
	on site MBR	< 40 storeys	1.46 $/m³	- not feasible
Humeau *et al.* (2011)	direct NF	50 persons	3.01 €/m³	7.80 €/m³ -
	direct NF	500 persons	3.01 €/m³	4.82 €/m³ -
	on site MBR	50 persons	3.01 €/m³	7.40 €/m³ -
	on site MBR	500 persons	3.01 €/m³	4.40 €/m³ -

1.8. OBJECTIVES AND RESEARCH QUESTIONS

On-site GW could be one part of overall water recycling in urban areas. There are still many problems, which need to be solved in terms of feasibility, costing and technology. In consideration of the main obstacles in wide dissemination, the idea was created to develop a new approach with significantly lower investment and operating costs but with similar good effluent results as conventional MBR treatment.

Based on this idea, the main aim of this thesis can be summarized in three main objectives:

1. Assess the current situation in research of GW recycling and analyze market potential and feasibility.
2. Develop a new approach with strongly reduced investment and operating costs based on the membrane bioreactor technology.
3. Provide a mechanistic model specifying most influential parameters on the treatment process and by this provide a tool for design of GW treatment plants in future.

The main research questions to assess above objectives were:

1. What are the best suitable technologies currently available for on-site GW treatment in terms of performance, ease of use and dependence on utilities? (*Chapter 1*)
2. What are bottlenecks in wide dissemination of GW recycling technologies? (*Chapter 1*)
3. How can power demand and maintenance costs be reduced significantly with small-scale membrane bioreactors? (*Chapter 2*)
4. How can membrane bioreactors be operated with strongly reduced or even without fouling control of the membrane system? (*Chapter 3 and Chapter 4*)
5. What are the principles of sustainable flux development of membrane filtration without chemical cleaning in GW applications? (*Chapter 4*)
6. What parameters beside biological parameters influence GW treatment mostly? (*Chapter 5*)

1.9. OUTLINE OF THESIS

Chapter 1 "Overview and Feasibility of Advanced Grey Water Treatment Systems for Single Households" gives a general overview of recent developments in GW recycling, a detailed comparison of advanced GW treatment systems available for single households and a cost evaluation in view of on-site membrane bioreactor (MBR) systems for single households. Data and information were collected from demo units, publications, manufacturers and suppliers of these systems and were compared with experiences found in research literature.

Chapter 2 "Flux Stabilization in On-site MBR Plants with Reduced Fouling Control" investigates if strongly reduced fouling control applied in different membrane bioreactor set-ups leads to sustainable flux stabilization and by this to a significant decrease of operational costs. The pilot unit was operated at a wastewater treatment plant in Salzburg/Austria for several weeks. The operation experience and outcomes were used further for the development of an energy efficient GW treatment concept.

Chapter 3 “Single Household Grey Water Treatment with a Moving Bed Biofilm Membrane Reactor (MBBMR)” describes the design and operating data of a new approach of GW treatment. The combination of moving bed biofilm reactor with membrane filtration (MBBMR) was studied for on-site GW treatment for a single household with four inhabitants over ten months. Synthetic GW with different loading and varying ambient temperatures was part of the study. Although different stress situations were applied, the results of the treated effluent achieved international reuse guidelines.

Chapter 4 “A Novel Fixed Fibre Biofilm Membrane Process for On-site Grey Water Reclamation Requiring No Fouling Control” proposes a second new grey water treatment process, based on a concurrently working hollow-fibre membrane as fixed biofilm support and filtration device. Bioreactor characteristics, influence of different aeration rates, membrane flux development as well as structure and composition of biofilm were monitored to evaluate the performance of the tested pilot unit.

Chapter 5 “Modelling of Moving Bed Biofilm Membrane Reactors (MBBMR) for On-site Grey Water Treatment” evaluates with a mechanistic model the pilot plant results of a combined moving bed biofilm process and membrane filtration (MBBMR) treating single household GW. 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 pilot test lasting ten months. Besides, a sensitivity analysis was made, which calculates the relative significance factor (RSF) of each model coefficient and by this provides comparability with other studies.

1.10. THESIS ROAD MAP

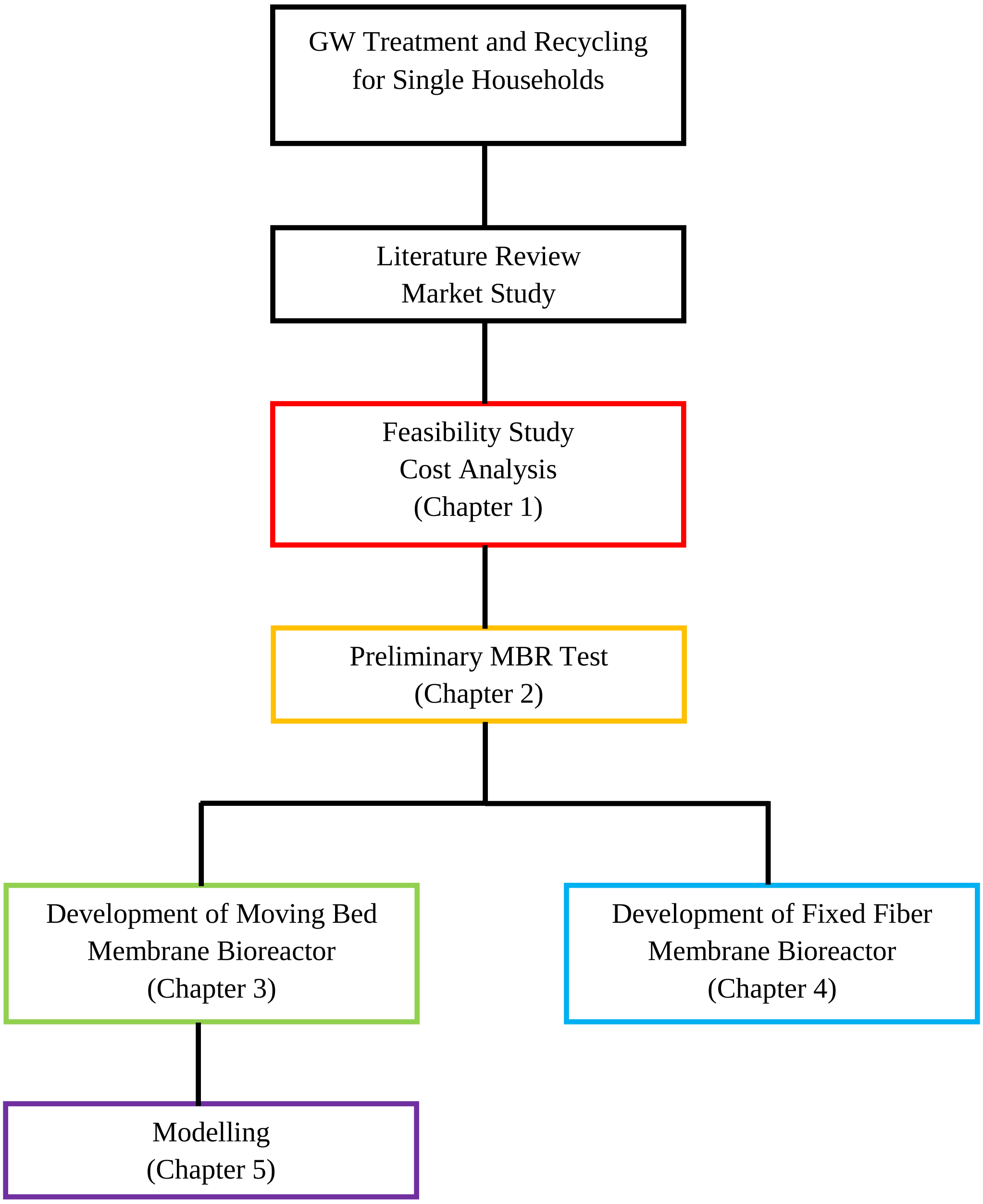

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Figure 1: Thesis road map

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CHAPTER 1:

Overview and Feasibility of Advanced Grey Water Treatment Systems for Single Households

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Overview and Feasibility of Advanced Grey Water Treatment Systems for Single Households

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Abstract

Although in recent years new developments for smaller applications have been made, investment costs have decreased and fresh water and sewage fees have increased, it has not been published so far if there are feasible grey water systems for single households with high quality effluent on the market. This paper gives a general overview of grey water treatment systems, a detailed comparison of available advanced grey water treatment for single households and a cost evaluation in view of on-site membrane bioreactor (MBR) systems for single households. The data and information were collected from demo units, publications, manufacturers and suppliers of these systems and were compared with experiences found in research literature. Results show that on-site MBR systems with daily grey water reclamation of more than 500 litres (10 persons) could be cost-effective. For single households the investment is still too high for a payback time of less than 15 years.

Keywords: advanced grey water treatment, feasibility, single household, on-site MBR

1. Introduction

Grey (Gray) water is defined as part stream of household wastewaters, which includes water from washing basins, bath tubs, showers and washing machines but sometimes excludes kitchen waters. In general grey water treatment does not include toilet flushing water (black water). In typical households up to 80% of the wastewater consists of grey water (Friedler and Hadari, 2006). Reported grey water flows start with 15 – 55 L/d/person (Nolde, 1999) up to 90 – 120 L/d/person (Morel and Diener, 2006) which could be theoretically treated and then recycled.

Grey water amount and characteristics strongly depend on location, grey water source (Jefferson, 2004), personal habits, personal activities (Eriksson, 2002) and even on population structure and income of households (Morel and Diener, 2006). In Germany, Nolde (2005) and Bullermann *et al.* (2001) made investigations in multi-storey buildings, which were combined in the guideline of the German Association for Rainwater Harvesting and Water Utilisation (fbr). They found partially big differences in the load and composition depending on the source but showed good biodegradability in terms of COD/BOD ratio and N and P nutrients. In contrast to this the studies of Jefferson *et al.* (2004) show a lack of nutrients and an unfavourable COD/BOD ratio, which could cause problems for biological treatment processes. The characterisation of grey water quality for different sources and different treatment systems in terms of design guidelines is still on-going and will require more studies. Besides, grey water may contain more than 900 other substances (Eriksson *et al.*, 2002), which can be important in terms of selection of treatment, reuse and disposal. Grey water is not free of microbial activity and possible pathogens. Depending on the source the total coliforms concentration can reach up to 7.0×10^8 CFU (colony forming units) per 100 ml (Eriksson *et al.*, 2002) and is therefore less than untreated sewage with total coliforms in a range of 10^7 to 10^{10} CFU per 100 ml (Crook *et al.*, 1998) but not without risks.

The recommended treatment methods for grey water depend on the composition of the grey water and its reuse application. Applications range from garden spraying, car washing and cooling water to washing machine and toilet flushing. These types are more or less sensitive in terms of human contact to the treated grey water. There is still no general valid international guideline for grey water treatment and reuse. The WHO guideline published in 2006 refers to agricultural reuse for irrigation only. Many countries have developed their own guidelines which can vary strongly (Li *et al.*, 2009). In Germany a legal guideline does not exist. Nevertheless, it is one of the countries with the highest number of grey water and rainwater treatment installations. The most popular guideline from the Association for

Rainwater Harvesting and Water Utilisation (fbr, 2005) gives values for agricultural reuse and toilet flushing. In Britain the BSI (British Standards Institute) published new guidelines in May 2011 i.e. BS8525. The guideline distinguishes between spray (pressure washing, garden sprinkler and car washing) and non-spray applications (toilet flushing, garden watering, washing machine). Non-spray limits are less strict except for reuse in the washing machine. Both require final disinfection with chlorine. In the US each state has its own reclaimed water guideline. A general directive is given in the EPA guideline for water reuse. Western Australia divides the reuse limits into exposure risks levels – levels of human contact (High, Medium, Low, and Extra Low). The highest exposure limit allows non-restricted agricultural reuse and non-restricted non-potable urban reuse in the new guidelines from August 2011 but requires further disinfection by chlorine or UV. As most countries do not have a national guideline for grey water treatment, the European Directive 2006/7/EG for bathing water quality has been used as basis by many European producers of grey water treatment plants, which specifies strict limits in terms of remaining nutrients and microbiological contaminants.
A wide range of treatment methods has been tested for different kinds of grey water. Most of the tested and published treatment methods achieve the WHO reuse guidelines for agricultural irrigation. For complying with the stricter EU guidelines final disinfection is required in addition to chemical or biological treatment. This disinfection can be achieved by sand or fibre filtration followed by UV radiation, ozone or chlorine dosing or by a physical barrier such as micro- or ultra-filtration membranes. For medium and high strength grey water biological treatment with Rotating Biological Contactor (RBC) with final disinfection or MBR could be a feasible solution (Li *et al.*, 2009). Nevertheless, sometimes pathogens can be found even after the disinfection unit, which could be caused by the so called "hopping phenomenon", which explains the transfer of bacteria from the grey water collection or treatment to the reuse system by aerosols, contaminated vessels and occasional sprays and spills (Friedler *et al.*, 2006).
While there are a great number of publications on different systems concerning their technical feasibility, only a few systems were described in terms of investment and amortisation of an on-site grey water treatment unit. In Germany, where RBC systems for 20 m^3/day have been used for more than 10 years and fresh water costs are comparably high, a payback time of only 6.5 years has been calculated (Nolde, 2005). In Israel, where fresh water and sewage fees are rather low, RBC treatment plants would only be feasible if they were used for minimum 5 storeys, which is equivalent to 28 flats. In comparison to this, on-site MBR systems would not be feasible at all within a suitable time frame (Friedler and Hadari, 2006). This shows that the

payback time strongly depends on freshwater supply and sewage disposal costs. They also applied governmental subsidies on their cost evaluation and found that the payback time would decrease accordingly (Friedler and Hadari, 2006). Humeau *et al.* (2011) compared direct nano-filtration and submersed MBR systems for 50 – 500 persons. They calculated much higher costs than Lazarova *et al.* (2003). The costs per m^3 treated grey water are higher than the fresh water costs in France (3.01 €/m^3) but lower than in high price countries like Germany (5.16 €/m^3) and Denmark (6.18 €/m^3).
From published literature no information can be obtained about which technologies are successful on the market in terms of quality and economic feasibility for single households. The aim of the author was to analyse information directly collected from suppliers and manufacturers and evaluate it with respect to different aspects. Many systems were found ranging from very simple to highly sophisticated technical solutions. Then a payback analysis will be conducted to verify if the implementation of such disposal is feasible for single households without subsidies.

2. Materials and Methods

2.1. Data Collection

The data were collected from pilot plants, internet research, direct enquiries and interviews with suppliers and manufacturers of small grey water treatment systems, review of publications, experience reports and manuals of private and governmental organisations.
The systems compiled were divided in low-tech and advanced treatment systems. The difference of low tech and advanced treatment was defined according to the stated quality of the treated grey water and the use of electro-mechanical facilities. Systems achieving water quality according to the EU Directive for bathing water with or without further disinfection and having one or more electro-mechanical items such as pumps or blowers which were directly used for the process (not for grey water pumping purposes) were defined as advanced treatment.
The evaluation criteria were divided in performance, ease of use and dependence of utilities as shown in **Table 1 - 1**.

Table 1 - 1: Evaluation criteria for grey water treatment systems

Performance	"+"	achieves EU bathing water guideline
	"o"	occasionally meets EU bathing water guideline, not shock load resistant
	"-"	fails to meet EU bathing water guideline
Ease of Use	"+"	maintenance free, installation by user, almost no user interference necessary
	"o"	installation by specialist, low maintenance, almost no user interference or automatic operation
	"-"	Installation by specialist, regular maintenance necessary, regular user interference necessary
Dependence on Utilities	"+"	no power, no chemicals, small area and space demand
	"o"	low power, no chemicals
	"-"	high energy demand or additional regular chemical or consumables for the process necessary

The suppliers of advanced treatment systems with the largest number of installations were contacted directly for further information including price and operational costs. The payback analysis was made with on-site MBR systems capable of < 500 L/d capacity. Four cases – 200, 300, 400 and 500 L/d (A0-B0-C0-D0) were evaluated, which would represent different numbers of persons per household. Per person 50 L/d grey water was estimated. Furthermore, a subsidy case (A1) of 50% investment subsidy and energy optimization of the treatment process for 200 L/d was evaluated.

2.1.1. Payback Analysis

The aim of the payback analysis was to review if small on-site MBR systems are feasible for single households compared to a zero investment case without treatment of grey water. The cost comparison was made for four different grey water productions per day treated by the same unit, which were each compared with a zero investment case.

The basis of the payback analysis represented the guideline for dynamic cost comparison for wastewater treatment plants (LAWA, 2005). The basic input parameters of the cost comparison were the investment and operational costs of small on-site MBR systems found in the survey and review of literature. The total investment was the sum of the gross unit price for the end user and the installation costs of a professional plumber. The annual operational costs included electrical power, maintenance, membrane replacement and chemical cleaning of the system. Each case with different flow was compared with a zero investment with the fresh- and sewage disposal fees as solely operational costs.

The reference date of the cost comparison was the day of investment. Therefore the increasing operational costs for n years were converted into a cash value (CV) at the day of investment. For this case LAWA suggests using the DFAKRP factor, which calculates as follows:

$$\mathrm{DFAKRP} = (1+\mathrm{p})\frac{(1+\mathrm{i})^{\mathrm{n}}-(1+\mathrm{p})^{\mathrm{n}}}{(\mathrm{i}-\mathrm{p})(1+\mathrm{i})^{\mathrm{n}}} \quad (1)$$

The cash value at the reference date is calculated with the DFAKRP factor multiplied with the total operational costs per year.

$$\mathrm{CV} = \mathrm{OC} \times \mathrm{DFAKRP} \quad (2)$$

DFAKRP	Discounting factor for operation costs after the reference date with price increase p
CV [€]	Cash value for operational costs after the reference date with price increase p
OC [€/y]	Operational costs per year
q [-]	Interest factor: q = 1+ i
i [%]	Interest rate
p [%]	Price increase for operation costs
n [y]	Period under review in years

For the zero investment case only the cash values of the fresh- and disposal fees were considered. The cash values for the on-site MBR cases (CVI) were the sum of total investment and cash value of the operational costs. The total cash value was then converted into annual costs by the annuities factor AF.

$$\mathrm{AF} = \mathrm{i}\frac{(1+\mathrm{i})^{\mathrm{n}}}{(1+\mathrm{i})^{\mathrm{n}}-1} \quad (3)$$

$$\mathrm{AC} = \mathrm{CVR} \times \mathrm{AF} \quad (4)$$

AF [-]	Annuities factor

AC [€/y]	Annual costs
CVR [€]	Cash value of total costs at reference date (CV operational costs + total investment)
i [%]	Interest rate
n [y]	Period under review in years

In each case the period under review was adjusted so that the CVR and annual costs were the same value for zero investment and treatment option. The determined period then showed the time frame when the plants got cost-effective for the user.

In **Table 1 - 2** the input parameters in the payback analysis are described. The data were collected from pilot plants, internet research, direct enquiries and interviews with suppliers and manufacturers of small grey water treatment systems and review of publications.

Table 1 - 2:: Input parameters of payback analysis

Parameter	**Description / Reference**
Grey water production	The value represents the mean grey water production per household and day.
Grey water reused	It was assumed that grey water collection and storage do not lead to losses. The reject through aerobic treatment was neglected (Merz *et al.*, 2007).
Operational days	It was assumed that the members of the household are away for 14 days per year with zero grey water production. Total operation days are 351 per year.
Investment costs	Investment costs were used from the conducted survey as gross acquisition costs of suitable systems for the end user. Costs for additional separation of drinking water supply and reuse water were not considered. It was assumed that these systems would be installed in new buildings only. The lowest investment costs for a system with 500 L/d found in the survey were used for the analysis.
Installation costs	Installation of the plant to an existing grey water system and start-up are usually simple. Plumber expenses were estimated to be one 8 hour manday at a rate of 55 EUR per hour and 20 EUR transportation expenses (Austrian Chamber of Labour, 2011).
Power demand	The power demand of the systems was taken from the conducted survey (Table 4) as a mean value of 11 different treatment systems. The calculated value was 3.3 kWh/m^3.
Power costs	Power costs per kWh were set to be 0.1919 EUR (E-control 2011, European mean value).
Maintenance	Per year 4% of the total investment was assumed for maintenance costs (fbr, 2005).
Membrane replacement	Membrane replacement is another major cost factor apart from energy demand. It was assumed that the membranes are designed for a rather low 5 LMH (L/m^2/hour – specific flow per m^2 membrane area) in order to avoid frequent chemical cleaning. Replacement costs were 50 EUR/m^2 (Judd and Judd, 2010). The mean life time was considered to be 5 years. The membranes themselves can be replaced without reinvestment of equipment.
Membrane cleaning	One cleaning per year is typically recommended by the suppliers. Membrane cleaning costs were estimated to be 10 EUR per year only. This is equivalent to one acid (250 g citric acid) and one alkaline cleaning (125 ml disinfection agent). Price basis was a drugstore in Austria.
Freshwater and sewage disposal costs	The mean freshwater (1.65 $€/m^3$) and sewage disposal costs (2.36 $€/m^3$) in Germany (DESTATIS, 2011) were used for the calculation.
Interest rate	The interest rate for the investment was calculated with 3%.
Price increase for operation costs	Increase was assumed to be 2 % per year.
Price increase freshwater and sewage disposal fees	Increase was assumed to be 2 % per year (DESTATIS, 2011).

3. Results

The data collected were reviewed and divided into two types of treatment systems: low-tech and advanced treatment. Systems achieving EU bathing water quality were further investigated concerning investment and operational costs. The review of the advanced treatment systems shows that only membrane based systems are widely used for single household applications with a daily treatment capacity of less than 500 L. This is why the payback analysis was made for on-site MBR systems only.

The systems available for grey water treatment are quite different in size, construction, ease of use and treatment process. The simplest treatment unit consists of mere coarse filtration, followed by direct reuse without raw and clear water storage. The most sophisticated systems have storage capacities and comprise different treatment steps including pre-filters, biological treatment, membrane filtration and final disinfection. An overview of available systems found in the research among suppliers and retailers is given in **Table 1 - 3**.

Low-tech systems usually do not achieve the values of the EU bathing water directive and therefore the treated grey water is not used for purposes where it can come into direct contact with humans. Most of these systems are for small applications only e.g. gardening or direct connection to toilet flushing. Their installation is simple as they are available as small skids and can be implemented directly by the user. These systems often do not have any electro-mechanical parts, which makes them cheap and low in maintenance.

The size of advanced grey water systems ranges from single households to multi-storey-buildings. Advanced grey water systems have sophisticated control units including solenoid valves, pumps and blowers. Investment and maintenance costs are higher but they also achieve the highest standards in terms of reuse. Installation needs to be done by specialized plumbers. The control unit operates the system automatically and allows setting several operation cycles such as holiday mode. Furthermore, most of these systems include connections for fresh- and/or rainwater supply to increase water savings. Some suppliers additionally offer thermal heat recovery from the treated grey water, which should make advanced grey water systems more feasible.

Advanced grey water systems for single households were investigated in more detail concerning technology, operation costs and investment costs. The survey shows that there is strong activity for these small systems in Germany and Australia. While several systems are available, the total number of installed units seems to be comparably low. One famous supplier in Germany supplied about 1,000 units for a capacity of less than 1 m³/day. The companies and technologies used are listed in **Table 1 - 4**.

Table 1 - 3:: Grey water treatment systems available on the market

Description of Treatment Systems	Performance	Ease of Use	Dependance on Utilities
Low-tech			
Chlorine disinfection	-	+	-
Coarse filtration + (disinfection with chlorine tablets)	-	+	+
Coarse filter + 100 µm fine filter	-	+	o
Sedimentation + (disinfection)	-	+	+
Coagulation/flocculation + Sedimentation + (disinfection)	-	o	-
Filtration through soil bed	o	+	+
Filtration over ceramic filter bed + (disinfection)	o	o	+
Reed bed	o	+	+
In-door reed bed with internal recirculation	o	o	o
Advanced Treatment			
Pre-membrane filter, sedimentation, UF membranes	+ (o)	o	o
Biological treatment with activated sludge technology based on black water treatment units + UV disinfection	+	o	o
Biological treatment with moving or fixed bed bioreactor + UV disinfection	+	o	o
Biological treatment with RBC (rotating biological contactors) + UV disinfection	+	o	o
Fine filter + activated carbon filters + MF filter	+	-	-
Multilayer filter bed + RO (reverse osmosis)	+	-	-
MBR (membrane bioreactor): external MF or UF-membranes submersed MF or UF-membranes	+	o	o
Ozonisation, pre-membrane filter, UF membranes, final disinfection	+	-	-

Except two, all companies use membrane filtration, either as MBR or direct UF/MF filtration, as treatment step. Furthermore, all suppliers except one, use biological treatment for decreasing the nutrients in the grey water. The processes include activated sludge technology, moving bed or fixed bed bioreactors and membrane bioreactors. No chemical processes are used by the suppliers of treatment systems for single households.

The mean specific energy consumption according to suppliers' information was calculated with about 3.3 kWh/m³. A deeper look into MBR applications showed that energy consumption for membrane filtration gets optimized compared to standard MBR by also using the cleaning air for the membranes for the aeration of the bioreactor. The filtration pressure through the membrane and as a result the energy demand is kept very low in order to avoid membrane blocking and chemical cleaning. Investment costs for these system lie within a range of 4,300 EUR – 9,000 EUR excluding installation. As option usually a rain- and/or freshwater connection is included to allow supplying additional water in case of grey water shortages. The use of sophisticated control units with PLC further allows direct user interference by SMS alarms. An optional heat recovery of grey water is also offered by one supplier. Within small single-household systems the on-site MBR is most popular amongst the suppliers, which is also reflected in the number its installations. Therefore, a payback analysis was made with different flow cases for small on-site MBR systems. As a first case a minimum daily flow of 200 L/day (A0) was verified. This first case represents the standard grey water amount for a four people household with about 50 L/d/person. The payback analysis shows that this case is not feasible, because the payback time would be more than 100 years. As the systems offered can usually treat more than 200 litres per day, higher daily flows with the same investment were verified. This includes a 300 L/day, 400 L/day and 500 L/day case. The analysis of the other cases shows that the payback time strongly decreases with the capacity. The reason for this is that sewage and freshwater fees proportionally increase more, while operation costs stay almost the same. The result is that the 300 (B0) L/day and 400 L/day (C0) cases are beyond 15 years depreciation time. The 500 L/day (D0) case proves to be feasible with a suitable time frame of less than 15 years. As a final case a subsidy and lower energy demand were assumed in the payback analysis of the 200 L/day (A1) case. A subsidy of 50% of the investment costs and 50% energy demand decrease show that even small 200 L/day systems are hardly feasible. Table 1 - 5 gives the summary of the cash value calculation of cases A1 (200 L/day + subsidy + energy optimization) and D0 (500 L/day).

Table 1 - 4:: Suppliers of advanced grey water systems for single households

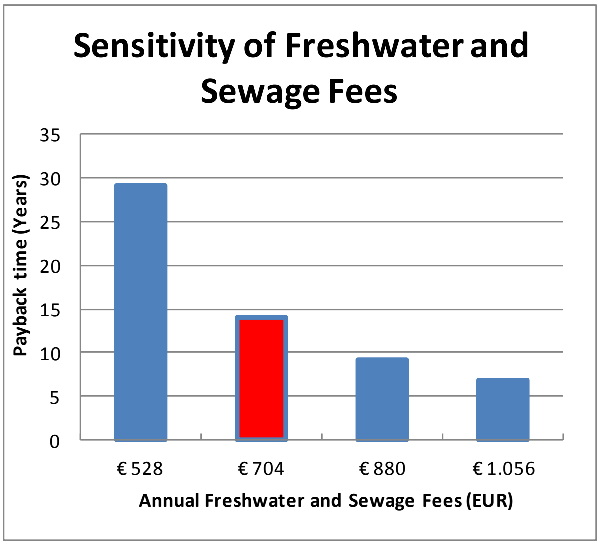
Company	Product Name	Capacity [L/day]	Process Description	Spec. Power Demand [kWh/m³]	Investment Costs
AquaClarus	Super Natural Grey	1,300	pre-filter, trickle bed, external UF, UV, clear water storage	1.46	n.a.
Aqua-Pluvia	Aquanus	300	pre-filter, grey water storage, UF, clear water storage	n.a.	n.a.
Dehoust	GEP-Wassermanage r WME-4	300	pre-filter, MBR, clear water storage	6.00	€ 5,400
EwuAqua	iClear 200 indoor	200	pre-filter, MBR, clear water storage	6.50	€ 5,800
Green Life GmbH	GWI 1.0-250 Indoor	250	pre-filter, MBR, clear water storage	6.00	€ 4,800
Hans Grohe	Pontos AquaCycle 2500	2,000	pre-filter, 2stage biological treatment, UV, clear water storage	1.25	€ 6,000
Hans Huber AG	GreyUse	1,500	pre-filter, MBR, clear water storage	3.80	n.a.
Intewa	Aqualoop	400	pre-filter, MBR, clear water storage	0.60	n.a.
Mall Umweltsysteme	GW / 600	600	pre-filter, MBR, clear water storage	1.72	€ 8,900
Nubian Water Systems	GT 600	600	pre-filter, feed water tank, fixed bed biological treatment and adsorption, UV disinfection, clear water storage	4.10	n.a.
Spin Flow	Spin Flow Grauwasser	500	pre-filter, MBR, clear water storage	3.00	€ 4,300
Water Gurus	Nova Grey	600	pre-filter, MBR, UV, clear water storage	n.a.	n.a.
Weise Water Systems	MicroClear® Aquacell 800	800	pre-filter, MBR, clear water storage	2.00	€ 4,600
				Min 0.60 Max 6.50 **Mean 3.3**	Min € 4,300 Max € 8,900 **Mean € 5,685**

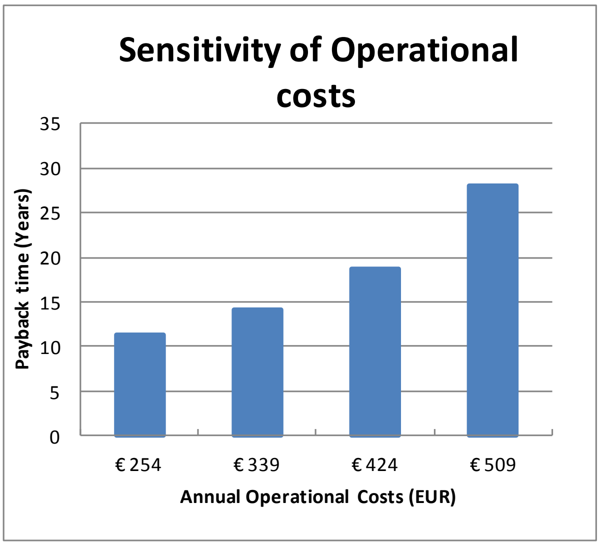
n.a. – data not available

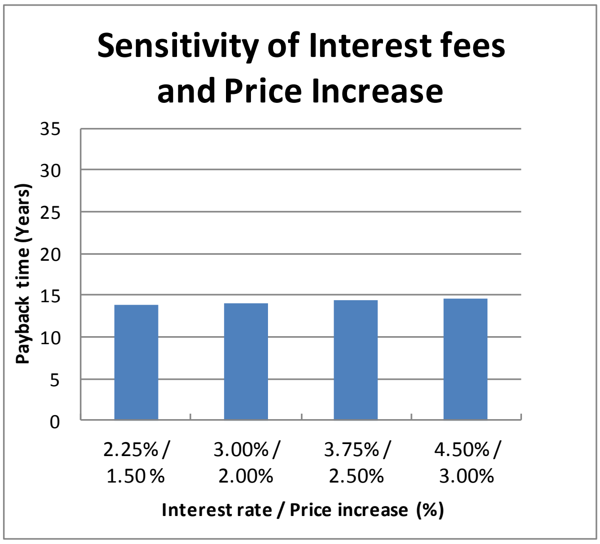
Table 1 - 5:: Payback analysis for cases A1 and D0

Case		A1		D0	
Basic input parameters		**No Invest**	**0.20 m³/d**	**No Invest**	**0.50 m³/d**
Grey water production	m³/d	0.200		0.500	
Grey water reused	m³/d	0.200	0.200	0.500	0.500
Years until cost-effectiveness	**y**		**19.7**		**14.0**
Investment					
Plant	€		€ 2,150		€ 4,300
Installation	€		€ 460		€ 460
Total investment	**€**	**€ 0**	**€ 2,610**	**€ 0**	**€ 4,760**
Operational costs					
Power demand	€/y	-	€ 22.23	-	€ 115.57
Maintenance	€/y	-	€ 86.00	-	€ 172.00
Membranes	€/y	-	€ 16	-	€ 40
Chemical cleaning of membranes	€/y	-	€ 10.0	-	€ 10.0
Sewage and freshwater fees	€/y	€ 282	€ 0	€ 704	€ 0
Total operation costs	**€/y**	**€ 282**	**€ 135**	**€ 704**	**€ 339**
Cash values (CV)					
CV operation of treatment plant	€	€ 0	€ 2,401	€ 0	€ 4,430
CV disposal	€	€ 5.011	€ 0	€ 9,190	€ 284
CV investment	€	€ 0	€ 2,610	€ 0	€ 4,760
CV total	€	€ 5,011	€ 5,011	€ 9,190	€ 9,190
Annual costs					
Annual costs		€ 341	€ 341	€ 812	€ 812
Costs per m³			4.86 €/m³		4.45 €/m³

The sensitivity of the input parameters was verified with case D0 (500 L/day). **Figures 1 – 1, 1 – 2, 1 – 3 and 1 – 4** show the payback time with varied investment costs, operational costs, freshwater/disposal fees and interest/price increase in a range of -25%, +25% and +50% of the initial value used for the cost evaluation.







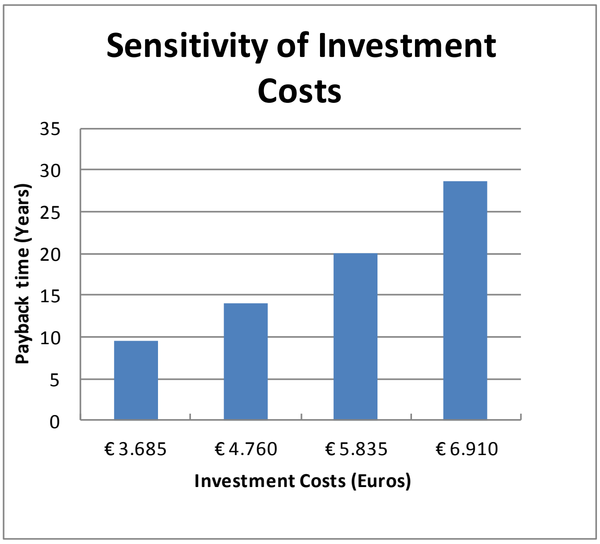


Figure 1 - 1: Sensitivity analysis of freshwater and sewage fees for case D0.

Figure 1 - 2: Sensitivity analysis of operational costs for case D0.

Figure 1 - 3: Sensitivity analysis of interest fee and price increase for case D0.

Figure 1 - 4: Sensitivity analysis of investment costs for case D0.

Higher investment and operational costs strongly increase the payback time of on-site MBR systems. Even a small increase of 25% would make them not cost-effective within 15 years anymore. On the other hand, if freshwater supply and sewage disposal got more expensive, the payback period would drop sharply and would also make smaller units cost-effective for the users. Interest and price increase rate had a rather low impact.

4. Discussion

The number of grey water reuse applications and systems found in the research around the world show an increasing public awareness and acceptance of grey water reuse. Non-existing standards and legal regulations led to quite different philosophies of treatment. Literature also shows the various numbers of physical, chemical and biological processes tested (Li *et al.*, 2009). However, most applications are very simple and also have a limited reuse range.

Many of the companies that offer advanced systems for grey water use a combination of biological treatment and a final membrane filtration step as process. This investigation proves the statements found in the literature (Pidou, 2006; Friedler and Hadari, 2008; Li *et al.*, 2009) that on-site MBR systems show good reliability and produce good reuse quality. So, both companies and researchers are convinced that on-site MBR systems will have a promising future.

Nonetheless, compared with the total number of new homes yearly built, the numbers of installed on-site grey water treatment MBR plants play an insignificant role. Bigger and therefore even more feasible systems have also only been built in a range of some hundred systems in Germany so far (Nolde, 2005).

Small on-site MBR systems can be feasible for more than 10 persons (500 L/day), because the investment into these systems is paid off by lower operational costs within 15 years. This result is much better compared to an earlier payback analysis, which shows that such systems are partly feasible for more than 50 persons only (Humeau *et al.*, 2011). The reason for this could be found in the higher investment costs of 12,700 EUR/m³ compared to the value conducted by the survey of 9,520 EUR/m³. Furthermore, Humeau *et al.* (2011) considered labour costs for the operation of the treatment unit, which is not applicable for private household applications and increased the operational costs significantly in his comparison.

The input parameters used in the analysis had a different influence on the results. A major factor if these small on-site systems could be feasible are the fresh- and sewage disposal costs. The higher these costs, the faster the investment in on-site MBR systems turns cost-effective e.g. 25% increased fees would lower the payback time of case D0 from 14 to less than 10 years. Furthermore, the operational costs of the treatment unit have a strong impact on the result. Maintenance and electrical power consumption together represented over 80%. A decrease of these costs would decrease the payback time as well. The interest and price increase had rather a low influence on the result.

A required daily flow like this is still too much for single households with a raw grey water amount of less than 200 L/day, but could be a start for further decrease of investment costs and energy demand. With a subsidy of 50% and half the energy consumption the payback time for single household applications can be lowered, but even then it is still longer than 15 years. The feasibility of such systems will require further improvements or even a new philosophy of treatment and processes.

The mean value of specific energy consumption according to suppliers' information was calculated with 3.3 kWh/m³ only, which is in contrast to experience reports that show higher values of up to 9.2 kWh/m³ (BMFLUW, 2009). A high energy demand prolongs the payback time, which also influences the acceptance of the users in terms of ecological considerations (Nolde, 2005). Another problem found during the research is that grey water reclamation often goes hand in hand with water saving technologies. So, saving fresh grey water may be good from an environmental point of view on the one hand, but on the other hand also negatively affects the feasibility and amortisation of treatment plants. This could be a further bottle-neck concerning the feasibility of small systems. From this point of view the connection of as many grey water sources as possible including washing machine and kitchen would be necessary. In contrast to this it is often not recommended by the suppliers to connect sources other than showers, bath tubs and washing basins.

One possibility to evade the economically unviable single-household application would be grey water collection for 2 or 3 households and common reuse. However, common reuse could conflict with psychological acceptance because common reuse may result in refusing the "extrinsic waste water" of others (Sheikh, 2010). Furthermore, maintenance and responsibility need to be arranged well in advance.

5. Conclusion

The author´s research provides evidence that innovations in on-site grey water MBR systems decrease the feasible capacity to 500 L/day, which is equivalent to the daily production of 10 persons. However, it is still not feasible to invest in a grey water treatment plant for a single household if high effluent quality (EU bathing water guideline) is requested by the usage. The current situation either requires subsidies of more than 50% investment and operation costs or further improvements of technology.

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CHAPTER 2:

Flux Stabilization in On-site MBR Plants with Reduced Fouling Control

Simon Jabornig

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Flux Stabilization in On-site MBR Plants with Reduced Fouling Control

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Abstract

In recent years small membrane bioreactor (MBR) plants have become popular for on-site grey and sewage treatment. The biggest advantage of on-site MBR plants is the membrane barrier, which at least provides acceptable water quality in terms of suspended solids and removal of pathogens in case of malfunctions of the bioreactor. Main disadvantages are still additional investment and high operational costs due to filtration, chemical cleaning and fouling control equipment. This study investigates if less fouling control applied in different bioreactor set-ups leads to sustainable flux stabilization and by this to a decrease of operational costs. The results show that flux stabilization could be achieved between 2 - 4 LMH (L m^{-2} h^{-1}) for two different bioreactor types, typically for grey and sewage treatment. The membrane aeration could be lowered by at least 95%, while back-flushing was still applied. As membrane costs for small applications usually play a minor role in the overall investment and membrane costs will decrease further, this operation method could be feasible in future.

Keywords: on-site MBR, fouling control, flux stabilization

1. Introduction

Flux is defined as the specific flow through the transversal membrane area and is used as performance parameter in membrane processes. During normal operation the flux decreases by time, which is normally the result of biofouling and scaling of the membrane surface and pores. Flux stabilization describes the effect of a continuing constant but comparably lower flow through the membrane over a long period of time during dead-end filtration without any fouling control. **Peter-Varbanets *et al.,* 2010**, describes this phenomenon with low pressure ultrafiltration membranes operated with different types of surface water and diluted wastewater. Sustainable or critical flux in conventional MBR processes is usually higher because of continuous fouling control such as air scouring, back-flushing or relaxation, which increases the back transfer of particles into the solution. Flux stabilization without fouling control mechanism is due to a heterogeneous but very active biofouling layer on the membranes (**Peter-Varbanets *et al.,* 2010).** In membrane bioreactors submerged membrane modules are often operated in a semi cross-flow configuration. The cross-flow is generated by air which is introduced below the membranes. The shear stress through the coarse bubbles and the cross-flow of liquid removes the sludge layer from the membrane. This operation method is usually also more energy consuming than classical dead-end filtration without cross-flow.

An interesting application of low energy consuming dead-end membrane operation would be small plants for grey and sewage treatment. Grey water includes sources like washing basins, bath tubs, showers and washing machines, while combined wastewater further includes kitchen waters and toilet flushing water. In some households a share of up to 80% of the wastewater consists of grey water (**Friedler & Hadari, 2006**). Typical flows start with 15 – 55 L d^{-1} and person up to 90 – 120 L d^{-1} and person (**Nolde, 1999**, **Morel & Diener, 2006**). Treatment methods of on-site grey water systems can be found in a wide range starting with simple coarse filtration up to multi-barrier concepts including biological treatment, membrane filtration and UV disinfection (**Li *et al.,* 2009**). Due to missing standards and depending on the requested effluent quality by the usage different low-tech and advanced treatment methods could be applicable for grey water. The treatment of sewage requires biological treatment either by activated sludge technology, sequenced batch reactors (SBR), MBR, reed beds or biofilm processes. In contrast to grey water, there are also often legal effluent quality regulations for sewage treatment units in European countries for up to 50 p.e. In recent years the number of small on-site MBR plants for grey and sewage applications has been raised and proved to be a feasible alternative if high effluent quality is required. However, the investment and operational costs of these units in comparison to conventional biological treatment are still higher. (**Friedler & Hadari, 2006, Humeau *et al.*, 2011**).

The major part of energy consumption for these systems, namely 70 %, is used for air scouring the membrane modules as fouling control (**Judd, 2006**). While in bigger MBR plants sustainable flux and critical flux with full fouling control (air scouring of membranes, back-flushing, relaxation and chemical cleaning) have been well investigated (**Bacchin *et al.*, 2006**), the issue of flux stabilization with missing or reduced fouling control for small on-site MBR applications remains unclear.

This paper describes the attempt to lower the operational costs of these small on-site MBR systems by using less fouling control and the flux stabilization effect with hollow fibre membranes. This type of membranes comes along with a high packing density and therefore increased membrane area can be applied within small tanks. Strongly reduced aeration and back-flushing without mechanical or chemical maintenance cleanings were tested in different applications and configurations.

2. Materials and Methods

2.1. Pilot Plant Set-up

The membrane filtration tests for this publication were made on a side stream MBR pilot plant at the municipal waste water treatment plant (WWTP) located in Salzburg. Two independent membrane lanes were installed in order to verify the results with one another and to exclude operational and mechanical influences. **Fig. 2 - 1** shows the P&ID of the pilot plant.

For both registers submersed hollow fibre membrane modules were used. The modules comprise several thousand hollow fibres with a nominal pore size of 0.2 µm. The fibres were wound up around a carrier cartridge in order to increase their packing density. One cartridge had an active surface of 3 m². Four cartridges were combined in parallel operation to one module. In total a membrane surface of 12 m² was installed. Each lane had a separate online measurement for flow and pressure during filtration and back-flushing of membranes. The modules could be operated in suction mode till -0.9 bar and in pressure mode till 3 bar according to the supplier´s information. Permeate and back-flushing pump were of self-priming centrifugal type equipped with a variable frequency drive in order to regulate flow and pressure. So, the plant could be operated with constant pressure, constant flow and with different load curves for the same. Additionally, the constant pressure mode should simulate the gravity driven production without permeate pump. Back-flushing, when applied, was made in the range of 20 – 25 L m^{-2} h^{-1}.

For air scouring a membrane compressor was used for both lanes. The air flow was simply regulated by pressure loss through a manual regulation valve to about 1 $m^3 m^{-2} h^{-1}$. The plant was operated in suction mode without air scouring of membranes in dead-end mode. Air scouring was only switched on during back-flushing.

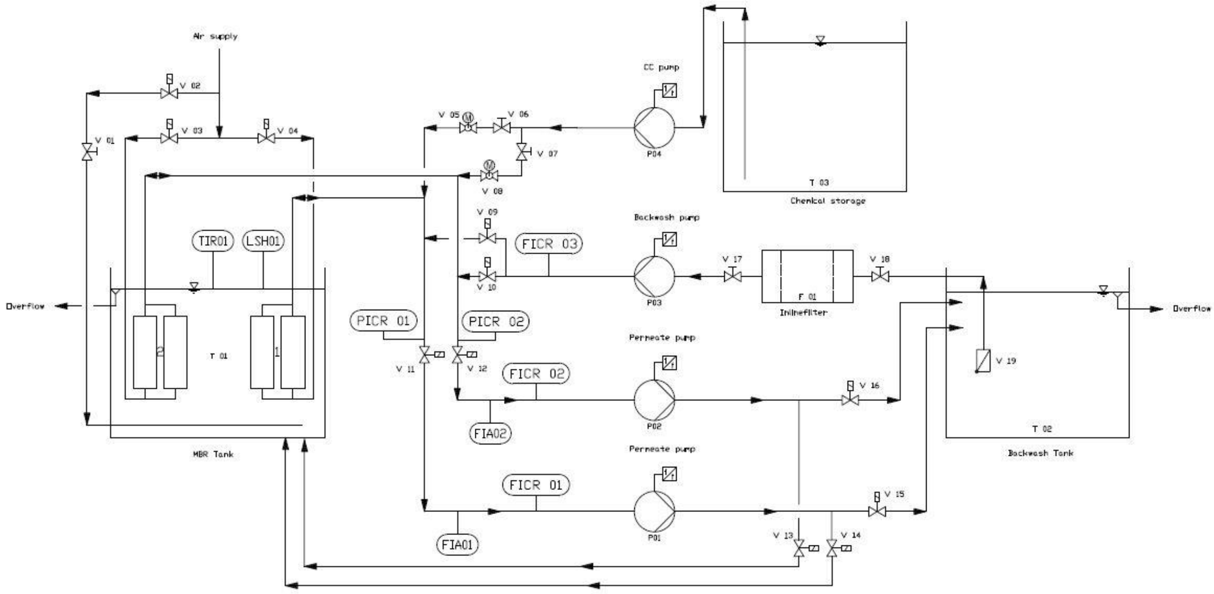

Figure 2 - 1: P&ID of pilot plant

The bioreactor for each test had a volume of 450 L and was equipped with a fine bubble diffuser system, a temperature measurement and a surplus sludge removal pump. The clear water was intermediately stored in a 450 L tank for quality measurement. The clear water overflow and the surplus sludge were rejected into the sewer system of the WWTP.

2.2. Operational Parameters Set-up

The fouling control with back-flushing and air scouring was strongly reduced. Air scouring was reduced by at least 95% and back-flushing by 50 – 100% compared to conventional fouling control settings in MBR treatment. Chemical cleaning was not applied during the tests. The strongly reduced fouling control should enforce the formation of a filter cake on the membrane. For the trials already used membranes were installed, which had been used in an activated sludge MBR for more than one year. The already fouled membranes should shorten the process of bioactive sludge layer formation. Furthermore, it should be assured that a certain number of pores were already blocked by irreversible fouling. After each trial the membranes were cleaned with a water flush from outside to remove the sludge layer. **Tab. 2 - 1** gives an overview of the different cycles and bioreactor conditions which were tested.

Table 2 - 1: Overview of operational parameters and bioreactor conditions

No.	Description	Filtration (sec)	Back-flushing (sec)	Air scouring (sec)
A1	High SS, constant flow 130 L/h	240	0	8
A2	High SS, constant pressure 0.2 bar	240	0	8
B1	High SS, constant pressure 0.1 bar	600	30	30
B2	Low SS, constant pressure 0.1 bar	600	30	30
B3	High SS, constant pressure 0.2 bar	360	20	20

The high suspended solid (SS) and low suspended solid bioreactor conditions simulated the operations in activated sludge or in a moving bed bioreactor. Activated sludge was introduced into the bioreactor from the WWTP three times a day and was kept in a range of 4 – 10 g L^{-1}. The lower suspended solid concentration was maintained between 0.05 – 1.0 g L^{-1}. Test time for each set up was 3 – 4 weeks. Previous tests and experience from literature showed that once flux stabilization was reached after 7 – 14 days, the flow would not change anymore if the bioreactor conditions remained constant respectively.

2.3. Measurements

The operational parameters flow, pressure and water temperature were measured by online instruments. The data were recorded every 1 - 2 sec. For every test day a mean value was calculated from the complete data and used for the analysis and charts. The flow measurements were additionally corrected with a temperature factor, based on the viscosity change of water.

$$j_{15°C} = \frac{j_m}{0.0266 \times T_m + 0.6} \qquad (1)$$

$j_{15°C}$ Flux calculated in L m^{-2} h^{-1} at 15°C

j_m Flux measured in L m^{-2} h^{-1}

T_m Actual temperature inside bioreactor in °C

The suspended solids in the bioreactor were measured manually with a 0.45 µm paper filter once a day. Additionally, the turbidity, colour and conductivity of permeate were tested with portable devices several times per day.

3. Results

3.1. Operation without Back-flushing and applied Flow Regulation

In test no. A1 the pilot plant was operated without back-flushing at a constant flow of 130 L h^{-1}. The flow was regulated and adjusted by a PID controller. Air scouring was set to 8 sec. every 4 minutes only. The permeate pump was stopped during air scouring. The suspended solid concentration of activated sludge reached levels of up to 10 g L^{-1}. **Fig. 2** shows the suction pressure development after back-flushing was lowered and then finally switched off.

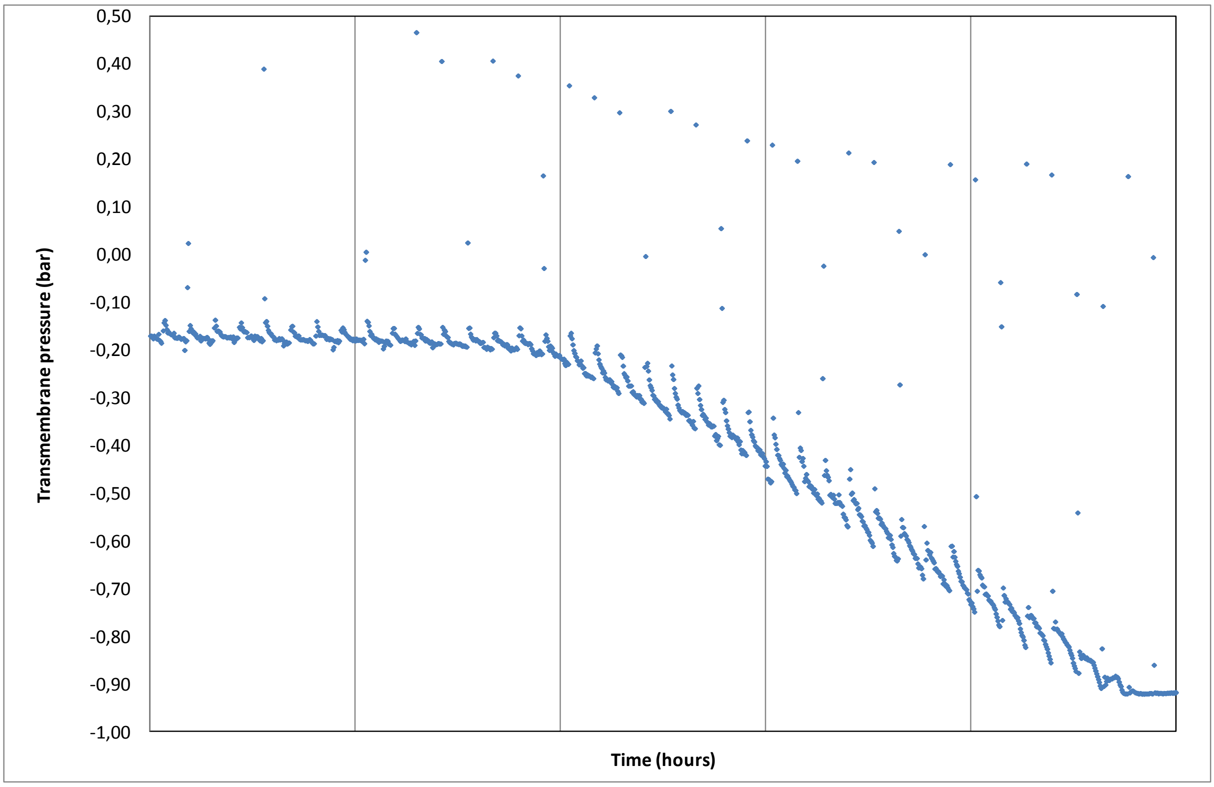

Figure 2 - 2: Transmembrane pressure (TMP) during constant flow operation without back-flushing

After only 1 hour the missing back-flushing led to a strong increase of the suction pressure which indicates that the critical flux has been exceeded. During that time the membrane surface was continuously covered with sludge and therefore the resistance of filtration further increased. The short air scouring pulse was not enough to ensure the back transfer of sludge from the membranes.

Within 70 minutes the PID controller had to increase the performance of the pump to maximum frequency to achieve the pre-set 130 L h^{-1}. A further increase was not possible due to cavitation of water at a pressure of lower than -0.9 bar. Anyway, the system was operated further to investigate if the flow would finally be zero. Surprisingly, after four days operation at these unfavourable conditions for membranes and equipment the flow was still in a range of 40 - 50 l h^{-1}. This could indicate that the thick and compressed filter cake as a result of the high suction pressure is still porous after several days. Channels must be there to allow water passing through the dewatered sludge layer and finally the membrane pores.

3.2. Operation without Back-flushing and Pressure Regulation

In comparison to Test B1, Test B2 was started without back-flushing but with a constant suction pressure of -0.2 bar. The constant pressure should simulate gravity driven operation of the membrane system. Air scouring was set to 8 sec. every 4 minutes. The suspended solid concentration of activated sludge reached levels of up to 8.5 g L^{-1}. Flux stabilization could be obtained at around 2 LMH after 7 days of operation and was kept constant till the end of the test period. **Fig. 3** shows that the flux started at about 4 LMH, which is a lower level than shown in the following tests B1, B2 and B3. The most probable reason could be the condition of the membrane after the first test with high pressure. The membrane was not chemically cleaned after the first test and a certain irreversible fouling may have blocked the membrane surface and the pores from the beginning. Therefore, the expected sharp drop of the flux after start-up is missing.

3.3. Operation at Constant Pressure with Reduced Fouling Control

In tests B1, B2 and B3 a simulated gravity driven dead-end filtration with reduced air scouring was tested for high and low activated sludge concentrations. The pilot plant was operated at a constant suction pressure of -0.1 bar and later with -0.2 bar. Between the manometer and the water level was a 40 cm geodetic level difference. Therefore, the actual transmembrane pressure was 0.04 bar lower. Fouling control was only started every 10 min. for 30 sec. After 4 weeks of operation the activated sludge concentration was reduced from up to 10 g L^{-1} to less than 1 g L^{-1} and operated for another three weeks. After a winter break of 8 weeks the system was operated again with high DS and additionally -0.2 bar suction pressure for 3 weeks. Fouling control was started in this test after 360 sec. for 20 sec. **Fig. 2 - 3** shows that both operation with low and operation with high sludge concentration start at about the same flux between 6 - 7 LMH and decrease to a range of 3 - 4 LMH within 1 week.

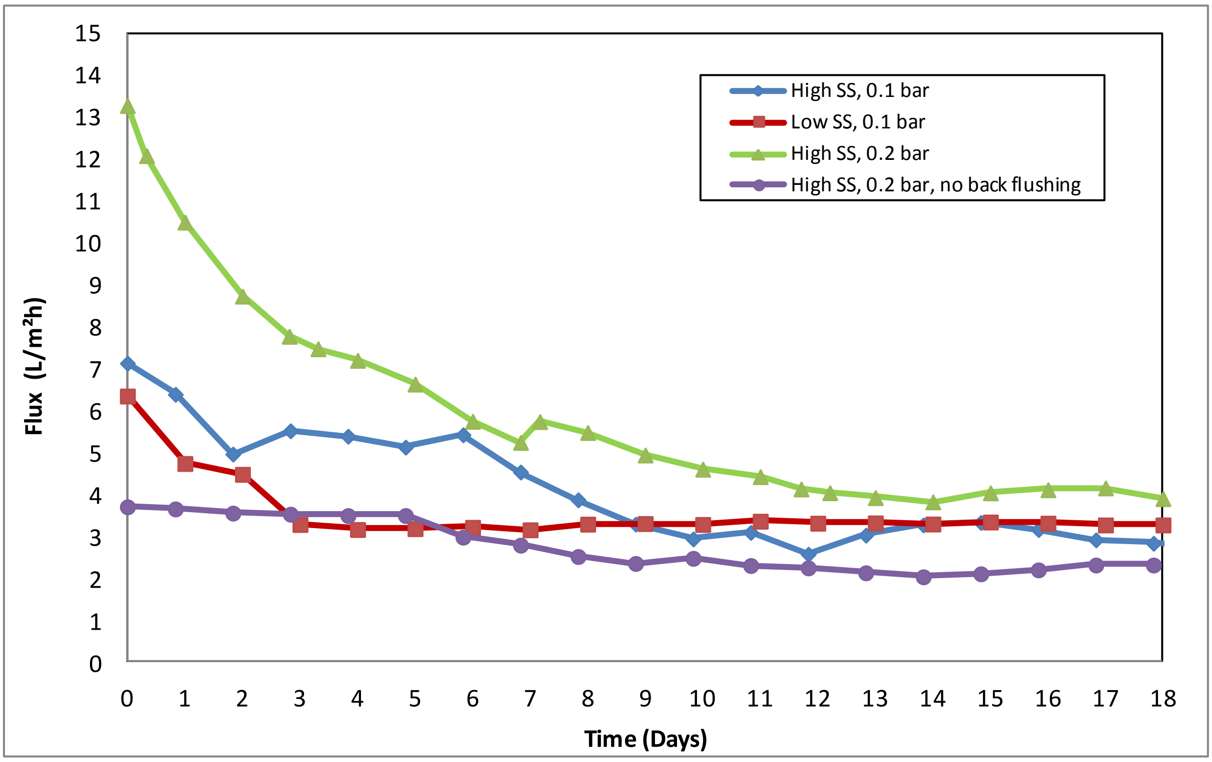

Figure 2 - 3: Flux development during three weeks operation with different TMP and suspended solid (SS) concentration

Anyhow, during the following two weeks the flux did not change anymore. Not surprisingly, the trial with higher suction pressure and high SS starts at about 13 LMH, which is about double the LMH at 0.1 bar pressure. However, the decrease in this type of operation is also fast and led to a stable flux at a similar but a little higher LMH within two weeks.

3.4. Quality of Effluent

The quality of the effluent of test B1, B2 and B3 were measured by a portable turbidity instrument. The results were comparable in all three set-ups. The mean value of 0.26 NTU for the operation at low pressure and high SS was the lowest value. The low SS set-up result was 0.36 NTU and the set-up with high pressure and high DS resulted in 0.30 NTU.

3.5. Sludge Layer on Membranes

Reduced fouling control led to the formation of a sludge layer on the membrane surface. After the tests the membranes were removed from the bioreactor and their weight was measured before and after manual cleaning with water flushing in order to check the quantity of sludge on the membrane surface. A mean value of 84 g m^{-2} was measured.

3.6. Energy demand

The energy demand of set-up A2, B1 and B3 was conducted with the data measured in the tests and compared with literature data. **Table 2 - 2** shows the calculated energy demand per day for each test set-up and the specific power demand per m^3 treated water without raw water lifting and bioreactor aeration.

The basic parameters of net flux and membrane area were used from the results of tests A2, B1 and B3. The maximum net output was calculated by multiplying net flux and membrane area. The permeate extraction energy demand was zero as it is gravity driven. Backwashing power demand was calculated with the installed power of the pump allocated on the time it was running. The aeration of membranes was only running during backwashing in full speed. The power demand was calculated accordingly. The energy demand of all processes was summed up and the specific energy demand for 1 m^3 treated water was calculated.

For all set-ups the specific energy demand was less than 0.6 kWh m^{-3}, in total less than 0.5 kWh day^{-1}. Energy demand of set-up A2 is the lowest but also has the lowest specific flow rate. Allocated on the flow it is lower compared to B1 and B3. If a typical flow for one people equivalent is around 150 L day^{-1}, then A2 can be applied for a household of 3 - 4 people, while B3 would be feasible for even 7 - 8 people.

Table 2 - 2: Operational parameters and power demand of set-ups A2, B1 and B3

Parameters		**A2**	**B1**	**B3**
Stabilized Flux	$L\ m^{-2}h^{-1}$	2	3	4
Membrane Area	m²	12	12	12
Max. Flow per Day	$L\ day^{-1}$	576	864	1,152

Energy Demand		**A2**	**B1**	**B3**
Permeate Extraction *)	$kWh\ day^{-1}$	0.000	0.000	0.000
Backwashing	$kWh\ day^{-1}$	0.000	0.254	0.254
Membrane Aeration	$kWh\ day^{-1}$	0.155	0.229	0.229
Total Energy Consumption	**$kWh\ day^{-1}$**	**0.155**	**0.482**	**0.482**
	$kWh\ m^{-3}$	**0.269**	**0.558**	**0.419**

*) Gravity driven

4. Discussion

4.1. Flux Stabilization and the Application in On-site MBR Plants

The results show that flux stabilization could be achieved with comparably low fouling control. For the tested hollow fibre membrane module the flux value was in a range of 3 – 4 LMH and was finally achieved in all experimental set-ups with back-flushing. Without back-flushing the final LMH was measured to be about 2 LMH.

High or low sludge concentrations as well as different transmembrane pressures didn't have much influence on flux stabilization and effluent quality. The reason why is probably the cake layer, which always builds up on the membrane surface due to missing fouling control. The structure of this cake layer was built up in a comparable way and quantity during all experiments and resulted in a similar filter resistance and barrier.

The high suction pressure test of up to -0.9 bar indicated that even a thick and very compressed filter cake allows a certain amount of water to pass through the membrane and the final flow will probably never be zero. Although this type of operation is more energy consuming and negatively effects the equipment through cavitation, this effect provides a certain safety during malfunctions as there will still be output from the plant.

4.2. Energy demand vs. Investment

The calculated energy demand of 0.269 - 0.567 kWh/m³ for the membrane process without bioreactor aeration and raw water lifting is compared to literature quite low. **Verrecht *et al.*, 2012**, collected values ranging from 3.0 - 11.5 kWh m^{-3} for complete on-site MBR treatment. The savings for air scouring energy demand were 95% compared to the standard operation of continuous aeration of membrane modules. The actual savings will probably be less because in small applications the air for air scouring is also used for oxygen supply to the bioreactor. Anyway, if membrane air scouring is separated from bioreactor aeration, the efficiency of oxygen transfer through fine bubbles can be increased and possible negative effects on the micro-organisms through sheer stress of coarse bubbles will be minimized.

4.3. Biological Activity of Sludge Layer on Membranes

As a mean value over the complete membrane surface 84 g m^{-2} was measured. This relatively high quantity was accumulated because of reduced fouling control. **Peter-Varbanets et al., 2010** showed that the sludge layer is quite active with a porous structure. Therefore, the flux probably did not go to zero, but kept a low but stable value. The question is, if the sludge layer, which reduces the flux, may act as additional bio-film and improves effluent quality or the degradation of nutrients.

5. Conclusion

The operation with strongly reduced fouling control and the flux stabilization effect without chemical cleaning could be applied for on-site MBR plants in a feasible range. However, the final flux value probably depends on further parameters, such as the module construction which affects the quantity and quality of the sludge layer. There is little doubt that the sludge thickness is the main reason for the filtration resistance. Membrane investment costs in small MBR applications play only a minor role and therefore the increase of membrane area and simultaneous reduction of fouling control could be cost-effective in the future.

Further research concerning long-term stabilization, quality of permeate and possible improvement of permeate quality by filtration through the sludge layer and additional effects on degradation through biofilm processes is pending.

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CHAPTER 3:

Single Household Grey Water Treatment with a Moving Bed Biofilm Membrane Reactor (MBBMR)

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Single Household Grey Water Treatment with a Moving Bed Biofilm Membrane Reactor (MBBMR)

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Abstract

A combination of moving bed biofilm reactor and membrane filtration (MBBMR) was studied for on-site grey water treatment for a single household with 4 inhabitants over 10 month. Synthetic grey water with different loading and varying ambient temperatures were part of the study. Although different stress situations were applied, the results of the treated effluent achieved NSF (International Standard /American National Standard) reuse guidelines. Mean flux of the membrane during permeating phase was 12.9 L/m²h. Additionally, natural hair colour could be removed by almost 80% and energy consumption of the pilot unit was optimized during the operation to less than 1.3 kWh/m³. The process configuration proved to be feasible for the implementation on on-site micro systems with high flow and load variation.

Keywords: on-site MBR, grey water, MBBR, BF-MBR, single household

1 Introduction

Grey water from showers, baths, wash basins, washing machines and kitchen is a major part of the overall household wastewater with a share of up to 80% (**Friedler & Hadari, 2006**). The daily amount per person can reach values of 15 – 55 L d^{-1} up to 90 – 120 L d^{-1} (**Nolde, 1999**, **Morel & Diener, 2006**). A wide range of treatment methods of grey water can be found in literature as well as already on the market. So far low and high-tech systems such as simple coarse filtration up to multi-barrier concepts including biological treatment, membrane filtration and UV disinfection exist together and have also been applied to different building sizes (**Li *et al.*, 2009**). An innovative and reliable technology in terms of advanced biological treatment and removal of pathogens represents the activated sludge membrane bioreactor (AS-MBR). **Table 3 - 1** lists recent studies about grey water treatment with AS-MBR plants and achieved treated effluent quality. All studies show excellent removal efficiencies of organics and microorganism but feasibility in view of operational costs and estimated investment remain unclear.

AS-MBR technology is applied only for a small number of on-site sewage or greywater treatment applications for single households till now (**Lesjean & Huisjes, 2008**). One obstacle for a broad dissemination could represent the investment and operational costs of commercially available units which are in comparison to conventional biological treatment still higher and therefore not cost-effective for small buildings. (**Friedler & Hadari, 2006, Humeau *et al.*, 2011**). Other acceptance problems may result from increased maintenance due to fouling of membranes (**Kraume *et al.*, 2010**) and in consequence higher energy demand for fouling control caused by air scouring of membranes (**Merz *et al.*, 2007**).

As a result of these disadvantages of the conventional AS-MBR systems, an alternative membrane configuration called biofilm membrane bioreactors (BF-MBR) were developed by **Leiknes & Ødegaard (2007)**. The BF-MBR reactor is divided into two compartments. One is filled with moving bed biofilm growth bodies and the other one is equipped with the membrane modules and acts as filtration chamber. Retained sludge is recirculated to the first chamber or removed by system by a pump. When operated with sewage, this process showed to have higher flux rates than conventional AS-MBR probably due to lower suspended solid concentration in the bioreactor. The low suspended solid concentration may also lead to reduced clogging of membranes and less sludge accumulation in the membrane housing. Thus less fouling control is needed to maintain a high flux.

In order to reduce investment and operational costs of small onsite MBR plants for grey water treatment, the authors of this study combined a moving bed biofilm reactor and membrane filtration system in one tank. The system was operated 10 month with synthetic low and high strength grey water. This new configuration should combine the advantage of high effluent quality and low space requirements of AS-MBR and the lower operational costs of the BF-MBR. The combined process was named Moving Bed Biofilm Membrane Reactor (MBBMR). This study describes the performance of this reactor type on a single-household scale and presents operational aspects and costs as well. The results of 10 month operation shows that the treated effluent of the combined process configuration meets international reuse guidelines and energy demand could be lowered to less than 1.3 kWh/m³.

Table 3 - 1: Recent studies on performance of grey water MBR

Reference	Membrane type	Water Source	BOD5		COD		TSS		TN		TP		Turbidity	
			mg/L		mg/L		mg/L		mg/L		mg/L		NTU	
			In	Out	In	Out	In	Out	In	Out	In	Out	In	Out
Lesjean & Gnirss (2006)	Flat plate	Bathroom and kitchen	-	-	493	24.0	90	< 1	21.0	10.0	7.40	3.50	-	-
Merz *et al.* (2007)	Hollow fibre	Sports and leisure camp	59.0	4.00	109	15.0	-	-	15.2	5.70	1.00	0.90	29.0	0.50
Kraume *et al.* (2009)	Flat plate	Synthetic grey water	-	-	194	17.5	-	-	10.8	2.81 [a)]	-	-	-	-
	Flat plate	Real grey water including kitchen	-	-	274	13.7	-	-	18.0	1.62 [a)]	-	4,40	-	-
	Flat plate	Showers	-	-	233	21.0	-	-	32.9	9.20 [a)]	-	0.85	-	-
Huelgas & Funamizu (2010)	Flat plate	Washing machine and kitchen sink	-	-	675	26.3	-	-	-	10.01 [a)]	2.37	-	-	-
Jong *et al.* (2010)	Hollow fibre	Washing, kitchen, cooking, bathroom and shower	-	-	808	6.57	2180	1.22	-	-	-	-	2131	1.63
Bani-Melhem & Smith (2012)	Hollow fibre	Cleaning, sinks and kitchen	-	-	463	65.0	78.0	-	12.2 [a)]	11.15	0.53	0.18	133	6.90

[a)]: TN calculated with NH_4-N + NO_3-N

2 Materials and Methods

2.1 Pilot Plant Set-up

A moving bed biofilm membrane reactor (MBBMR) was studied for a four person's household treating 200 L grey water per day. **Figure 3 - 1** shows the process flow scheme of the MBBMR pilot plant. The moving bed biofilm bioreactor and the membrane filtration unit were placed in the same tank with a capacity of 350 L which also included sufficient balancing volume for incoming grey water flushes. In low level this tank still had a volume of 110 L. As moving bed, cylindrical biomass carriers (d = 36 mm, h = 20 mm) made of HDPE (density 0.95-1.10 kg/L) with a specific area of 320 m^2/m^3 were used. The necessary moving bed biofilm area was calculated with an expected BOD_5 influent concentration for mixed grey water without kitchen waters proposed by the Association for Rainwater Harvesting and Water Utilisation in Germany (**fbr, 2005**) of 187.5 mg O_2/L, a daily grey water amount of 200 L, and a loading rate of the biomass carries with 0.004 kg BOD/m^2d. This resulted in a biofilm area of 9.4 m^2 or roughly 30 L of biomass carriers. The diffuser below the membrane module was concurrently used for membranes scouring, mixing and oxygen supply of the bioreactor. The air flow provided by a small compressor was set to 0.3 m^3/m^2h relating to membrane area. Aeration operated in an ON/OFF cycle 5 min/10min and later 1 min/10 min in order to further reduce the energy consumption.

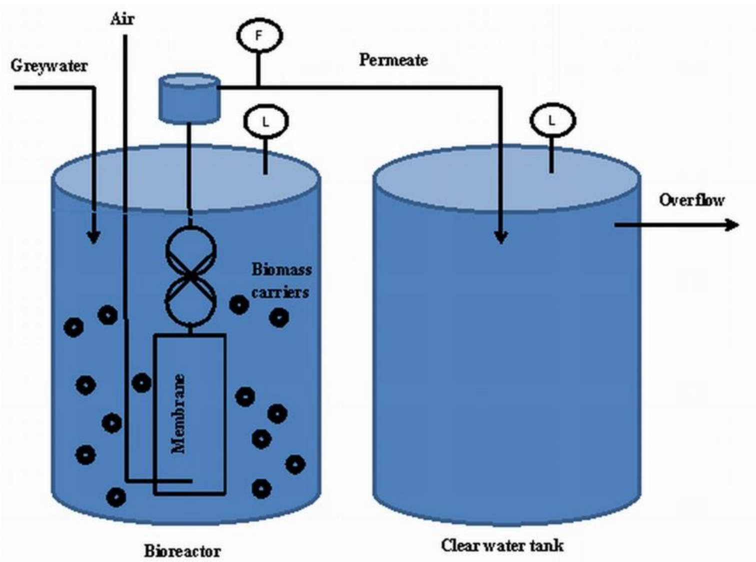


Figure 3 - 1: Flow diagram of Moving Bed Biofilm Membrane Reactor Pilot Plant

The membrane module was equipped with hollow fibre membranes with a nominal pore size of 0.2 µm. The module had an active membrane surface of almost 6 m². The fibres were wound up around a carrier cartridge in order to increase their packing density and save space. The fibres inside the cartridge were protected by a plastic cover sleeve from moving biomass carriers. The top and bottom of this cartridge had several rectangular openings with 20 x 5 mm which were small enough that the carries (cylindrical diameter 36 mm) could not pass. Nonetheless the overall area of the openings in the carrier cartridge was big enough that back-flushing and air scouring could remove accumulated sludge inside the cartridge sufficiently.

A low module construction increased the balancing volume in the bioreactor. A strong variation of arriving grey water is typical for single households; therefore the unit was designed for a quasi-batch operation. In other words the bioreactor worked continuously but membrane filtration started only at a time when it was assumed that as few as possible fresh grey water arrived at the treatment unit. By this the balancing volume of the bioreactor was worked off with the membrane filtration at a rather short time. This should ensure that after a grey water flush, e.g. in the morning, no untreated water is bypassed directly into the clear water tank. In consideration of typical grey water production in a single household which is usually in the morning, noon and in the late afternoon, the permeate extraction took place only two times per day between 4 – 6 a.m. and 4 – 6 p.m. The filtration time was adjusted on the actual membrane flux which reduced by fouling over time.

The membrane filtration system consisted of a submersed permeate and back-flushing pump. The membrane module itself could be operated in suction mode till -0.9 bar and in pressure mode till 3 bar according to the supplier´s information. Back-flushing, when applied, was made in a range of 20 – 25 L/m^{2}h. Back-flushing was started after 6 min filtration for 15 sec. During the tests it was further reduced to 15 min filtration, followed by 30 sec back-flushing. Aeration of membranes was switched off during filtration and back-flushing. So the membranes were operated in dead-end configuration throughout the tests. Chemical cleaning was made once after four month operation using diluted hypochlorite (Danklorix, Henkel).

The bioreactor was a completely mixed reactor in view of the biomass carriers. They could not settle due to their lower density compared to water – even with biomass on it. The carriers were moving through the reactor forced by aeration. The surplus sludge which was washed off from the carriers or was built in the sludge flocs in the suspension partly settled on the bottom. The reason was that bioreactor aeration was made via the membrane aeration diffuser. This diffuser was situated about 20 cm above the tank bottom and so the washed off sludge flocs, which entered this low turbulence zone, could settle. The settled sludge was siphoned off on the bottom by stopping the process and opening the drain valve of the bioreactor. The drained volume with a relatively high concentration of suspended solids was about 2 L per week only. By this, a total suspended solids (TSS) concentration between 100 – 500 mg/L could be maintained throughout the pilot tests. As pre-treatment an overflow screen with a mesh size of 1 mm was installed to remove hair, fibres and debris from the influent. The treated water was stored in a 350 L clear water tank which was equipped with a gravity overflow to the sewer system. The overall operation parameters of bioreactor and membrane filtration are listed in **Table 3 - 2**.

Table 3 - 2: Operating parameters of the MBBMR pilot plant

Moving bed bioreactor	**Design**
Bioreactor tank low level (L)	110
Bioreactor tank high level (L)	350
Hydraulic retention time (h)	24
Loading rate biofilm (kg BOD5/m²d)	0.004
Biofilm area (m²)	9.4
Volume of biomass carriers (L)	30
Aeration ON/OFF (min/min)	1/10, 5/10
Suspended solids in bioreactor (mg TSS/L)	100-500
Membrane filtration	
Membrane area (m²)	6
Pore size (µm)	0.2
Permeate cycle duration (min)	6 - 15
Permeate extraction (h/day)	1 - 4
Back-flushing duration (sec)	15 - 30
Air-scouring ON/OFF (min/min)	1/10, 5/10

In order to prove the constancy of the process for the application on single household several different conditions and loads were tested during the 10 month operation. This included start-up with and without seeding, high and low ambient temperatures (due to seasonal change of ambient temperature), low and high COD loading through cleaning and washing days, performance after 14 days' without influent due to vacation of inhabitants and finally hair colour removal efficiency.

As seed sludge either activated sludge from a waste water treatment plant or seed microorganism (Biofuture BFL 5600SS) were introduced in a quantity of 2 x 50 g respectively, equivalent to 0.45 g/L at low level of bioreactor (110 L) and 0.14 g/L at high level (350 L).

For the colour test one of the most common hair colours in the world was selected, the natural Henna colour. First a pre-mixture with 30 g colour and 100 mL warm water was made. This mixture was further diluted with 30 L. The dilution was directly filtered through the membrane to investigate the removal efficiency without supplementary bioreactor degradation.

2.2 Raw grey water characteristics and feeding

Synthetic grey water with a composition comparable with combined wastewater from showers, baths, washing basin and washing machine were used for the tests. The amount of ingredients was according to NSF standard for testing of onsite grey water treatment systems (**NSF/ANSI 350, 2011**). A daily mean amount of 4 people or 200 L, equivalent for 50 L/day and person, was fed within three cycles over the day. One day per week was chosen as cleaning day with additional cleaning agents in the raw water and two days were assumed to be washing days with additional washing powder and softeners. In order to simulate the strong variation of inflow 40% of raw grey water was introduced between 7 – 9 a.m. and 7 – 9 p.m., respectively. The remaining 20% was fed between noon and 1 p.m. **Table 3 - 3** shows the average influent concentration of the raw grey water parameters. The high deviation values indicate the increased loading during cleaning and washing days. The nutrient ration was COD:N:P 100:2.28:0.25, which implies rather limited nutrients available for the microorganism but is typical for grey water without kitchen waters. **Jefferson *et al.*, 2004,** also measured limited COD:N:P ratio for the bath, shower and hand basins sources were averaged at 100:2.25:0.06, 100:2.91:0.05 and 100:1.77:0.06 respectively.

Table 3 - 3: Chemical and physical raw grey water characteristics

Parameter	Unit	Raw Grey Water		Samples
T	**[°C]**	14.31	±4,01	74
COD	[mg O_2/L]	240	±101	65
BOD	[mg O_2/L]	168	±111	5
pH		8,41	±0.56	71
Turbidity	[NTU]	133	±101	74
Colour	[Pt/Co]	564	±397	74
Conductivity	[µS/cm]	439	±208	74
TSS	[mg/L]	72.4	±56.5	35
PO_4-P	[mg/L]	0.15	±0.16	58
NH_4-N	[mg/L]	0.58	±1.06	23
TN *)	[mg/L]	6.50	-	1

*) measured during normal days without cleaning and washing machine detergents

2.3 Analysis

Analyses during the 10 month trial were made for raw grey water, bioreactor suspension, 24h composite samples from the clear water tank and membrane permeate. The average values presented in this study were calculated as an arithmetic mean of the data collected at the different sampling dates. The standard deviation is indicated as +/- x. The measurements included BOD_5 (DIN EN 1899-2 / SOP 028), COD (DIN 38409-41-1 / SOP 027), TN (DIN EN ISO 10304-2 / SOP 046), NH_4-N (DIN 38406-5-1), TP (EN1189-6 / SOP 046), PO_4-P, TSS (DIN 38409-2 / SOP 034), pH (DIN 38404-5), Turbidity (Hach 2100P), Conductivity (Hach Sension156), Colour (Hach DR/890), as well as the membrane parameter transmembrane pressure (TMP) and specific flow called flux. *E.coli* was measured with HachLange paddle tester with an incubation time of 24 – 48 hours at 37 °C +/- 1 °C.

The concentrations of the suspended solids in the bioreactor was measured directly with a turbidity meter (Hach 2100 P) or by filtration over a 0.45 µm paper filter and drying at 105 °C. For the measurement of the biomass on the carriers, 10 pcs were randomly picked from the bioreactor and were weighed after drying at 105°C. After removing the dried biomass, cleaning and drying, they were weighed again. The weight difference was assumed to be the dry weight of biomass on one carrier. The mean weight of the carriers was multiplied with the total number of carriers (~880) in order to calculate the total biofilm mass. The F/M ratio was calculated with the mean concentration of BOD_5 (168 mg/L) multiplied with the daily amount of grey water (200 L) and divided through the total biofilm mass.

2.4 Calculation of energy consumption

The pilot unit was equipped with three big consumers, namely the common blower for bioreactor and membrane scouring (29 W), the permeate pump (40 W) and the back-flushing pump (70 W). The energy consumption was calculated with the installed power multiplied with the actual runtime of the equipment per day as kWh/day. For the study six different cases were investigated in order to find the lowest energy in consideration of bioreactor aeration (> 50 % oxygen saturation), membrane scouring and output of plant. Cases A1, A2 and A3 describe a low (5 L/m²h), medium (15 L/m²h) and high flux (30 L/m²h) status of the plant. High flux means that the daily amount of treated water was extracted very quickly through the membrane and therefore energy was saved for permeate and back-flushing pump. Low flux meant longer runtime and therefore higher energy consumption in total. The aeration at these cases was kept constant at 5 min ON and 10 min OFF cycles in order to compare the actual energy demand. In cases B1, B2 and B3 a rather low mean flux of 7.5 L/m²h, due to fouling of membranes, could be maintained continuously. It was used as constant low limit but aeration times of bioreactor and thus, the membrane, were then varied. In B1 the bioreactor was continuously aerated, in B2 33% and B3 only 9% of the total time. Cycle time was set to 1 – 5 min ON and 10 min OFF. B3 represented the final and in terms of energy demand the optimum set-up. This set-up was applied for the last 5 month of testing. The specific energy consumption in kWh/m³ of each case was also calculated as total energy consumption per day divided through the daily amount of treated grey water (200 L).

3 Results and Discussion

3.1 Treatment performance of bioreactor

The MBBMR pilot unit with a capacity of 4 people equivalent (200 L/day) was tested 10 month with synthetic grey water. Different operational conditions and stress situations typical for single households were simulated in order to verify the performance of the process. This included a start-up with and without seed microorganism, high and low influent COD loadings, and a seasonal change of ambient temperatures during the year, restart after vacation and the loading with natural hair colour. During the whole test period the biological, chemical and physical characteristics of the raw water, bioreactor suspension and treated water were measured. Furthermore, especially the performance of membrane filtration and energy consumption were also part of this study.

3.1.1 Start-up with and without seed microorganisms

The pilot unit was started with new biomass carriers without biofilm and without seeding of additional suspended microorganism. Although the synthetic grey water contained 0.5% of treated effluent from a waste water treatment plant and therefore some microorganism were introduced right from the start, almost no biological treatment took place inside the bioreactor at the beginning. So the first two weeks the plant worked more or less as direct membrane filtration plant. The removal efficiency of COD (73%, 6 samples), TSS and turbidity (99%, each 6 samples) was in average good which indicates that major parts of the organic load of the grey water ingredients were already particulate or built conglomerates in the water and could be filtered away. The results were slightly better than in previous studies with direct ultrafiltration of grey water with a reported reduction of 45-70% (**Ramon *et al*., 2004**) which could be a result of using higher transmembrane pressures in previous studies.

After two weeks additional 50 g activated sludge was fed into the bioreactor. Surprisingly, the process showed in the beginning after seeding poorer results than without biological degradation. In the first month after seeding the removal efficiency for COD was in average 70% (7 samples), for TSS 97% (3 samples) and for turbidity 98% (9 samples) which is slightly less than with direct filtration. Again one month later performance increased to COD (78%, 4 samples), TSS (99%, 10 samples) and turbidity (98%, 6 samples).

In a second trial seeding was tested with 100 g (2 x 50 g) microorganism adapted to grey water composition. This time the COD development after seeding was analysed daily to double-check the findings from seeding with activated sludge. Again right after seeding a peak in the COD values were measured. In **Figure 1** the development of this COD peak in the treated effluent is shown. The very low COD at the time of seeding were due to dilution of clear water which was filled into the bioreactor first. After seeding the COD values increase continuously. The peak reached the plateau after 10 days and stabilised after 18 days of treatment.

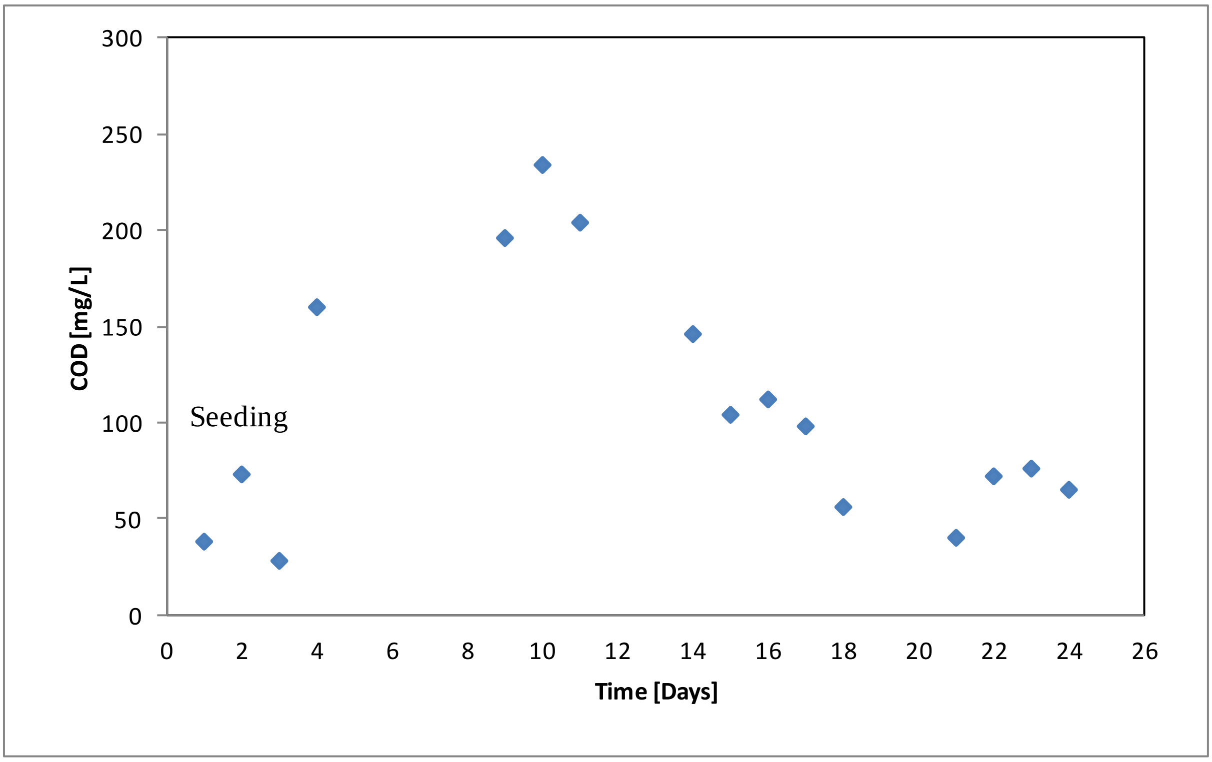


Figure 3 - 2: COD development in treated water after start up with seed microorganism

The peak right after seeding may result from adaption of microorganism and degradation of particulate COD on the membrane surface by the microorganism, which was formerly retained. The adaptation of the microorganism to the environment and growth on the biomass carriers finally stabilized the degradation and increased the performance. Specialized microorganism compared to activated sludge may reach stabilization faster but a peak cannot be avoided in both cases. In literature seeding of small onsite grey water unit was not subject to detailed studies so far. **Merz *et al.* (2007)** did not inoculate the AS-MBR demo-unit but also did not measure biological parameters the first days for comparison. The results may suggest that seeding of onsite grey water treatment including membrane filtration is not necessary at all. Any additional negative influence on the membrane performance apart from normal fouling similar among tests with and without seeding could not be verified.

3.1.2 Performance of bioreactor including low and high loading rates

During the 10 month operation of the MBBMR pilot plant the goal was to simulate the composition and production of grey water of a single household as best as possible. So two days per week the raw grey water included additional washing machine detergents and softeners and once a week bath room cleaning chemicals. Raw grey water was filled into the reactor in flushes in the morning, noon and late afternoon. Regular samples were taken from the raw water, treated water in the clear water tank and directly from the membrane outlet. The treated water parameters of 10 month testing are presented in **Table 3 - 4**. The mean and single sample maximum of the **NSF/ANSI 350 (2011)** effluent criteria for onsite residential and commercial water reuse treatment systems could be achieved (BOD_5< 10 mg/L, TSS < 10 mg/L, Turbidity < 5 NTU, *E.Coli*< 14 MPN/100mL, pH 6 – 9). The results showed excellent removal rates for BOD_5 (95%), Turbidity (99%), TSS (98%), Colour (91%), NH_4-N (78%) and PO_4-P (91%). The treated effluent further didn't contain any *E.coli* (4 samples). COD removal was in average at around 64% which is lower compared to literature values ranging from 86 – 99% (**Merz *et al.*, 2007, Jong *et al.*, 2010**). A reason could be the rather poor nutrient ratio in the feed grey water which was only in a range of COD:N:P 100:2.28:0.24.The oxygen saturation which should be above 50% according to certain guidelines **(fBr, 2005)** could not be achieved in average but would probably need only slightly more aeration time of the bioreactor.

Table 3 - 4: Treated water and permeate characteristics

Parameter	Unit	Clear water tank		Removal	Samples	Permeate		Samples
T	**[°C]**	13.21	±4,01		72			
COD	[mg O_2/L]	86.1	±39.93	64%	70			
BOD_5	[mg O_2/L]	8.40	±2.80	95%	5			
pH		8.34	±0.46		85			
Turbidity	[NTU]	0.98	±1.28	99%	75	0.99	±0.50	58
Color	[Pt/Co]	50	±23	91%	90			
Conductivity	[µS/cm]	453	±180		91			
TSS	[mg/L]	1.75	±1.85	98%	32			
PO_4-P	[mg/L]	0.01	±0.04	91%	59			
NH_4-N	[mg/L]	0.13	±0.25	78%	26			
TN *)	[mg/L]	1.35	±1.23	79%	3			
O2-saturation	%	44%	±12%		88			
Flux	[L/m²h]	12.9	±10.0		79			

*) measured during normal days without cleaning and washing machine detergents

The average flux was around 12.9 L/m²h during permeating phase. The flux relating to daily treated water was constant with 1.39 L/m²h. The flux during permeating phase was higher than reported in literature for conventional AS-MBR treating grey water. **Merz *et al.*, 2007** measured a mean flux of 8 L/m²h and **Huelgas & Funamizu (2010)** 9.17 L/m²h. However the flux was lower compared to BF-MBR plants with fluxes up to 50 L/m²h which were operated with combined waste water (**Leiknes & Ødegaard**, **2007)**. A reason could be the quite different composition of the grey water compared to combined waste water which may form sticky conglomerates of non-degraded soap and detergents in the bioreactor and therefore is subject to increased fouling on the membrane surface and pores. Another objection for a correct comparability to literature could be the discontinuous membrane operation in this study which was in a range of 2 – 4 hours per day only. **Figure 3 - 2** shows the flux development of the pilot plant during April 2012 – January 2013. Typically for membrane operation the membranes started with a high flux and sharply declined due to fouling and scaling of membrane. The decline slowed down after 20 days and finally stabilized at a rather low value between 5 - 10 L/m²h. The further decrease of the flux in August after 120 days in operation was due to a breakdown of the back-flushing pump. Therefore after 130 days a recovery cleaning with 0.2% NaOCl was performed. The flux could be restored. Similar was concluded by **Kraume *et al.*, 2010,** who also needed a recovery cleaning after 3-4 month to maintain sustainable operation.

The chemical cleaning didn't affect the bioreactor performance significantly. COD peaks could not be observed after cleaning. One reason could be the special membrane configuration which was used during the tests. The membranes were protected from the biomass carriers inside a cartridge with an enclosed sleeve with small openings on the top and bottom. During cleaning the chemicals were back-flushed into the carrier cartridge and were not released directly to the bioreactor. After one hour reaction time the permeate cycle started again and pumped the exhausted chemicals into the clear water storage. The diffusion of chemicals into the surrounding media through the cartridge openings was probably minor and didn't affect the biomass on the carriers.

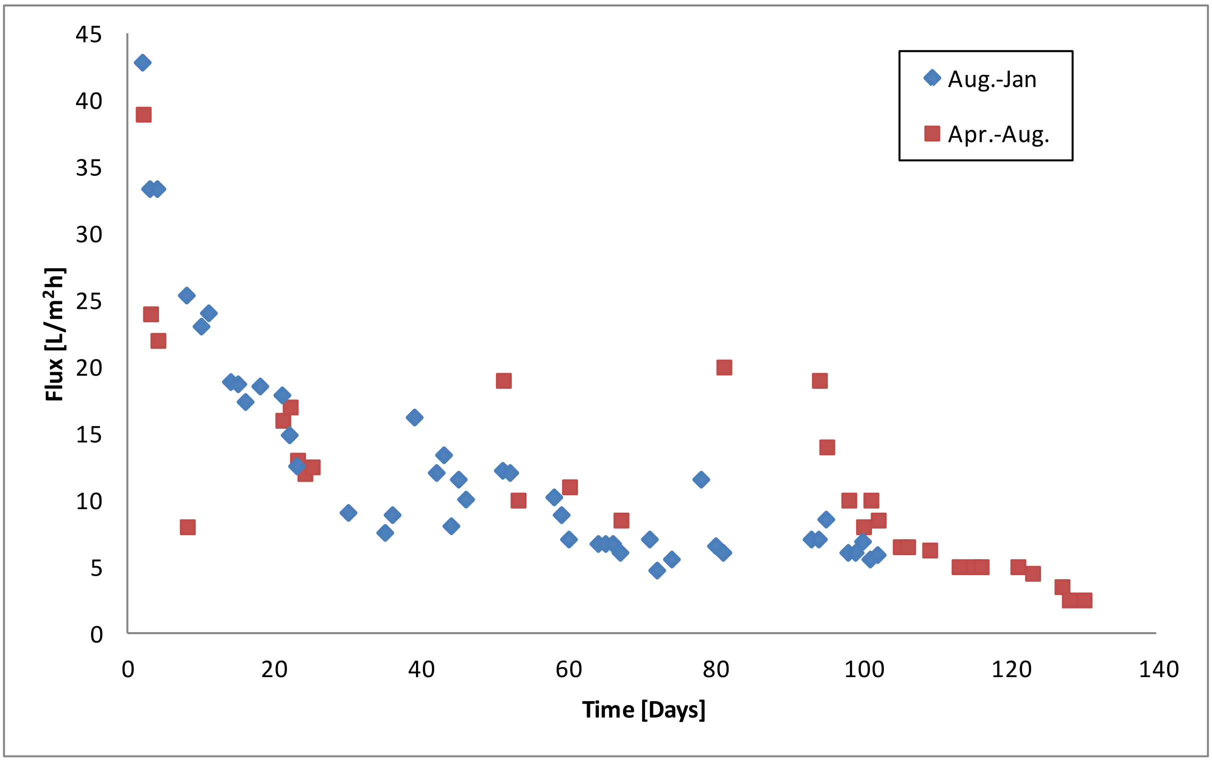


Figure 3 - 3: Mean flux of pilot plant during 10 month operation in permeating phase

New or recently chemically cleaned membranes achieved fluxes over 40 L/m²h while after several weeks the flux dropped significantly to less than 10 L/m²h. A high flux is preferable for the operation because treated water from the bioreactor can be withdrawn more quickly and therefore less energy is needed for the filtration process. The decline of flux maybe slowed down with increased scouring and back-flushing of membranes. On the other hand this would significantly increase the energy demand.

The strongly reduced back-flushing and air scouring restored the flux after each permeating phase but showed to be effective only within a short period of time. In the long run it turned out to be not strong enough to maintain the high initial flux which continuously decreased. The limited effect may be explained by a formation of an irreversible biofouling and scaling layer on the membrane with on-going operation. To verify the actual effect of back-flushing and intermediate air scouring dependent on the fouling condition of the membrane, the flux drop during the permeate cycles at a constant 0.2 bar pressure were observed and is presented in **Figure 3 - 4**. The diagram shows that the flux decline over time during one cycle reduces with lower flux and will be almost zero at a certain bottom line. The flux with this membrane configuration indicated to be finally stabilized is in a range of 5 L/m²h provided there is no failure of fouling control equipment. The sustainable or stabilized flux is lower than the critical flux which was measured in a range of 8 – 10 L/m²h.

Flux stabilization with strongly reduced fouling control at a low but constant value or operation of membranes on a deep sub-critical level may be the preferable operation method for small micro MBR systems. Then user interference in terms of maintenance and chemical cleaning could be probably lowered to a minimum. **Peter-Varbanets *et al.*, 2010**, also reported about this phenomenon with low pressure ultrafiltration membranes without any fouling control when different types of surface water and diluted wastewater were used. They could measure stabilized flux values between 4 and 10 L/m²h.

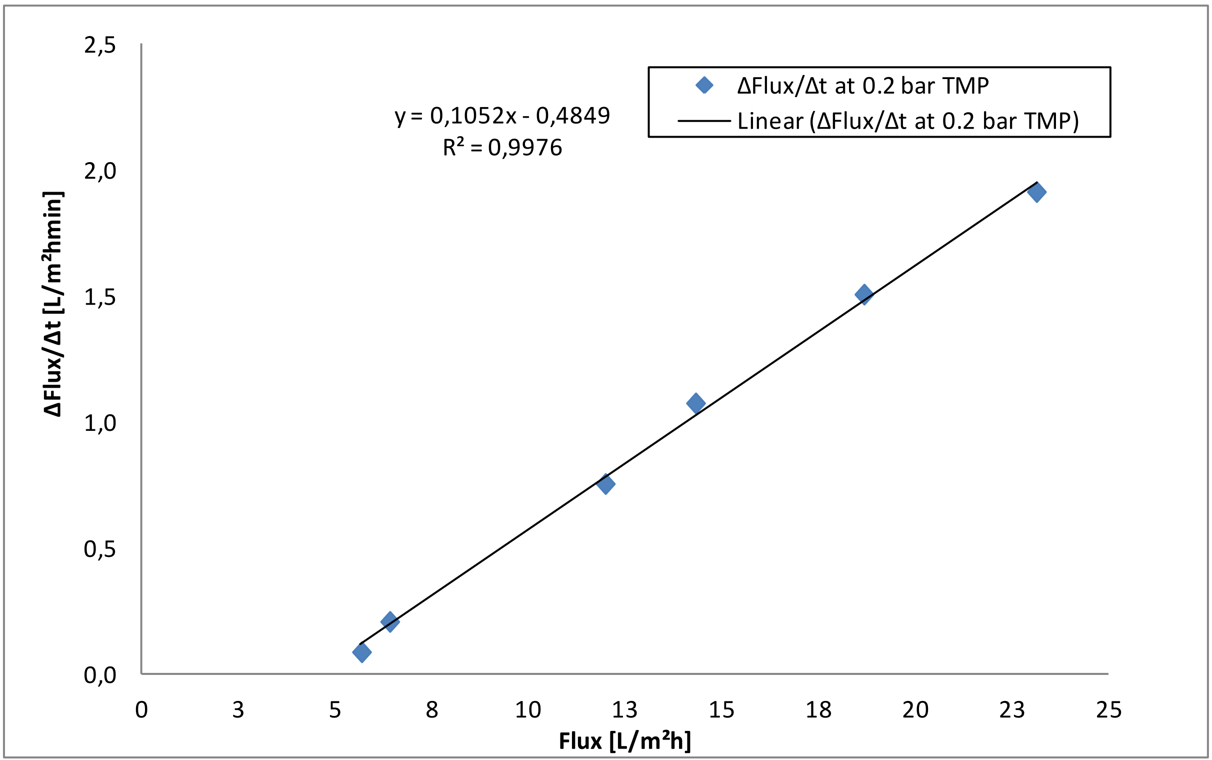


Figure 3 - 4: Flux drop during permeating phase of 6 min with constant pressure at different initial flux

3.1.3 Impact of high and low ambient temperatures

Although raw grey water is produced with almost the same temperature, the ambient temperature may vary strongly over the year. This could lead to operational problems especially in the case when the treatment unit is situated outside the main building.

The question was if this temperature change could significantly affect the bioreactor and membrane performance due to the 24 hour retention time in the process. The pilot tests were made from April 2012 to January 2013 in a non-temperature controlled facility. Ambient temperatures were between -10°C to +35°C. The raw grey water had a temperature in average of 14.31°C +/- 4.01. The results show that the ambient temperature had less influence on the bioreactor than expected. The retention time of about 24 hours was short enough to maintain a rather constant temperature in the bioreactor. Only during very cold days the system cooled down below 5°C. In summer the temperature in the bioreactor rose only on some days above 20°C. The minimum and maximum treated water temperature was 0.7°C and 25.6°C. The comparison of the treated water characteristics at low (< 10°C) and medium temperature (>10°C, < 20 °C) is presented in **Table 3 - 5**. The removal rates are quite similar at both temperature levels. COD and NH_4-N removal is slightly higher (59%/66% and 72%/83%), while TSS and PO_4-P removal is less efficient at higher temperatures (94%/89% and 97%/93%). pH and conductivity were not influenced significantly. While the biological processes and particle retention remain on a rather constant value, the membrane flux is strongly affected by the temperature. In cold temperatures the flux was in average only the half than at medium and high temperatures. The decreased viscosity of water which changes about 30% every 10°C is one part of the explanation and need to be considered in the design of these small units. The other issue is of course fouling which could have several explanations, is still hard to predict and is further subject to several unanswered questions. Numerous studies about membrane fouling were reviewed by **Drews (2010).** She summarized that possible fouling factors could be EPS (extracellular polymeric substances), HRT (hydraulic retention time), MLSS (mixed liquor suspended solids), PSD (particle size distribution), SMP (soluble microbial products), SRT (sludge age), TEP (transparent exopolymer particles), microbial population and rheology. Grey water composition in general will be strongly dependent on personal habits. Therefore the bioreactor performance and thus, the membrane fouling will probably vary strongly. In view of this uncertainness in the operation a low design flux concept in a range of 5 – 10 L/m²h for the single household systems might be advantageous.

Table 3 - 5: Performance of pilot unit with bioreactor temperature (TB) < 10 °C and 10°C < TB< 20°C

Parameter	Unit	T_B<10°C	Std-Dev.	Removal	Samples	10°C<T_B<20°C	Std-Dev.	Removal	Samples
COD	[mg O_2/L]	92.8	±25.15	59%	31	85.1	±48.4	66%	40
Flux	[L/m²h]	7.97	±2.82		22	16.6	±9.22		37
pH		8.45	±0.35		28	8.21	±0.53		38
Turbidity	[NTU]	0.83	±0.92	99%	29	0.95	±1.55	99%	39
Color	[Pt/Co]	45.1	±31.3	92%	28	51.2	±16.29	92%	39
Conductivity	[µS/cm]	356	±71.0		30	477	±177		39
TSS	[mg/L]	3.90	±1.30	94%	30	6.82	±14.78	89%	40
PO_4-P	[mg/L]	< 0.01	< 0.01	97%	10	0.01	±0.03	93%	11
NH_4-N	[mg/L]	0.07	±0.13	72%	29	0.16	±0.25	83%	27
O2-saturation	%	48%	±5%		12	0.46	±14%		10

3.1.4 Vacation

Another stress situation for the bioreactor is several days without influent due to a vacation of the inhabitants. The question was if the biofilm on the carriers will withstand several days without nutrients and microorganism will start again with degradation of fresh grey water without interruption. **Figure 5** shows the raw grey water and treated effluent COD right before and after the vacation. The first days after the system was restarted there was no change in the effluent COD values. A possible reason could be a dilution effect in the bioreactor because the microorganism degraded the complete COD during the vacation. Therefore a potential lack of degradation of the microorganism in the beginning resulted by long time without food is balanced by the retention time in the bioreactor. This could be an advantage of a combined bioreactor and balancing tank with more volume than actually needed for the biological process. Short peaks of increased COD in the effluent were measured 5 days after restart. After 4 more days the COD stabilized at pre-vacation level. The results may indicate that a vacation of several days may not affect the performance of a stabilized on-site MBBMR plant a long time and in general not significantly.

Besides the results are similar but not that distinctive compared to start-up of the system with and without seeding where COD peaks can be seen after some days of seeding. This could be an indicator that microorganisms are not wiped out completely but work limited in the beginning and grow again very quickly.

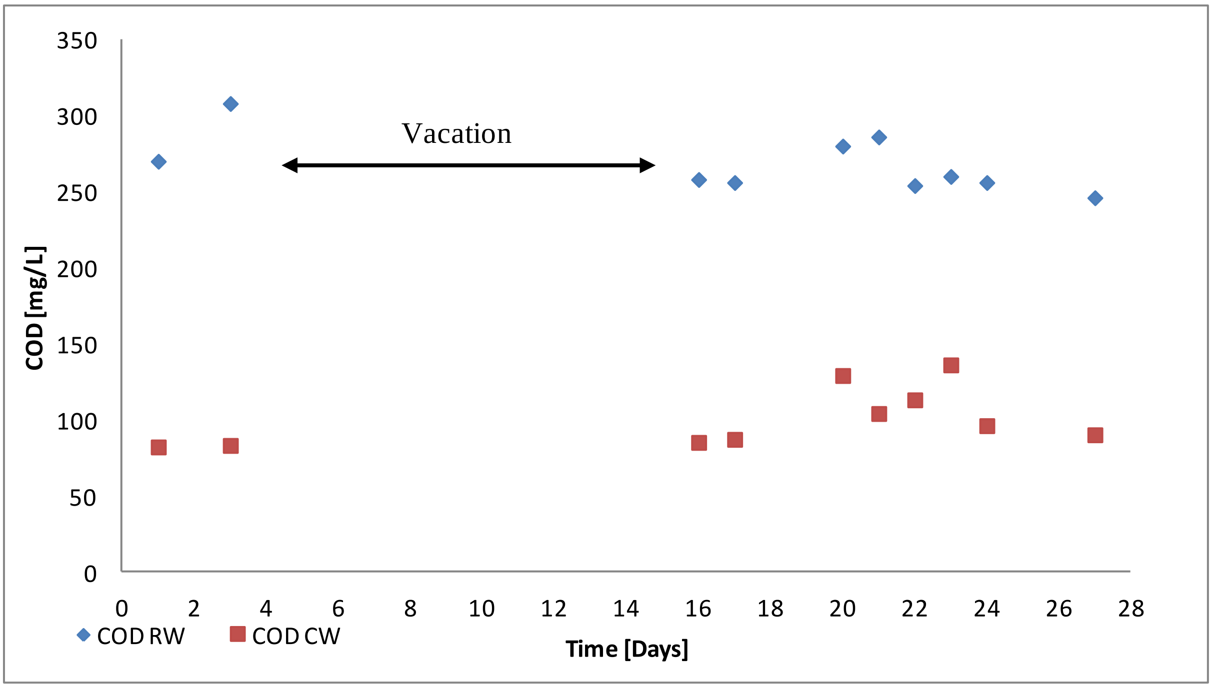


Figure 3 - 5: COD raw grey water (RW) and treated water (CW) before and after vacation

3.1.5 Colour removal (Henna natural colour)

Treated effluent from grey water treatment systems will be used for toilet flushing, irrigation, washing purposes and even for washing machines. Beside chemical stabilization and removal of pathogens, a very important factor for the widely acceptance is the aesthetic aspect. E.g. if hair colour is used and introduced into the wash basin, the colour should not be visible in the toilet flushing water anymore. Therefore a test with natural Henna colour was made with direct membrane filtration in order to investigate if colour can be removed sufficiently. The average colour removal efficiency was 78% (361 Co/Pt units) as mean within 1 hour filtration. A comparison of feed and permeate quality can be seen in **Figure 3 - 6**.

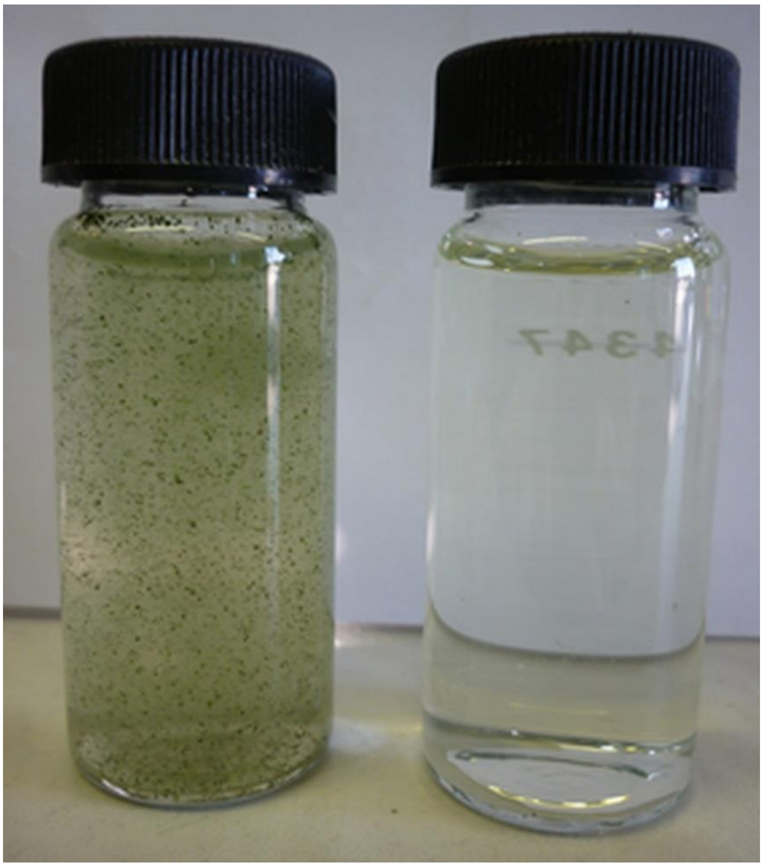


Figure 3 - 6: Comparison of feed and treated water when Henna colour is used

The treated water showed strongly reduced colour. The results may suggest that the membrane process alone is sufficient to provide acceptable removal efficiencies. With the additional degradation and dilution in the bioreactor the results might be even better and the water could be used for toilet flushing. Further colours were not tested and need to be part of further studies. However, a complete removal of colour, especially highly dissolved chemical colours is probably not possible at all. In this case the user of such a treatment system needs to be properly informed and advised in advance which colour are allowed to be utilized. This would be a similar issue when it comes to the usage of chemical cleaning agents for bathrooms and wash basins which might destroy the microorganism in the bioreactor.

3.1.6 Biomass development on carriers and in suspension

First biomass development on carriers could be visually verified after 5 weeks in operation. The biofilm thickness strongly increased till the end of the tests. A comparison of the biofilm thickness and structure on the carriers between week 5 and week 20 in operation is shown in **Fig. 3 - 7**. New carriers are black without any residues on the surface. Biomass growth started inside the carriers with a thin, light grey coloured film. With on-going operation the biofilm developed also at the other parts of the carriers and further grew even on the outside with highest aeration turbulence. The thickness of the biofilm generally increases strongly and colour changed to grey-yellow.

Figure 3 - 7: Comparison of the biofilm thickness and structure between after start-up (top left), week 5 (top right) and week 20 in operation.

After 10 month operation the actual biomass weight on the carriers was measured. The mean dry weight of the biofilm was 99 mg per carrier. Extrapolated on 30 L of carrier material there is a total of roughly 87 g of biomass which would be equivalent with an F/M ratio of 0.385 kg BOD_5/(kg day) related to mean loading of 168 mg BOD_5/L and 200 L grey water per day. The mean organic loading rate was 0.305 kg BOD_5/(m^3 day) or 0.436 kg COD/(m^3 day). **Merz *et al.* (2007)**, measured only 0.16 with grey water AS-MBR configuration, **Buntner *et al.* (2001)** with a low-strength waste water biofilm MBR process 1 – 3 kg COD/(m^3 day). Higher loading rates would be probably possible if bioreactor volume could be further reduced. In view of the carrier volume which is only 27 % of total bioreactor volume at minimum water level, this would be in fact possible. On the other hand higher loading rates are additionally restricted by the membrane module height and a certain hydraulic retention time which has positive effects in terms of load equalization needed for typical grey water flushes in the morning and evening.

The concentration of suspended solids in the bioreactor which could be defined as washed off biomass from the carriers and particles already present in the raw water increased only slowly during the tests and didn't exceed maximum concentrations of over 500 mg/L. **Fig. 3 - 8** shows that the concentration of suspended solids increased after a complete drain of the bioreactor in week 14 only from 150 NTU (~86 mg/L) to 250 NTU (~143 mg/L) within 7 weeks. The reason could be slow biomass growth in general which may be the result of limiting nutrients available (**Jefferson *et al.*, 2004**) and additionally that suspended solids settled permanently in the low turbulence zone between membrane aeration and bottom of the tank. The settled sludge was siphoned off once a week in a quantity of about 2 L/week. Peaks in the diagram indicate increased aeration flushes and partly detachment of already settled sludge from the bottom.

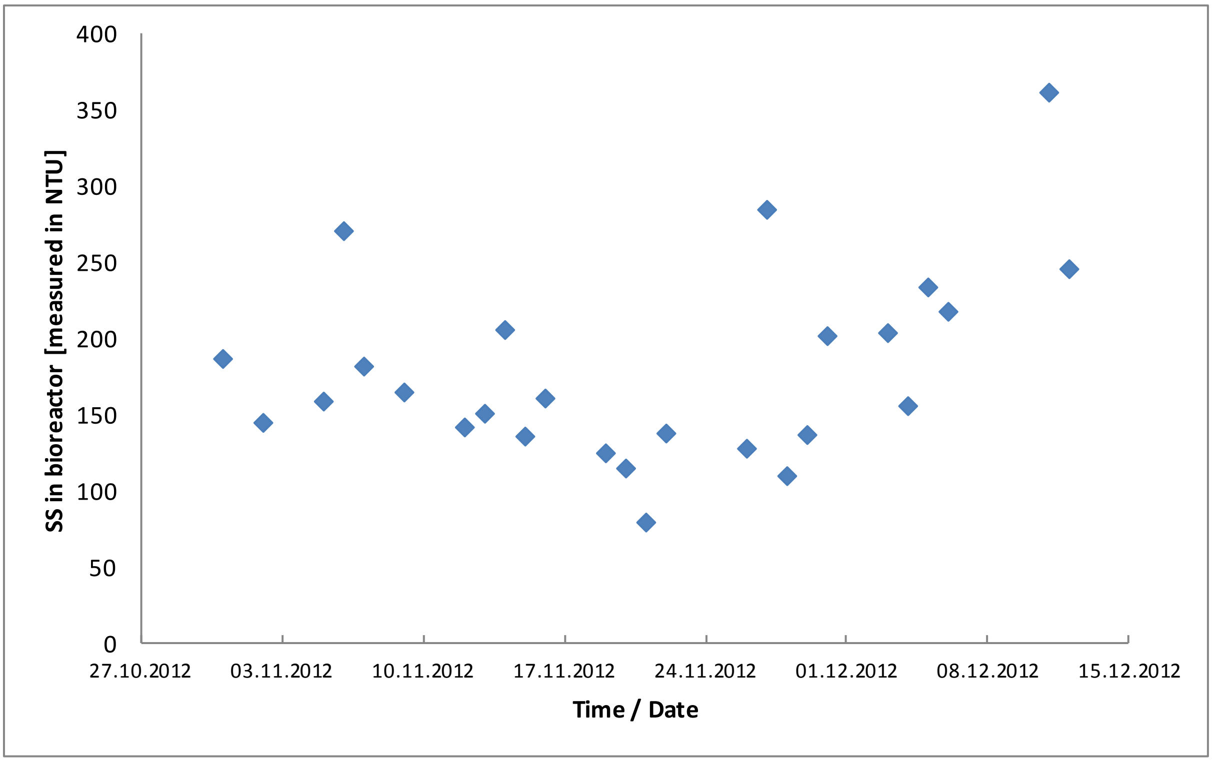
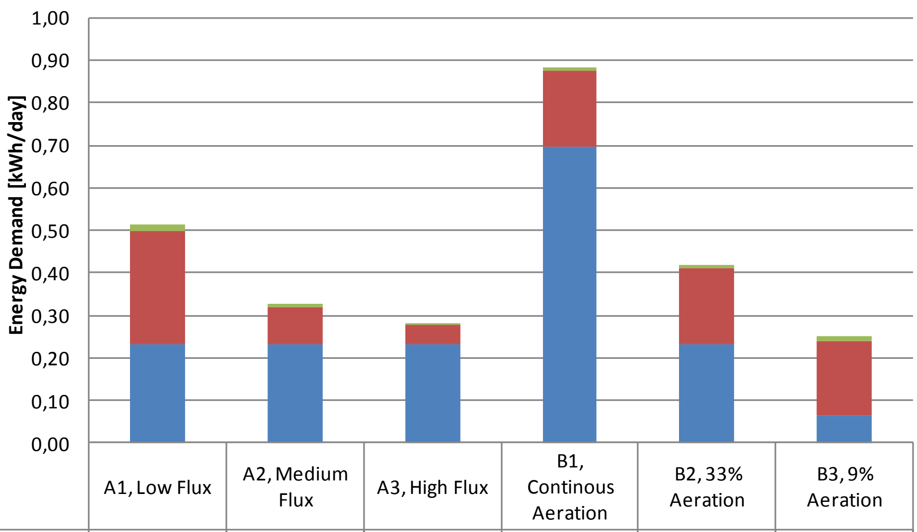


Figure 3 - 8: Development of suspended solids concentration in the bioreactor during 7 weeks of operation

Detailed analysis of surplus sludge is pending. Visual examination of the drained sludge showed a highly stabilized, black coloured and high density appearance with slight earthy smell.

3.2 Energy demand

In 10 month of operation the MBBMR plant went through different conditions were studies concerning the energy demand and operational costs were done. **Figure 3 - 9** shows the calculated energy demand according the runtime of the main consumers during the test period. Condition A1 – A3 represent the energy demand dependent on the status of the membrane fouling. If new or cleaned membranes (> 30 L/m²h, A3) were used, the main energy demand is needed for the aeration of the bioreactor. The energy demand for permeate extraction is reduced because of shorter cycle time respectively. On the other hand if membranes were severe fouled and flux is only about 5 L/m²h (A1) the main energy demand is due to permeate pumping and back-flushing. Continuous aeration (B1) didn't increase the sustainable flux but strongly increased energy demand of blower. Due to the low suspended solid concentration in the bioreactor the aeration of bioreactor and thus, the membrane could be further reduced to finally 9% of total time (B3) without facing problems with sludge accumulations on the membrane. At this aeration rate still enough oxygen was supplied to bioreactor and the oxygen saturation could be kept at almost 50% suggested by fBr guideline (**fBr, 2005**). The specific energy demand measured were A1 2.57 kWh/m³, A2 1.63 kWh/m³, A3 1.39 kWh/m³, B1 4.42 kWh/m³, B2 2.10 kWh/m³ and B3 1.26 kWh/m³. Studies for energy demand were not sufficiently made so far for on-site grey water MBRs. **Verrecht *et al.*, 2012** reported energy consumption of conventional small MBR plants for sewage treatment between 3 – 11 kWh/m³. A main reason for lower energy consumption in this study could be the strongly reduced scouring of membranes (< 9% of total time) and lower energy demand for biological treatment because of lower strength of waste water. Raw water lifting was also not included in this study. Anyhow, **Jabornig, 2013,** suggested in a feasibility study about small grey water MBR plants less than 1.5 kWh/m³ in addition to a reduction of investment costs in order to be cost-effective within 15 years. It depends on the actual additional membrane costs if a low but sustainable flux with increased membrane surface is preferable in opposite to increased operational costs due to high fouling control and high flux.



	A1, Low Flux	A2, Medium Flux	A3, High Flux	B1, Continous Aeration	B2, 33% Aeration	B3, 9% Aeration
Back-flushing	0,02	0,01	0,00	0,01	0,01	0,01
Permeate Extraction	0,27	0,09	0,04	0,18	0,18	0,18
Bioreactor + Membrane Scouring	0,23	0,23	0,23	0,70	0,23	0,06

Figure 3 - 9: Energy consumption of main equipment of MBBMR pilot plant

4 Conclusion

A combination of a moving bed biofilm reactor and membrane filtration in one reactor (MBBMR) which treated 10 month grey water with different loads and temperature simulating a single household achieved renowned international effluent reuse guidelines. A start-up with seed organism was not advantageous concerning degradation performance or membrane fouling. The strong variation of the ambient temperature and loading did not remarkably influence the treated water quality. Natural colour in grey water, which is a typical stress situation in single households, could be removed up to 80% and was almost invisible in the treated effluent. The energy consumption with the final set-up could be lowered to 1.26 kWh/m³ only, when a stabilized flux during permeating phase of 5 - 10 L/m²h is applied. The combined results lead to the conclusion that this reactor type could be feasible for grey water treatment for single households. Further studies concerning energy optimization through less membrane fouling and reduction of investment costs which would lead to faster amortization of these units are pending.

Nomenclature

AS-MBR	Activated sludge membrane bioreactor
BF-MBR	Biofilm membrane bioreactor
BOD_5	Biological oxygen demand
COD	Chemical oxygen demand
EPS	Extracellular polymeric substances
F/M	Food/Microorganism(Biomass)
HRT	Hydraulic retention time
LMH	Flux in L/m²h
MBBMR	Moving bed biofilm membrane reactor
MLSS	Mixed liquor suspended solids
NH_4-N	Ammonia
NO_3-N	Nitrate
ORL	Organic loading rate
PH_4-P	Phosphate
PSD	Particle size distribution
SMP	Soluble microbial products
SRT	Sludge age
T	Temperature
TEP	Transparent exopolymer particles
TN	Total nitrogen
TP	Total phosphorous
TSS	Total suspended solids

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CHAPTER 4:

A Novel Fixed Fibre Biofilm Membrane Process for On-site Grey Water Reclamation Requiring No Fouling Control

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A Novel Fixed Fibre Biofilm Membrane Process for On-site Grey Water Reclamation Requiring No Fouling Control

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Abstract

On-site grey water treatment and reuse in urban areas bears the potential to reduce huge quantities of wastewater and lower freshwater shortages. Until now dissemination of small, single household applications has been rather limited as simple and high quality water producing, but also cost-effective treatment units have not been developed so far. This paper proposes a new process, based on a concurrently working hollow-fibre membrane as fixed biofilm support and filtration device. Bioreactor characteristics, influence of different aeration rates, membrane flux development, as well as structure and composition of biofilm were monitored to evaluate the performance of the tested pilot unit. The introduced process achieved international water reuse guidelines, worked soundly and could, compared to conventional micro MBR, significantly reduce energy demand (<1.4 kWh m^{-3}). Fouling control by air scouring and chemical cleaning was not required once flux had stabilized. The biofilm analysis showed a porous, spongy-like structure. Microbiological investigation revealed a community of sheathed bacteria and nematodes that could play an important role in the flux stabilisation effect. In general, the study confirmed the suitability of the presented process for grey water treatment and provides valuable design data for future optimization and systematic analysis.

Keywords: flux stabilization, MBR, fixed biofilm process, membrane filtration, biofilm characterization

1 Introduction

The term grey water (GW) comprises a part of the residential wastewater, including effluents of showers, baths, wash basins (i.e. low-pollutant-load GW) and laundry- and kitchen sinks (i.e. high-pollutant-load GW). The amount of GW per person can represent up to 80% of total wastewater, e.g. Friedler & Hadari 2006, corresponding to 15 - 120 L d^{-1}, respectively (Morel & Diener 2006, Nolde 1999). Surveys showed that the treatment and reuse of GW would be generally accepted as long as four aspects, namely hygienic safety, aesthetic aspects, environmental tolerance and economical aspects are considered and fulfilled (Nolde 2005). One promising technology that could fulfil these criteria is the membrane bioreactor (MBR) technology. This process consists of a bioreactor used for the biological degradation of dissolved pollutants and a membrane-filtration-unit which acts as physical barrier for suspended activated sludge, suspended solids of any kind and pathogens or any other microorganisms. The combination of these processes ensures that the effluent after MBR passage is generally of superior quality and can be unrestrictedly reused in private households, e.g. toilet flushing or garden irrigation. Previous studies about GW treatment by MBR technology underlined the efficiency of this strategy to achieve highest reuse guidelines (Lesjean & Gnirss 2006, Merz *et al.* 2007, Kraume *et al.* 2009, Huelgas & Funamizu, 2010, Jong *et al.* 2010, and Bani-Melhem & Smith 2012). Despite of numerous advantages of MBR technology, there are still some major drawbacks of on-site GW reclamation. High investment costs and long payback periods for small-scale applications, high maintenance effort due to relatively frequent chemical membrane cleanings, high energy consumption for mixing and fouling control of membranes (i.e. removal of organic and inorganic precipitations on the membrane surface or in the pores through back-flushing, air scouring and chemical cleanings) are only some downsides of conventional MBR systems (Friedler & Hadari 2006, Humeau *et al.* 2011, Jabornig 2013, Jabornig & Favero 2013, Kraume *et al.* 2010, Merz *et al.* 2007). In other words, there is a huge effort needed to keep the membrane free from biofouling and scaling in order to maintain a continuously high flow through the membrane. In big municipal MBR plants fouling control strategies have been well investigated (Drews 2010, Judd 2006, Le Clech *et al.* 2006). The studies conclude that these plants would need sophisticated online monitoring of MBRs' chemical and physical condition. To our knowledge, detailed principles of fouling in micro on-site GW MBR applications have not been studied so far.

This study describes the approach to lower investment and running costs of micro MBR units by a new membrane bioreactor philosophy. A pilot unit was constructed which used a free floating hollow-fibre membrane as a concurrently working biofilm support and filtration device. Traditional fouling control of the membrane was therefore not performed. In literature this specific bioreactor set-up for the on-site treatment of GW has not been investigated so far. However it is known that flux stabilization can occur within one week without fouling control as reported by Peter-Varbanets *et al.* 2010 who used dead-end ultrafiltration membranes operated with lake, river and diluted wastewater-types. Phattaranawik and Leiknes 2011 further suggested as a recycling strategy to use non-intact hollow fibres membranes could be used as biofilm support. In a six month lasting monitoring campaign, the proposed combination of both processes was tested in order to prove three main hypotheses.

(i) Biodegradation of GW in a fixed-fibre biofilm membrane bioreactor does not need additional suspended activated sludge or biofilm growth carriers. Yet, the evolving biofouling on the membrane, i.e. the generated biofilm, is mainly responsible for degradation of dissolved pollutants. (ii) Sustainable flux without fouling control is feasible for on-site GW treatment and (iii) microbiological activity is responsible for the porous structure inside the evolving membrane fouling layer (biofilm).

The main aim of this study was to show the suitability of this process for GW treatment and to provide basic design data for advanced optimization and systematic engineering. Therefore the authors evaluated characteristics of the physico-chemical process conditions, the effect of three different bioreactor aeration rates on the process and the structure and composition of biofilm and biomass development.

2 Materials and Methods

2.1 Pilot Plant Set-Up

The pilot plant was constructed as depicted in **Fig. 4 - 1,** consisting of a completely mixed, aerobic cylindrical bioreactor tank with a diameter of 60 cm including a hollow-fibre membrane unit, concurrently acting as biofilm support and filtration device. The effluent of the membrane was collected in a treated water tank. Fouling control of the membrane by continuous air scouring, back-flushing and chemical cleaning was not applied during operation in order to enhance biofilm formation and to avoid excessive detachment of the latter. The whole unit was equipped with a programmable logic controller. The general operation parameters of the biofilm bioreactor and membrane filtration are listed in **Table 4 - 1**. Each tank had a maximum capacity of 350 L. The bottom water level in the bioreactor which stopped membrane filtration and ensured full immersion of membranes was set to 110

L, providing a minimum retention time of six hours and enough balancing volume for the influent peaks. Every day a volume of 100 - 200 L, equivalent to 50 L day^{-1} and person, was fed in three flushes discontinuously to the bioreactor to simulate natural inflow peaks in the morning (7 a.m.), noon and evening (7 p.m.) with a flow of 10 L min^{-1}. The bioreactor was operated as discontinuously fed process with independent inflow and effluent times. During all set-ups medium-load synthetic GW was freshly prepared (see **Table 4 - 2** for physicochemical characteristics). As pre-treatment of the influent GW and protection for the membrane fibres an overflow screen with a mesh size of 1 mm was installed to remove hair, fibres and debris. High-density polyethylene hollow-fibre membranes (0.95 - 1.10 kg L^{-1}), exhibiting a permanent hydrophilicity through a silicone finish were used. These fibres used for common filtration and biofilm support, were of symmetrically shaped with an outer diameter of about 0.4 mm, wall thickness of 50 µm and a pore size of 0.2 µm. Fibres were combined to bundles and potted with polyurethane resin into a main header. Each bundle consisted of about 1650 fibres (type C-MEM, Co. SFCU, Austria). Eight bundles connected on both ends formed a filtration unit of 12 m^2 surface. The specific area of the whole unit was calculated based on the outer diameter of the membrane fibres and amounted to 8777 m^2 m^{-3} which is significantly higher compared to biomass growth bodies, found on the market (320 – 500 m^2 m^{-3}), e.g. Jabornig and Favero 2013. The required biofilm support area was calculated for an expected biological oxygen demand (BOD_5) influent-concentration of mixed GW (without kitchen waters), as reported by the Association for Rainwater Harvesting and Water Utilisation Germany 2005 (187.5 mg O_2 L^{-1}). This represents a daily GW amount of 200 L and an organic loading rate of the biofilm of 4 g BOD_5 m^{-2} day^{-1}, resulting in a required biofilm (membrane) area of 9.4 m^2. Biofilm surface and flux were the two design bottlenecks for selection of the membrane area. Peter-Varbanets *et al.* 2010 reported on transmembrane-pressure independent final fluxes of < 5 L m^{-2} h^{-1} with low-pressure ultrafiltration membranes and no fouling control. As GW is entering the bioreactor in peaks, it is advantageous that there is no continuous permeate withdrawal in order to avoid untreated GW from a flush entering the clear water tank. Up to 8 hours per day permeate was filtered through the membranes. The remaining time the plant was aerated only and worked as a batch reactor. Each membrane bundle was connected to one vertical main header which was also immersed in the bioreactor. Fibres of the bundles floated horizontally and due to this construction and their relatively low density could not settle during operation. The bioreactor aeration device consisted of a fine bubble ceramic diffuser (type Resun 20 x 50 mm, Co. Resun, China) which was situated at a distance of 20 cm on the opposite side of the bundles in tank. Membrane bundles were not directly aerated to preven strong shear-stress that could lead to biofilm detachment, but aeration provided enough air for bioreactor mixing and oxygen transfer. The initial air flow was set to 4.6 L min^{-1} or 0.38 L min^{-1} m^{-2} membrane area, respectively, which was previously suggested by Phattaranawik and Leiknes 2011**.** In this study, influence of aeration on biodegradation, flux and fouling was investigated by different air-flow-rate regimes, listed in **Table 4 - 1** (set-up A1, A2 and A3). Permeate withdrawal was performed with an inline centrifugal pump (type AL1, Co. Intewa, Germany) which was automatically started two times per day in between 3 – 7 a.m. and 3 – 7 p.m. Filtration time was adjusted to the actual membrane flux and so permeate pump runtime was usually lower in a range of 2 - 8 hours d^{-1}. Transmembrane pressure (TMP) was kept constant at 0.3 bar. Surplus sludge from the reactor-bottom was drained during the out-of-tank wash of the membranes. The effluent was stored in a 350 L treated water tank which was equipped with a gravity overflow to the sewer system.

In order to prove the constancy of the process for on-site single household application several different conditions were tested during the 6 month operation. This included start-up without seeding, high and low ambient temperatures (due to seasonal change of ambient temperature) and different aeration rates (set-ups A1, A2 and A3). Each set-up lasted for at least 30 days. The first set-up started after a four weeks start-up phase. Set-up A1 and A2 were repeated. Data from A2 repeat were shown due to technical problems with pump. There were no operation breaks between each set-up. Longer runtimes of each set-up were not considered because flux stabilization was expected within 7 days which is in accordance with reported stabilization times in literature (Peter-Varbanets *et al.* 2010)**.** After each set-up the membrane was removed from the bioreactor and washed gently with clear water in order to restore membrane flux. By this procedure deposits and excess biofilm on the membrane were removed. Chemical cleaning was not applied to maintain the biological activity for a fast recovery of the biofilm. Integrity testing of membrane was performed with 0.5 bar pressured air after each restart. Visible bubbles appearing from the immersed module would indicate membrane damage or open connections.

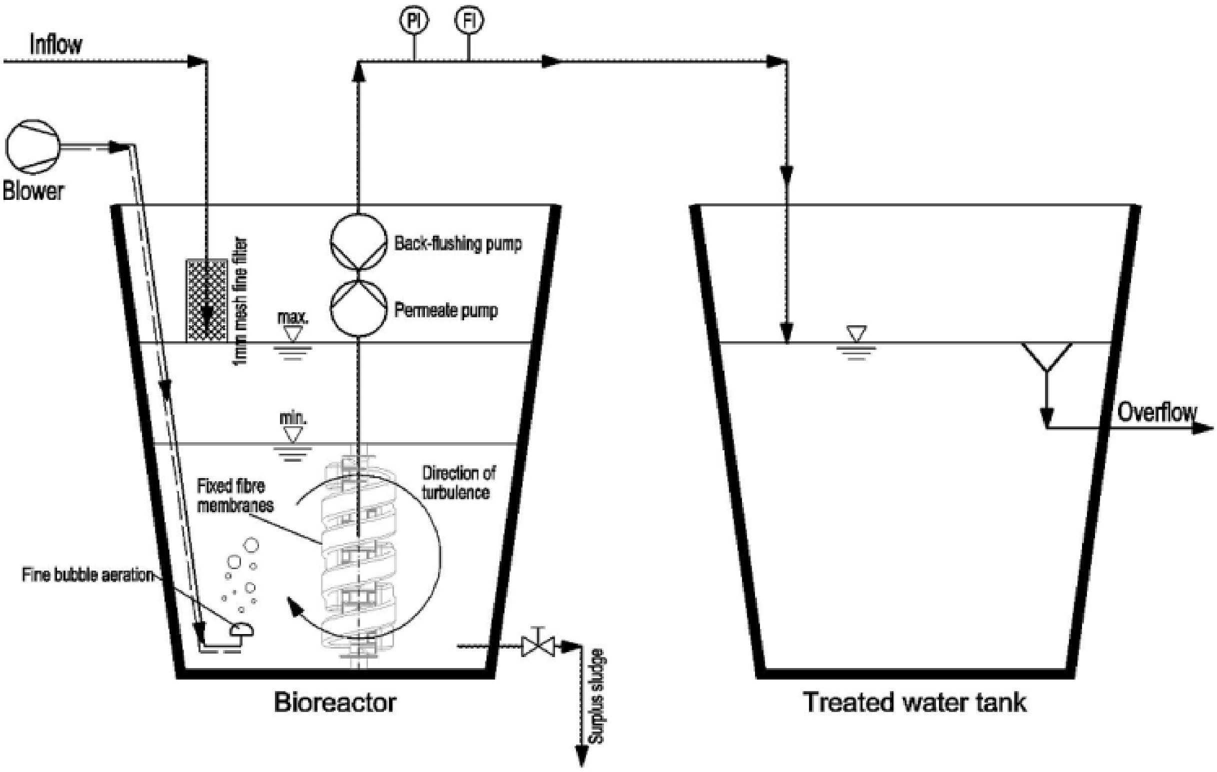

Figure 4 - 1: Schematic diagram of the fixed-fibre biofilm membrane pilot plant

Table 4 - 1: Overview of operating parameters of the bioreactor and membrane module

Set-up		**All**	
Top water level volume bioreactor [L] :		350	
Bottom water level volume bioreactor [L] :		110	
HRT dependent on filling level [h]:		6 - 24	
Permeate cycle duration [min]:		15	
Runtime permeate extraction depending on actual flux [h day^{-1}]:		2 - 8	
Set-up	**A1**	**A2**	**A3**
Aeration rate [L min^{-1}]:	2.3	4.6	6.8
Specific aeration rate [L m^{-2} min^{-1}]:	0.19	0.38	0.57

2.2 Raw water characteristics

Grey water consisted of shower gels, shampoos, tooth paste, conditioner and deodorants and was freshly prepared and mixed with potable water in a raw water tank (200 L). The amount of wastewater components was in agreement with NSF international standard for testing of on-site GW treatment systems (NSF/ANSI 350 2011). At the beginning a start-up phase lasting 30 days was allowed for initial biofilm formation. Microbiological contamination (e.g. human skin) of natural GW was simulated with 2 L of secondary effluent of a wastewater treatment plant, which was added to every 100 L fresh mixture. The measurements which were taken from the treated water tank (24 hour composite sample) or permeate pipe (grab sample for turbidity only) were done daily in the morning, except for weekends. The analyses included COD (Hach Lange dichromate method), total nitrogen (TN, Hach Lange method 10071), ammonium (NH_4-N, Hach Lange method 10023), phosphate (PO_4-P, Hach Lange method 10127), pH (Hach Lange SensION156), temperature (Hach Lange SensION156), turbidity (Hach Lange 2100P), conductivity (Hach Lange SensION156) and colour (Co/Pt, Hach Lange DR/890). Filtration pressures and -flow were measured with digital online sensors (type Tecsis P3276). Total aerobic bacteria and total coliforms were visually measured with Hach Lange paddle tester (No. 2610910) with an incubation time of 24 hours at 37 °C ± 1 °C. The concentration of suspended solids and settled sludge of the bioreactor bottom were measured by filtration of 100 mL (0.45 µm paper filter, Sartorius type 388) and drying at 105 °C. The deposit and biomass on the membrane was measured by the same principle. The weight difference of unused and already used ones was extrapolated on the whole number of fibres.

2.3 Analysis of biofilm sample

Biofilm samples including membrane material were taken from the pilot plant 10 weeks after start-up in set-up A3 and were microscopically analysed in wet state regarding structure and microbial composition.

Stereo- and light-microscopy

The general morphological overview of the biofilm, covering the membrane fibres, was obtained via stereo-microscopy (Nikon SMZ 1500). More detailed investigation on biofilm structure and visualisation of protozoa and microorganisms was achieved by light microscopy (Nikon Optiphot-2) using the NIS Elements D 3.0 software. To be able to differentiate between actual biomass and organic matter, the DNA of microorganisms was stained with

DAPI (4.6 diamidino-2-phylindole). Therefore, prior to fluorescence microscopy, specimen (ca. 20 µL liquid sample plus membrane/biofilm particle) were overlayed with 10 µL of DAPI-solution (1 µg mL^{-1}) and incubated for at least 10 minutes in the dark. To further allow for a natural image acquisition of the biofilm specimen - i.e. immersed in GW solution and not squiched by the cover slip - an elevated frame (nail varnish) was prepared around the sample onto which the cover slip was placed.

Confocal laser scanning microscopy

Image acquisition was performed with a confocal laser scanning microscope (Leica CLSM, SP5II, Leica Microsystems, Germany) and the Leica Application Suite Advanced Fluorescence software. Signals of total biofilm (organic matter; i.e. DAPI negative) were captured for the wavelength range of (excitation 405 nm; 546 nm – 608 nm) and are shown in yellow (false colour). Biological fluorescence (DAPI positive) was excited at 405 nm and detected between 419 and 466 nm and is displayed in blue (false colour). Images were acquired in sequential mode with a HC PL Fluotar 10X lens and minimum pinhole settings. Several areas of the sample were captured via z-stacked images (depth 11.73 µM ± 0.85 µM; 21 slices) and combined to a final maximum projection image to allow for higher structural resolution. Additionally single layers of this stack were used for subsequent biofilm/biomass ratio evaluation.

Image analysis

To determine the ratio of biofilm/biomass and the mean porosity of the biofilm structure, image analysis was performed with the Fiji software (www.fiji.sc). The ratio was calculated as quotient of total area of yellow signals (DAPI negative) and total area of blue signals (DAPI positive) from three layers (top, middle, bottom) of each stack.

Porosity [%] was assumed as background (neither yellow, nor blue signal) of maximum projection images. Prior to analysis background noise of the two colour channels (yellow, blue) was reduced.

2.4. Statistical Analysis

The average values presented in this study represent arithmetic means of the data collected at the different sampling dates. The standard deviation is indicated as ± x. The COD analysis of different set-ups further depicts median, min./max. -values, the lower- (25% quantile) and higher quartile (75% quantile). The Kruskal-Wallis test including Mann-Whitney U pairwise comparisons (Bonferroni corrected) was performed to detect statistical differences of the investigated set-ups A1, A2 and A3 regarding COD- and ammonium concentration (Zar 1986). Prior to this, normal distribution was checked with the Shapiro-Wilk test (Shapiro and Wilk 1965). All statistical analyses were performed with the PAST software (Version 2.17) (Hammer et al. 2001).

3 Results and Discussion

3.1 Raw water characteristics

Table 4 - 2 lists the average influent concentration of physicochemical parameters. Some parameters of the NSF international standard could not be achieved during the tests even though the dosing recommendations of ingredients were strictly followed (slightly lower temperature and slightly higher turbidity and pH). This could also be a result of different brands of moistures or shampoos that were used as wastewater components. The actual COD:N:PO_4-P ratio was 100:0.98:0.26 and represented a rather low and maybe limiting nutrient ratio for biological degradation compared to typical domestic wastewater with COD:N:P 100:9.30:1.63, e.g. Metcalf & Eddy 2003. Anyhow, e.g. NSF/ANSI 350 2011 suggest in their recommendations 100:1.33:083 while in natural GW Jefferson *et al.* 2004 reported of COD:N:P ratios of 100:2.25:0.06, 100:2.91:0.05 and 100:1.77:0.06 for the bath, shower and hand basins sources, respectively. Total aerobic bacteria (37°C) were measured in the feed with $10^5 – 10^7$ cfu mL^{-1} and total coliforms with $10^3 – 10^5$ cfu mL^{-1}.

Table 4 - 2: Main chemical and physical raw water parameters compared to recommended NSF values

Parameter [Unit]	Raw water tank		(n)	NSF
T [°C]	20.82	± 4.29	48	25 - 35
COD [mgO_2 L^{-1}]	252.9	± 30.23	15	200 - 400
pH	7.88	± 0.30	49	6.0 – 7.5
Turbidity [NTU]	94.47	± 54.41	53	30 - 70
Colour [Pt/Co]	428	± 193	53	-
Conductivity [µS cm^{-1}]	431	± 42	49	-
TSS [mg L^{-1}]	87.21	± 47.20	53	50 - 100
PO_4-P [mg L^{-1}]	0.66	± 0.28	14	-
NH_4-N [mg L^{-1}]	0.10	± 0.07	15	-
TN [mg L^{-1}]	2.49	± 0.36	7	3 - 5

3.2 Long Term Treatment Results

The fixed-fibre biofilm bioreactor was operated for 6 months with different set-ups and a varying daily raw water volume of 100 – 200 L. Measurements of physicochemical parameters were made in regular intervals from composite samples of the bioreactor, the membrane permeate and the treated water tank. The treated effluent characteristics of the complete test period are listed in **Table 4 - 3**. Generally, the process and effluent quality showed to be stable as long as no malfunctions of equipment or interruption in operation occurred. The average COD of treated GW was reduced from 253 mg L^{-1} to 77.9 mg L^{-1} reflecting a removal efficiency of roughly 70%. Studies with conventional GW AS-MBR processes reported rates of 85 - 96% (Kraume *et al.* 2009, Merz *et al.* 2011). Moving bed membrane bioreactors fed with GW could achieve a comparable result with a decrease of 240 to 84 mg COD L^{-1} and < 10 mg BOD_5 L^{-1}, e.g. Jabornig & Favero 2013. In international urban water reuse guidelines, e.g. NSF/ANSI 350 2011, certain COD limits are not mandatory but BOD_5-limits of 10 mg L^{-1} are common. With the implemented aeration rate of 0.57 L m^{-2} min^{-1} in set-up A3 this limit could finally be fulfilled. The final steady flux during all campaigns, as shown in **Figure 4 - 2,** seemed to stabilize in a range of 1 – 2 L m^{-2} h^{-1}. With eight hours runtime, the selected 12 m² membrane area allowed for a maximum permeate volume of roughly 200 L day^{-1}. Thus, measured flux is significantly lower than in conventional MBR treatment but is also not entirely comparable because of the different operation philosophy and lack of fouling control. The operation of membranes in a bioreactor without fouling control has not been investigated so far and therefore no comparable fluxes were available. For the investigation and confirmation of long term stabilization of flux, a trial over several months is pending. In direct membrane filtration slightly higher fluxes were measured by Peter-Varbanets *et al.* 2010. They monitored flux stabilization between 5 - 10 L m^{-2} h^{-1} with a dead-end membrane (flat sheet) filtration process operated with different river-, lake- and diluted wastewaters, but observed stabilisation also after a similar duration (one week). Here, chemical cleaning was not carried out during the 6 month period and might not be necessary at all, once flux remains constant after stabilization. In conventional MBR treatment, frequent cleanings, at least every 3 months, are reported for similar GW composition (Jabornig & Favero 2013, Kraume *et al.* 2009). Operation without chemical cleaning and fouling control would significantly lower investment costs, maintenance effort and would additionally reduce oxidative stress of bioreactor and membrane material. Thus, less equipment and membrane replacement costs can be expected. The power demand of the set-up A3 with the highest aeration rate and longest runtime of permeate pumps was 1.4 kWh m^{-3} which is still only half of the average consumption of GW treatment units available on the market (Jabornig 2013). Additionally, no consumables (e.g. coagulants or cleaning chemicals) were necessary during the six month testing phase. The regular integrity test with pressured air after each restart confirmed the durability of the selected membrane. During all tests ascending air bubbles from the membrane or from connections could not be discovered.

Table 4 - 3: Average long term treated water characteristics of pilot plant

Parameter [Unit]	Treated water tank		Removal Rate	(n)
Temperature [°C]	20.6	± 4.77		54
pH	7.93	± 0.34		52
Turbidity [NTU]	2.47	± 2.00	-97%	56
Colour [Pt/Co]	57	± 33	-87%	56
Conductivity [µS cm^{-1}]	441	± 24		52
COD [mg O_2 L^{-1}]	77.9	± 26.9	-69%	55
TN [mg L^{-1}]	0.99	± 0.61	-60%	7
NH_4-N [mg L^{-1}]	0.05	± 0.08	-51%	49
PO_4-P [mg L^{-1}]	0.14	± 0.15	-78%	47
TSS [mg L^{-1}]	0.96	± 1.25	-99%	56
Flux in permeating phase [L $m^{-2}h^{-1}$] *)	4,1	± 5.5		53
O_2-saturation [%]	35.6%	± 15%		53
Permeate Turbidity [NTU] *)	1.38	± 0.64	-99%	54

***) grab sample**

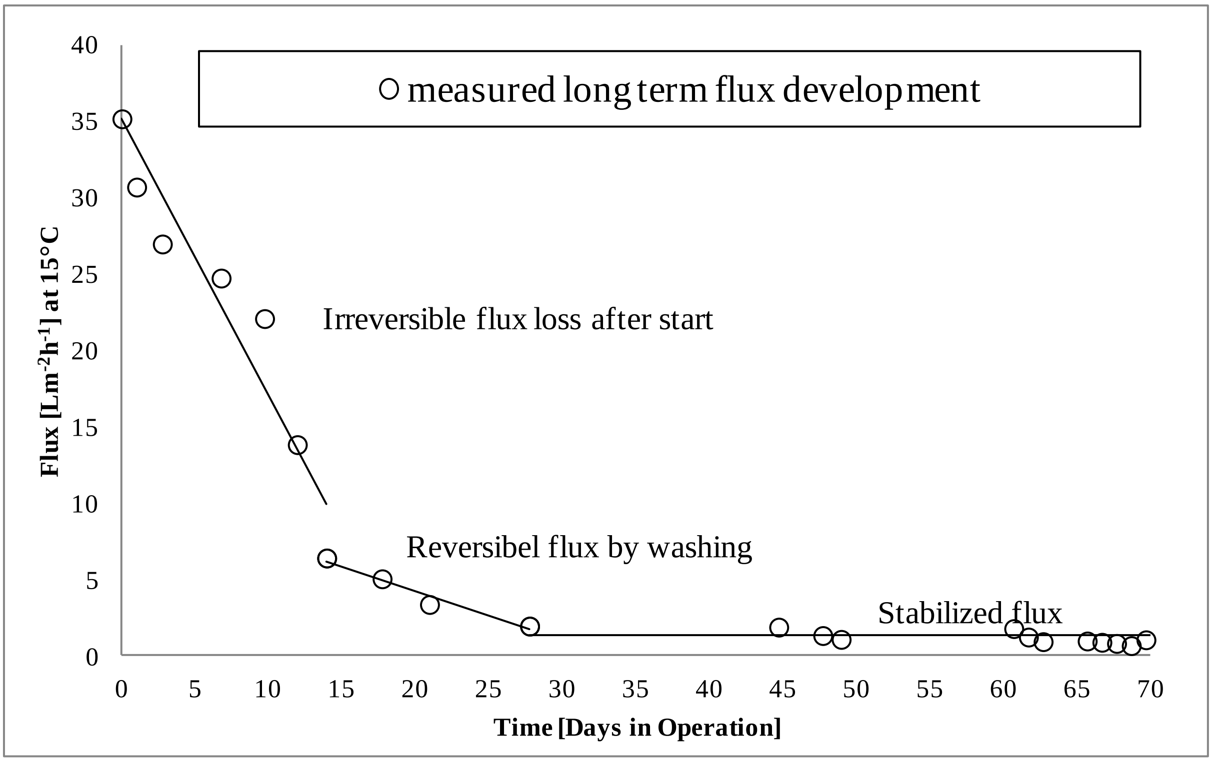

Figure 4 - 2: Three stages of long-term flux stabilization of the fixed-fibre biofilm membrane reactor adapted from actual measured values by linear regression. First flux decrease describes the irreversible loss by pore blockings after start-up. The second flux decrease during operation is reversible by out-of-tank washings and the third trend represents the final flux bottom line after stabilization.

3.3 Influence of Aeration Rate on Degradation (Set-up A1 – A3)

After a start-up phase of 30 days three different air flow rates were applied to determine the influence of aeration on the process. The efficiency of biological degradation was measured with COD as indicator parameter at regular intervals (i.e. daily or every second day). **Figure 4 - 3** shows the impact of the three tested aeration rates on the treated GW COD concentration. For final COD concentration a significant difference among the three set-ups was detected ($p=0.0474$). Further pairwise comparison through Mann-Whitney U test revealed that these significant discrepancies lay between experimental set-up A1 and A3 ($p=0.019$). With aeration regimes of 0.19 and 0.38 L m^{-2} min^{-1} mean values of COD were distinctly higher than compared with 0.57 L m^{-2} min^{-1}. Average COD was 58 ± 9 mg L^{-1}. Median values for set-up A1, A2 and A3 were 70.5, 78.5, 59.0 mg L^{-1}, respectively. A higher standard deviation of COD effluent concentration at lower aeration rates was also expressed in the 25 % and 75% quantiles that ranged from 64.8 – 83.3 (A1), 55 – 96.8 (A2) and 52 – 63 mg L^{-1} (A3), respectively. Although min-values were almost equal with 48, 44, and 46 mg L^{-1}, the max-value strongly increased in A1 and A2 compared to A3 with 135, 116 and 74 mg L^{-1}

indicating limited oxygen supply and therefore an unstable biodegradation. The COD measurement was confirmed by the oxygen saturation in the bioreactor. Mean oxygen saturation in the bioreactor during the aeration rates of 0.19 and 0.38 L m^{-2} min^{-1} was 31.6% and 28.9%, respectively, while with 0.57 L m^{-2} min^{-1} it increased to 66.1%. The sharp increase of saturation would point out that more oxygen is available as actually needed and therefore the ideal aeration rate lies between 0.38 – 0.57 L m^{-2} min^{-1}. Phattaranawik & Leiknes 2011 reported of similar aeration rates in a range of 0.33 – 1.67 L m^{-2} min^{-1} but used different raw water with approximately 25 % higher COD values. Lower or missing aeration (data not shown) had strong negative effects on COD degradation and flux, respectively. Higher aeration rates neither enhanced bioreactor performance, nor influenced flux and would just increase energy consumption. Ammonium concentration in the raw water was very low with values under 0.2 mg L^{-1}. The average nitrification efficiency was 51%. The total nitrogen removal was in all three settings in an acceptable range of 60%. No significant differences between the aeration rates were measured for both parameters (p=0.0978).

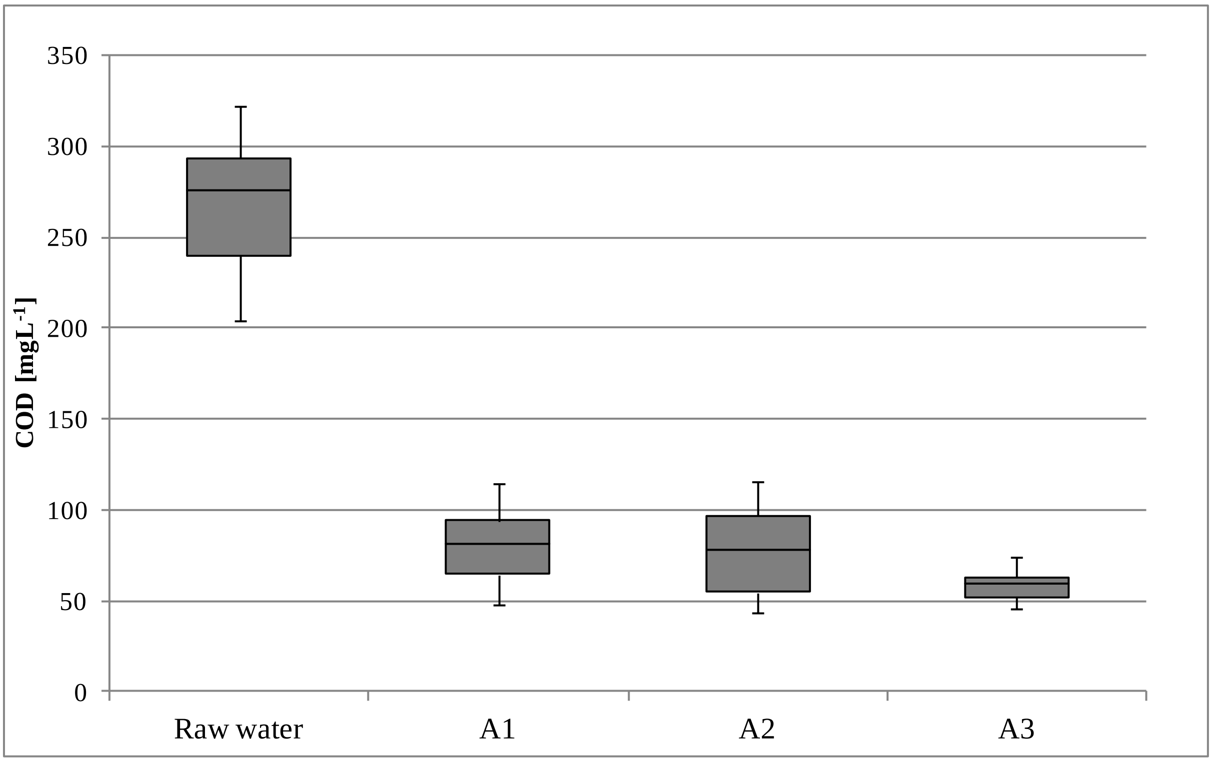

Figure 4 - 3: COD removal efficiency depending on aeration rates (A1 0.19 Lm-2min-1, A2 0.38 Lm-2min-1, A3 0.67 Lm-2min-1). Bar chart analyses of each set-up including median, 25% and 75% quantile and whisker.

3.4 Influence of Aeration Rate on Biomass Development (Set-up A1 – A3)

In this study the membrane was not directly aerated in order to avoid excessive sludge detachment comparable with fouling control in common MBR operation. Convective turbulence and microbiological activity was solely responsible for biofilm structure and thickness. Phattaranawik & Leiknes 2011 measured a very thin biofilm on a directly aerated biofilm membrane and concluded a high oxygen transfer rate compared to conventional growth bodies. Although in this study the biofilm thickness was higher, limited oxygen transfer was not assumed. The reason may be due to the simultaneous filtration through the membrane. While in conventional biofilm processes oxygen and nutrient transport is mainly due to diffusion, in this study the inner layers of the biofilm were fed with additional oxygen saturated water by filtration. Besides biological parameters, also accurate filtration time for this new membrane process need to be considered in order to maintain sufficient oxygen- and nutrient-transport into the biofilm. Excess sludge formation and detachment from the membrane fibre was not studied in detail, but could be due to microbiological activity. The excess sludge settled along with other suspended solids from the raw water at the bottom of the bioreactor. The biofilm load on the membrane surface evolved autonomously to 10.2 g m^{-2} within four weeks after start-up and also restored very quickly to a similar extent after washing-events of membrane structures between set-ups. Probably because of indirect aeration in the tank, the changes of aeration rates did not affect biofilm load on membranes. The biofilm load of this study is comparable with the load of 9.28 g m^{-2} that was detected on common biomass carriers used for grey water treatment (Jabornig & Favero 2013), but considerably higher than the 3.8 g m^{-2} , that were measured by Phattaranawik & Leiknes 2011. The reason is probably due to the air scouring of membranes, compared to the non-direct aeration used in the current study that kept the biofilm thickness in a constant range. The OLR ratio was calculated with a mean value of 0.44 kg COD kg^{-1} biofilm (dry mass). The average organic loading rate of the biofilm was measured with 4.21 g COD m^{-2} day^{-1}. Leiknes and Ødegaard (2007), measured similar loading rates of 3.4 and 3.9 g filtered COD m^{-2} day^{-1}, respectively, when they developed a moving bed bioreactor followed by membrane filtration. In a previous study a mean of 5.1 g COD m^{-2} day^{-1} was measured, while operating a combination of moving bed and membrane filtration for GW (Jabornig & Favero 2013). Hence, the results of this study were in accordance with earlier publications and may indicate that similar loading rates for the current process can be applied. Interestingly, the suspended solid concentration in the bioreactor remained on a similar low range during the whole pilot tests with a mean of 17 ± 23 mg L^{-1} only. Different aeration rates had minor influence. The reason may be due to the single sided, separated aeration device which led to good settlement of sludge in low turbulence areas of the bioreactor. The low ratio of suspended sludge in the bioreactor of only 4 – 5 % compared to the biofilm load supports the hypothesis that the biofilm on the membrane was mainly responsible for degradation. Sludge from the bottom of the bioreactor was removed only three times in 6 month with an overall amount of 13.1 L (week 10 - 5.5 L, week 14 - 3.2 L and week 22 - 4.4 L). The mean TSS concentration of the bottom sludge was 14.3 g L^{-1}. This relatively low surplus sludge amount confirms high stabilized sludge in the system, resulting from long sludge retention.

3.5 Fouling characteristics and Influence of Aeration Rate on Flux (Set-up A1 – A3)

Figure 4 - 4 underlines that flux stabilization could be obtained for each set-up within 15 days. In set-ups A1 repeat, A2 and A3 final flux values between 1 - 2 L m^{-2} h^{-1} were obtained. The final mean flux with highest aeration rate was slightly higher compared to lower aeration rates. Compared to initial flux of 35 m^{-2} h^{-1}, there was only a difference of < 3%. Thus, and in consideration of other possible impacts (e.g. temperature, measuring error) the final flux was concluded to be independent of aeration rates. The initial flux of 35 L m^{-2} h^{-1} right after start-up could not be observed again. The primary fall in flux from 35 to less than 10 L m^{-2} h^{-1} within 10 days is probably the result of irreversible blockings of pores by organic and inorganic particles < 0.2 µm which may only be restored by chemical cleaning. Microorganisms are generally larger than 0.2 µm and would not pass the membrane pores Out-of-tank washes could elevate flux to roughly 7 – 9 L m^{-2} h^{-1} which decreased again shortly after restart. Once stabilized flux was reached, the value kept constant and seemed to be the long-term bottom line for this operation scheme. Similar stabilization time, but higher flux was reported by Peter-Varbanets *et al.* 2010, who used water or diluted wastewater with comparable lower organic loading and without considering the membrane as biofilm carrier. **Figure 4 - 2** also shows that there are mainly three stages of flux decline. After start-up it was strongest with up to 4.5 L m^{-2} h^{-1} per day. The decline slowed down after 14 days and after 30 days finally flux stabilized at 2 L m^{-2} h^{-1}. Below 2 L m^{-2} h^{-1} the system seemed to be in a steady state and flux did not decrease any further. After first flux stabilization, the membrane was removed and gently washed with clear water to remove biomass and deposits from the surface. By this procedure, flux could only be restored to 7.9 L m^{-2} h^{-1} which was 77% less filtration capacity compared to the start. Out-of-tank washes removed surface deposits from the membrane but could not restore the high original start-up level As the biofilm layer on the membrane surface was not present at the beginning, non-degraded GW components could

have enter the membrane directly and maybe then enforced irreversible fouling. Later flux declines were always reversible with an out-of-tank wash. Thus, the biofilm layer could have additionally protected pores from further irreversible fouling.

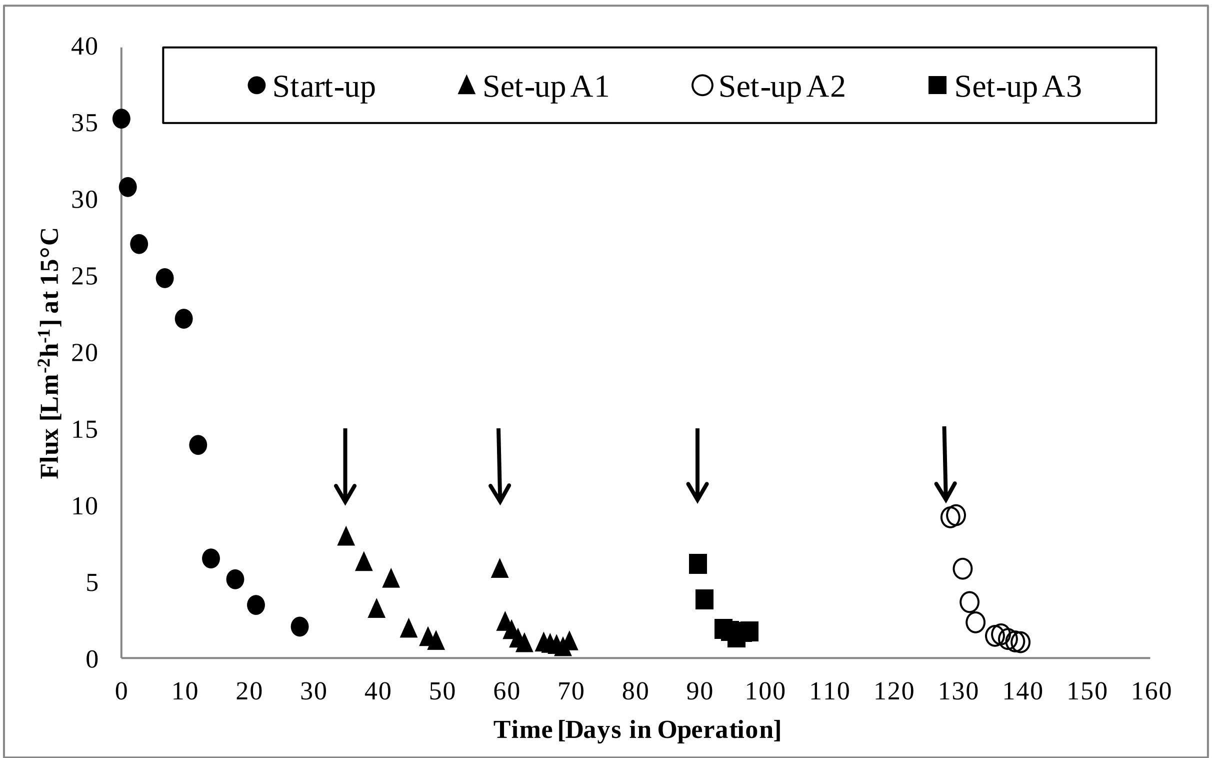

Figure 4 - 4: Flux stabilization after start-up (0.38 Lm-2min-1) and at different aeration rates (A1 0.19 Lm-2min-1, A2 0.38 Lm-2min-1, A3 0.67 Lm-2min-1), temperature corrected to 15°C, arrows indicate out-of-tank washes of membrane module.

3.6 Biofilm Analysis of experimental set-up A3

A membrane sample, wrapped in a biofilm layer from experimental phase A3, was analysed by stereo-, light- and confocal laser scanning-microscopy to investigate biofilm structure and microbial composition. **Figure 4 - 5a and 4 - 5b** show the typical structure of the biofilm layer, found on the membrane surface. The biomass and organic deposits formed a rather thick layer with 100 – 150 µm around the membrane fibre. The structure was spongy, woven like and rather weak. Gently mixing or washing removed the layer completely. Even after 10 weeks of operation the membrane itself seemed to be almost clean and still had a white appearance like a new membrane. Further increase of the magnitude of the light microscopy (**Figure 4 - 6a and 4 - 6b**) revealed that huge parts of the biofilm consisted of sheathed bacteria, like *Sphaerotilus natans* or *Leptothrix discophora*, a facultative aerobic species typically found in wastewater (Witzig *et al.* 2002). Additionally coccoid and spirillae-like microorganisms, protozoa and nematodes could be detected. Peter-Varbanets *et al* 2010, concluded that active biomass is responsible for flux stabilization during dead-end membrane filtration. Although biofilm analysis was performed only for set-up A3, similar characteristics of flux stabilization in all other configurations, indicate same functioning of the biofilm-layer throughout the experimental phases.

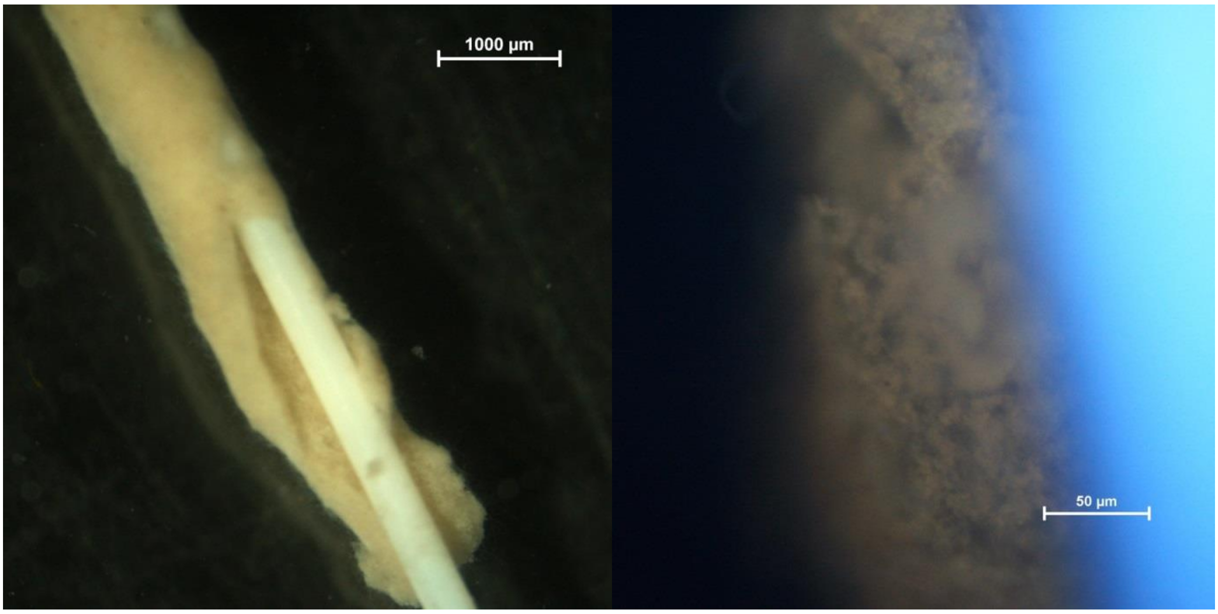

Figure 4 - 5: a) Stereo-microscopy of membrane fibre with biofilm-cover; scale bar represents 1000 µM; b) Light microscopy picture of biofilm cross section touching UV-fluorescent membrane; scale bar represents 50 µM;

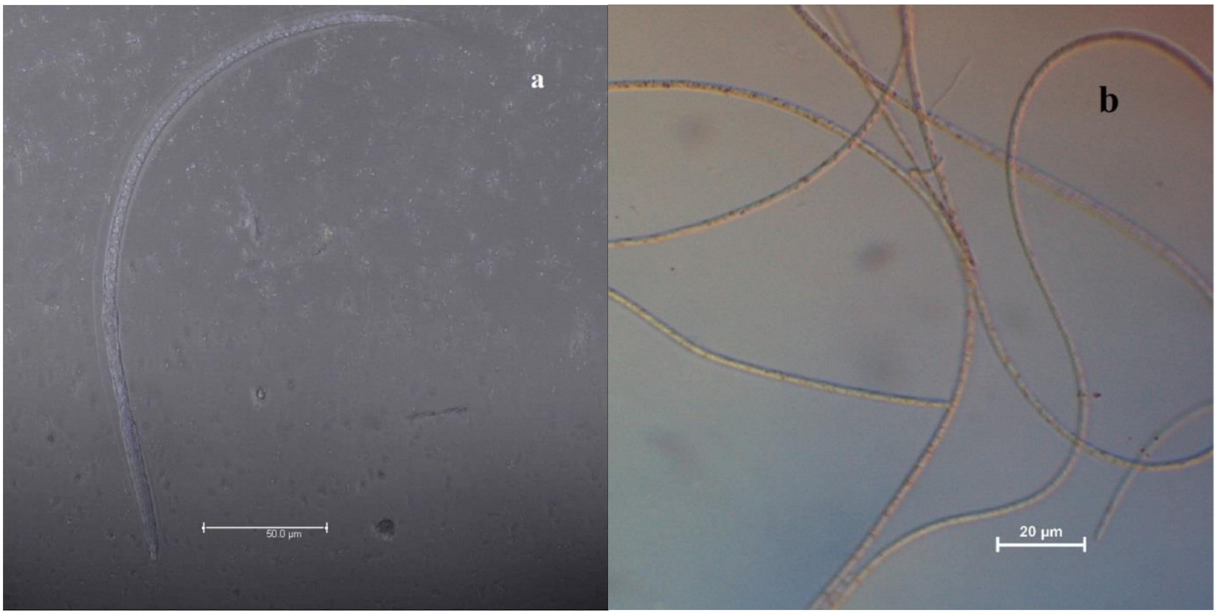

Figure 4 - 6: a) CLSM of nematode found in biofilm structure; 50 µM scale bar; b) Light-microscopy picture of sheathed bacteria found in biofilm; scale bar represents 20 µM;

Confocal laser scanning with DAPI-stained biofilm samples was conducted to investigate the porosity and ratio of microorganisms and organic deposits. Porosity of the submerged biofilm sample from experimental line A3 averaged around 54.9 ± 2.09 %, confirming the highly perforated and spongy structure. The large number of nematodes found in the biofilm sample by light microscope, might further explain this loose biofilm characteristics. Like earthworms in soil habitats, nematodes may act as key players for disaggregation of the biofilm layer and thus play an important role in flux rate stabilisation.

The estimated ratio of biofilm/biomass without nematodes obtained through image analysis ranged between 5.9 to 31.9 with an average of 20.2 ± 10.3. It has to be considered however, that DAPI staining may lead to underestimation of total biomass, as it is only staining DNA and not allowing for a clear determination of the borderline of each cell. Thus, the true ratio might be narrower.

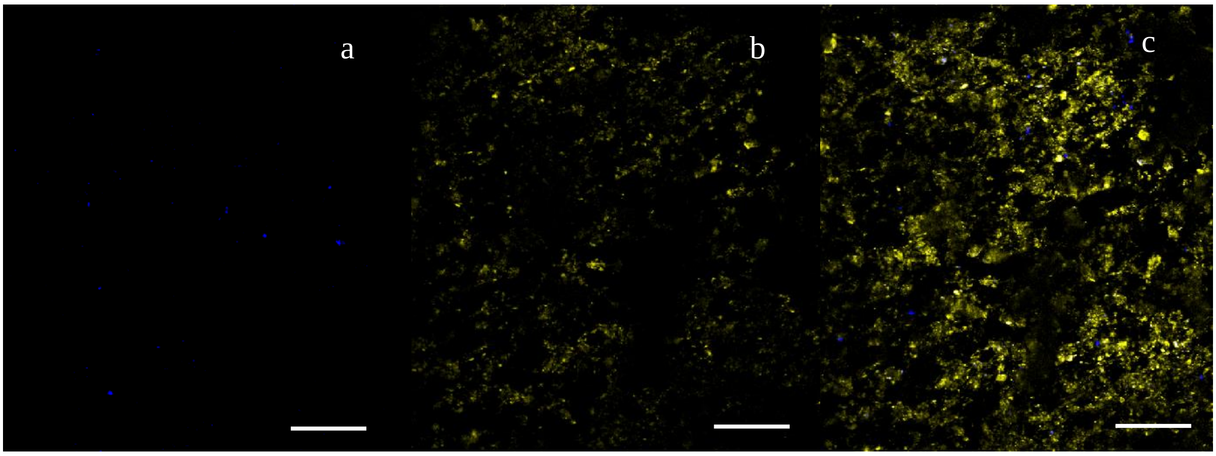

Figure 4 - 7: a) CLSM picture of DAPI-positive signals (biomass) (1 slice); b) CLSM picture of DAPI-negative signals (biofilm) (1 slice); c) CLSM picture of DAPI-positive and negative signals (biofilm and biomass); maximum projection of 21 slices; all scale bars represent 100 µm;

4 Conclusion

Here we introduced the application of a combined fixed-fibre biofilm membrane system for the treatment and reclamation of on-site GW. Following conclusions can be drawn from a 6 month test period:

- Treated water achieved on average international guidelines for unrestricted urban reuse.

- After one week of operation the flux decline was strongly reduced for all set-ups and seemed to stabilize without fouling control on a flux level of 1-2 L m^{-2} h^{-1}. Confirmation of long-term flux stabilization is pending.
- The power demand of the process was around 1.4 kWh m^{-3}. Further consumables were not necessary.
- Chemical cleaning was not required once flux remained constant after initial stabilization.
- Different aeration rates between 0.19 – 0.57 L m^{-2} min^{-1} had a clear influence on the COD treatment performance and showed only minor influence on the flux.
- Flux stabilization effect is due to strong biological activity in and around the biofilm.
- Biofilm analysis revealed numerous nematodes, sheathed bacteria and protozoa, which form spongy, woven-like and porous biofilm structures. These microorganisms are most probably responsible for the flux stabilization effect without fouling control.

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CHAPTER 5:

Modelling of Moving Bed Biofilm Membrane Reactors (MBBMR) for On-site Grey Water Treatment

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Modelling of Moving Bed Biofilm Membrane Reactors (MBBMR) for On-site Greywater Treatment

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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 grey water. 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 ten month lasting pilot test. The model shows good predictions of readily biodegradable COD (S_S) and ammonium removal (S_{NH}), as well as biomass concentration on carriers (X_{HB}) and in suspension (X_{HS}). Besides, a sensitivity analysis was made, which calculates the relative significance factor (RSF) 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 grey water reclamation.

Keywords: grey water, modelling, MBR, moving bed biofilm, MBBMR

1 Introduction

Wastewater recycling in private households is a promising way to overcome potable water shortages in arid regions. Reused grey water (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. Literature reports of numerous combinations of the same (**Li *et al.*, 2009**). However, treating and recycling grey water 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 micro- or ultrafiltration. While in the bioreactor long sludge ages can be applied, which remove particulate and dissolved nutrients to a high extent, the physical membrane barrier additionally withdraws all other suspended solids and microorganism. This significantly reduces plate count and the effluent quality resembles disinfection. Several publications document that the process combination meets the guidelines and fulfils the expectations of users (**Jabornig & Favero, 2013, Kraume *et al.*, 2010, Merz *et al.*, 2007**). However, the contradiction of increasing acceptance by enhancing treated water quality on the one hand, and on the other hand reducing investment and operating costs for a reasonable payback period 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. (**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**) and MBR applications for mixed sewage (**Fenu *et al.*, 2010**) and grey water (**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 (MBBMR). 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 the main parameters. Data from previous publications (**Jabornig & Favero, 2013)** and new data sets 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 paid to typical situations and environment of single household applications and by this showing their impact on process stability and treated water quality.

2 Material and Methods

2.1 Grey Water Characteristics

The pilot plant was operated with 180 L day^{-1} synthetic grey water freshly mixed of different shower gels, shampoos, soaps, deodorants and other typical grey water ingredients according to **NSF/ANSI 350 (2011)** with COD:N:P 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 roughly represents a four people 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 grey water was assumed to be 35% - 60 % - 5%, typical for grey water (**Hocaoglu M. *et al.*, 2013)**. Feeding of grey water was done in batches in the morning (40%), noon (20%) and evening (40%) in order to simulate actual inflow conditions.

2.2 Pilot Plant

The pilot plant was operated for a period of 10 months and combines a moving bed biofilm process with a microfiltration membrane (hollow fibre, HDPE, C-MEM) with a nominal pore size of 0.2 µm in one bioreactor. It was operated in batches dependent on the actual inflow of grey water 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 grey water. Extraction of treated water (permeate) was done in two batches, each lasting maximum three hours, so in total six hours per day. Equipment selection and design of the bioreactor was made according to **Jabornig & Favero (2013),** which included PLC-controlled permeate and back-flushing pump (40 W, < 4 L/min), non-continuous time controlled co-current aeration device for membrane and bioreactor (29 W, 30 L/min) and cylindrical biomass carriers (HDPE, 0.95 kg m^{-3}). 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°C – 30 °C.

2.3 Model characteristics

2.3.1 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 am, 25 L from noon to 1 pm and 80 L from 7 to 8 pm. Raw water is entering the system as short flushes with 10 L min^{-1}. Effluent by permeate extraction was 90 L from 4 to 7 pm and additional 90 L from 4 to 7 am 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 one to three weeks after start-up of membrane filtration even without or strongly reduced fouling control (**Jabornig & Favero, 2013; Peter-Varbanets *et. al.*, 2010**) and was therefore used as output during permeating phase of the plant. Due 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:

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

$$\frac{dV_B}{dt} = Q_{in} - Q_{out} \qquad 2$$

This concept came along with the assumption that change of concentration of material A in batch reactor was not only dependent on degradation by a certain process rate r_A but also on water level.

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

Temperature influence on the process rates in the batch reactor (e.g. for high inflow and low ambient temperatures) were also implemented into 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.

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

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

2.3.2 Bioreactor

The basic elements of the mechanistic model used in this study in **Table 5 - 1** were adapted from **Hocaoglu M. *et al.*, 2013**, who applied their ASM1 based model on grey water treatment with conventional activated sludge membrane bioreactors (AS-MBR). The model includes the following constituents in the bioreactor: (1) degradation of readily biodegradable chemical oxygen demand (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). Six 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. The process additionally included two switching functions for ammonium and phosphate due to typical limitations of these parameters in grey water. Ambient conditions of the pilot plant were considered in the process by an experimentally derived temperature dependent growth rate.

$$\mu_H = k_{T_0} e^{(k_T(T - T_0))} \quad 6$$

(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 hydrolysis step of S_{RB} was included because of the retention and accumulation of additional particulate matter through the membrane barrier (**Hocaoglu M. et al., 2013**). (4) Microbial decay by endogenous respiration was included as a linear function generally used in ASM models. (5) The process of irreversible sludge settlement was simulated by a linear function. Settlement was assumed to be dependent of the settling velocity of free floating biomass in the suspension. The settling velocity was again dependent on the aeration characteristics and was experimentally derived. Important in this context is that total heterotrophic biomass X_H was

assumed as a sum of heterotrophic biomass in biofilm X_B and suspension X_{HS}. The biomass in the biofilm was further dependent on the aeration characteristics of the bioreactor.

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

(6) 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.

The overall process was simulated in MS EXCEL software on the basis of numerically solved differential equations with an output time step of 1 hour for long term development and 1 min for short term degradation of S_S and oxygen uptake rate. All coefficients and parameters were iteratively calibrated based on experimentally derived steady state data from the pilot tests. 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):

$$ARD = \frac{1}{N} \times \sum_{i=0}^{N} \frac{|(m_i - p_i)|}{m_i} \times 100\% \tag{8}$$

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}, X_{HS}.

Table 5 - 1: Stoichiometric matrix of grey water MBR bioreactor model (adapted from Hocaoglu M. et al., 2013)

	Material i	1 Soluble Substrates S_S	2 Particulate Substrates X_S	3 Soluble microbial products S_P	4 Particulate microbial products X_P	5 Resulting residual soluble COD in the MBR S_{RB}	6 Ammonium S_{NH}	7 Phosphate S_{PO}	8 Oxygen S_{O2}	9 Biomass / Biofilm X_H	Process rate ρ_j
j	Process	g_{COD} m^{-3}	g_{COD} m^{-3}	g_{COD} m^{-3}	g_{COD} m^{-3}	g_{COD} m^{-3}	g_N m^{-3}	g_P m^{-3}	g_{O2} m^{-3}	g_{COD} m^{-3}	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								$\rho_2 = k_h \frac{\frac{X_S}{X_H}}{K_X+\frac{X_S}{X_H}}X_H$
3	Hydrolysis of S_{RB}	1				-1					$\rho_3 = k_{hr} \frac{\frac{S_{RB}}{X_H}}{K_X+\frac{S_{RB}}{X_H}}X_H$
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	$\rho_4 = b_H X_H$
5	Settlement of sludge in bioreactor									-1	$\rho_5 = \frac{w_S\, A}{V_B}X_{HS}$
6	Aeration of bioreactor								1		$\rho_6 = \frac{Y_{O2}\, Q_A R_A}{V_B}$

2.3.3 Sensitivity analysis

A sensitivity analysis was made with the relative sensitivity factor (RSF) function according to **Jiang et al. (2005)**. The 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.

$$RSF = \frac{\theta}{Y}\frac{dY}{d\theta} \qquad 9$$

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

$$Y'(\theta) = \frac{Y(\theta+h)-Y(\theta-h)}{2h} \qquad 10$$

A simple approach was preferred to recently reported global methods (Cosenza *et al.*, 2014, Sin *et al.*, 2011). 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).

3 Results and Discussion

3.1 Hydraulic characteristics

The pilot plant was fed with daily 180 L fresh grey water in flushes in the morning, noon and evening, simulated as short peaks. Treated water was extracted batch wise through the membrane six hours 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 5 - 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 is situated in the same tank resulted in the need to separate inflow and effluent temporally as best as possible. Any other configuration showed to have higher bypass flows of non-treated grey water 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 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 app. 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 microorganism 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.

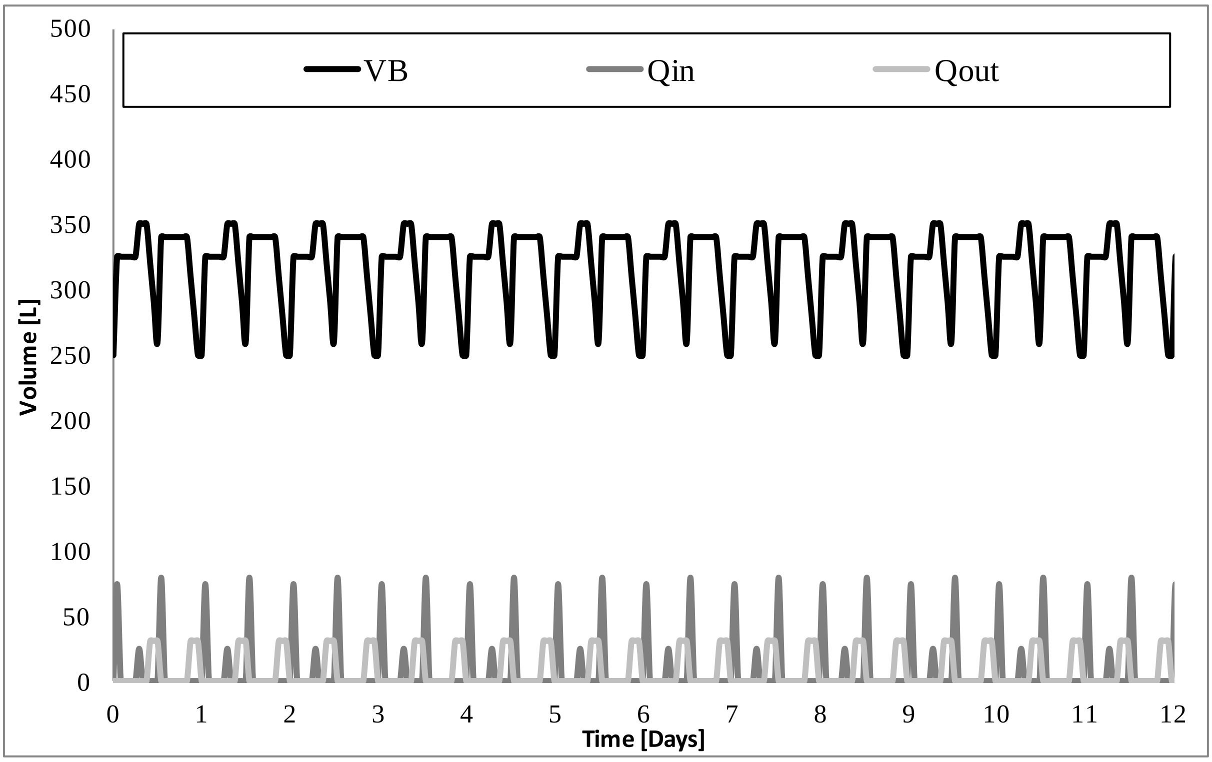

Figure 5 - 1: Modelled hydraulic characteristics of bioreactor volume (V_B), inflow (Q_{in}) and effluent (Q_{out})

3.2 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 being 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. In order to fulfil the observed 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 due to 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 five weeks after start-up of the pilot plant when the moving bed carriers were finally filled up with biofilm. Steady state of bioreactor could be maintained till the end of the pilot tests. **Figure 5 - 2** shows the modelled and measured concentration of biofilm biomass carriers and suspension. The actual concentration of biomass in the model is related to the filling level and therefore subject to variation when the water level changes due to feeding and permeate extraction. The trend of X_{HS} compared to X_{HB} over the day was more even because of settling and continuous removal of regrown biomass. The settled sludge on the bottom of the tank was siphoned off in an amount of 8 L $month^{-1}$ and was measured in a concentration of 14 - 16 g L^{-1}. The bottom sludge degradation characteristics itself 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.

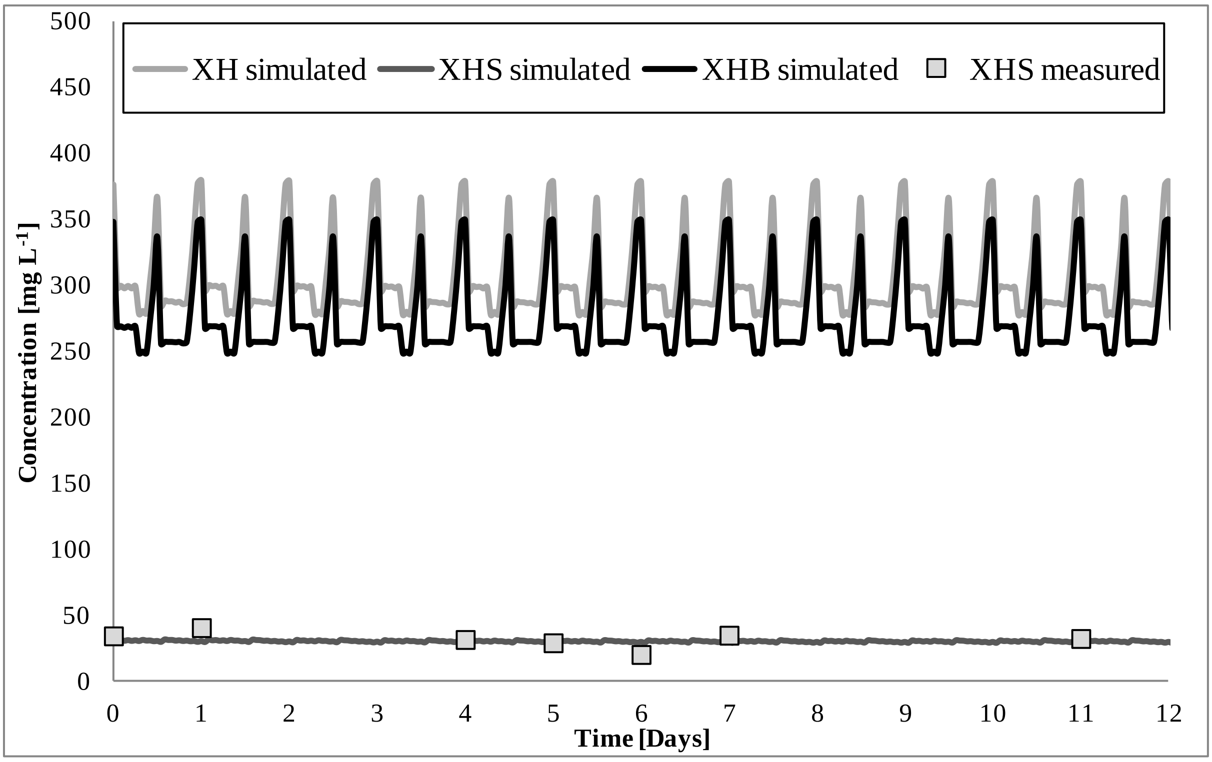

Figure 5 - 2: Simulated and measured concentration of biomass in bioreactor at 9.1% aeration rate

Readily biodegradable COD (S_S) was measured in the treated water tank during the stabilized phase of the pilot tests in range of 40 – 60 mg L^{-1}. The simulation of this parameter shows the typical variation over the day resulting from changing reactor level. Nonetheless, the variation appeared only in the simulation because actual measurement combined the treated water from six hours of filtration and therefore high peaks were usually omitted. The readily biodegradable COD value was comparably higher than in other grey water MBR studies (**Merz *et al.*, 2007**, **Kraume *et al.*, 2010**, **Hocaoglu M. *et al.*, 2013)** indicating slow degradation and also verifies the fitted low growth rate of biomass of the model. Oxygen as restriction could be excluded because it was available in excess even in lowest aeration rate. The limitation of growth rate and shortened S_S removal could be further enforced by a low diffusion rate into the biofilm and by a low concentration of nutrients of ammonium and phosphate. This assumption follows the low ammonium concentrations which were measured in the treated water tank 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. However, peaks up to 0.25 mg L^{-1} right after dosing of grey water 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 in a value of 0.69 day^{-1} showed to be lower than in other black- and grey water studies. The slow growth rate of biomass was also consistent with the low oxygen consumption of the process. **Figure 5 - 3** shows that right after feeding of fresh grey water the model predicted a sharp oxygen decay which was enforced by degradation, dilution and temperature increase. After this fall, oxygen was consumed by the microorganism slowly, almost linearly. Similar oxygen uptake characteristics could be confirmed by the O_2-measurements in the bioreactor of the pilot plant, respectively.

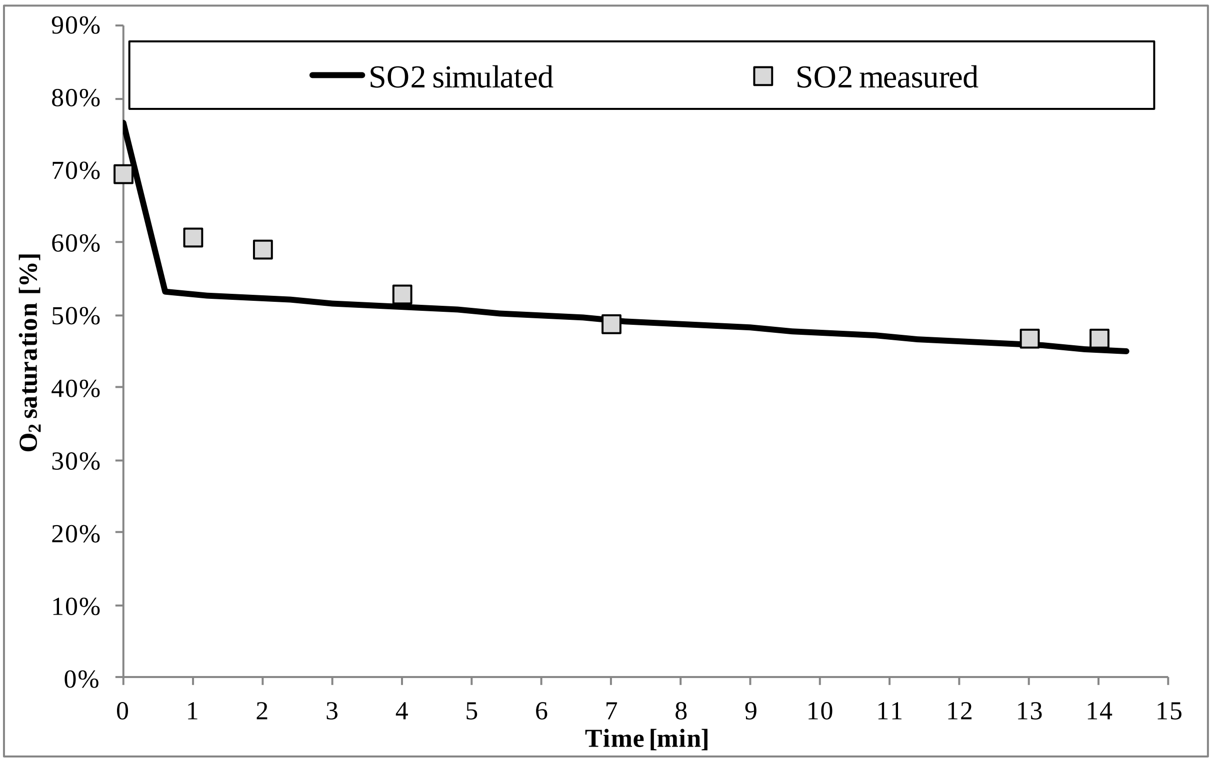

Figure 5 - 3: Oxygen saturation decay in the bioreactor with switched-off aeration blower shortly after feeding fresh grey water. The straight line shows a comparison between modelled and measured O_2-saturation.

3.3 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}. All other parameters were at least moderately influential in one of them. Results also show the strong influence of growth rate and yield factor on the treatment performance. 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 paid 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 proofed to be further dependent on the sedimentation velocity (w_S).

4 Conclusion

The treatment combination consisting of moving bed biofilm process and membrane filtration (MBBMR) in one bioreactor could be successfully modelled and calibrated with obtained pilot plant results. 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.
- 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 made 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 greywater kept balanced even at freezing level.

5 Acknowledgments

We would like to thank the company SFC Group and RHV Siggerwiesen for providing the pilot testing facilities.

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}	Phosphorous 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 phosphorous per mass of COD in biomass [gN g^{-1} COD]
i_{XN}	Nitrogen fraction of particulate microbial products [gN g^{-1} COD]
i_{XP}	Phosphorous fraction of particulate microbial products [gP g^{-1} COD]
k_a	Aeration rate of bioreactor [-]
k_h	Maximum hydrolysis rate of XS [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 phosphorous
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}]

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_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_{O2} O_2 transfer coefficient diffuser [mg O2 L^{-1} air]

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CONCLUSIONS AND OUTLOOK

2. CONCLUSIONS AND OUTLOOK

2.1. CONCLUSIONS ARISING FROM THIS THESIS

This present thesis has extended on the one hand the understanding of the current grey water treatment market, and on the other hand provides alternative treatment concepts in order to reduce energy demand and maintenance efforts in the future.

2.1.1. Current Market Situation of GW Treatment

A market study including a payback analysis conducted in 2013 has shown that there are numerous low-tech and advanced GW treatment systems available on the market. Unfortunately, wide dissemination is still pending. One main reason has been identified to be either low effluent quality or high investment and operating costs. Literature results have shown that MBR technologies have proven to achieve highest reuse water guidelines (e.g. EU bathing water quality) but investment and operating costs have still been too high for a feasible use in single households with a capacity of < 500 L day^{-1}. Nonetheless, the payback analysis has also indicated that these small systems would become cost-effective if energy demand could be lowered from average 3.3 kWh m^{-3} to less than 1.5 kWh m^{-3}. (*Chapter 1*)

2.1.2. Investment Price vs. Operating Costs

The market study has also shown that membrane costs in micro-MBR plants for less than 500 L day^{-1} were only an underpart of less than 5 – 10 % of total investment of roughly 5,000 EUR net. Related to long-term operation of 10 years, the increase of membrane area would pay off quickly if energy demand and maintenance could be lowered by 50%. (*Chapter 1*)

2.1.3. Feasibility of Flux Stabilization in On-Site MBR plants

Membrane air scouring as a part of fouling control is responsible for main energy consumption in immersed membrane filtration in MBR treatment. Strongly reduced membrane air scouring (less than 10% of filtration time) with or without additional back-flushing strongly decreased the output of a conventional on-site MBR plant. Yet, flux has not dropped continuously, but has stabilized after 7 – 10 days on a low but constant level in a range of 2 – 4 L m^{-2} h^{-1}. Energy demand could be lowered by this low-flux operating philosophy to less than 0.3 kWh m^{-3}. (*Chapter 2*)

2.1.4. Feasibility of Moving Bed Biofilm Membrane Reactors (MBBMR)

Grey water treatment with a new approach combining a moving bed biofilm reactor with membrane filtration in one reactor (MBBMR) has achieved average values of international reuse guidelines. In a ten-month pilot test the process has proven to be robust and low in maintenance. Even effluent quality did not change remarkably by different loads and ambient temperatures resulting from vacation and washing days. The energy consumption of the overall process could be lowered mainly by reduction of the fouling control of the membrane filtration system to less than 1.4 kWh m^{-3}. (*Chapter 3*)

2.1.5. Start-up Procedure of Bioreactors in MBBMR Process

The start-up of a grey water bioreactor without additional seed microorganisms had neither remarkable influence on the process run-in time nor on flux development. (*Chapter 3*)

2.1.6. Feasibility of Fixed Fibre Biofilm Membrane Process

A combined fixed-fibre biofilm membrane system for the treatment and reclamation of on-site GW requiring no fouling control has also proven to achieve average values of international guidelines for unrestricted urban reuse in a six month lasting campaign. Flux has stabilized without fouling control on a flux level of 1 – 2 L m^{-2} h^{-1}. The confirmation of long-term flux stabilization is pending. Chemical cleaning was not required once flux remained constant after initial stabilization. (*Chapter 4*)

2.1.7. Principle of Flux Stabilization Without Fouling Control

After 7 - 10 days in operation of a fixed-fibre biofilm membrane system without fouling control a biofilm layer on the membrane has built up. The biofilm analysis has revealed numerous nematodes, sheathed bacteria and protozoa, which had formed spongy, woven-like and porous biofilm structures. These microorganisms are most probably responsible for the flux stabilization effect without fouling control. By their strong activity, the microorganisms kept the sludge layer permeable, which can be compared to earthworms in soil habitats. Similar to classical biofilm processes the biofilm on the membrane was also mainly responsible for the degradation of soluble pollutants. Additional activated sludge or biofilm growth bodies were not necessary during the pilot tests. (*Chapter 4*)

2.1.8. Modelling of MBBMR Process

A modelling study of the MBBMR process showed a good prediction of highly variable bioreactor water level due to unbalanced grey water production mainly in the morning and evening. Therefore, special attention needs to be paid to filtration time in order to avoid that untreated grey water passes the membrane. The calibration of the bioreactor model also confirmed literature data that there is a lower growth rate of microorganisms in grey water treatment compared to sewage applications. A reason could be the limiting nutrient ratio for grey water in terms of phosphorous and nitrogen content. Consequently, surplus sludge production is low and settles irreversibly on the bioreactor bottom. Removal might not be necessary for a long time or even not at all. The model has also confirmed experience from pilot testing that ambient temperature and short-term load fluctuation does not affect the stability of the proposed MBBMR process. (*Chapter 5*)

2.2. GENERAL OUTLOOK ON GW RECYCLING

Historically, GW recycling is nothing new. Decentralised water recycling has been practised for many centuries to conserve fresh water. With the development of supply-driven centralized water distribution, suddenly fresh water was available in sufficient quantities and quality. Currently, water recycling is seen instead in a way more elaborate, dirty and time-consuming and has almost disappeared in industrialised countries.

The design of buildings, regardless of single household or multi-storey apartments, reflects this way of thinking and represents one major obstacle of wide-spread GW reuse: not existing separate pipe installation for different waste water parts, fresh and recycled water. In existing buildings the implementation of GW treatment, storage and recycling system would not be feasible because of extensive structural alteration works. For new or renovated buildings it would be easier but still comes along with extra costs for an additional place of installation and piping system. Rainwater harvesting, which has become quite popular recently, could be also a transition step towards water recycling. Interestingly, rain water harvesting is even less cost-effective than GW recycling but is in the public perception generally more accepted. Savings of rainwater usage occur only through reducing fresh water costs but do not change overall waste water amount and fees. Nevertheless, both technologies complement each other during supply shortages and could feed the same reuse piping system. GW treatment could be offered as add-on kit in order to increase productivity and savings.

Another possibility to shorten pay-back time would be subsidies similar to the public support program for renewable energy. Despite water recycling is undisputed ecologically reasonable, there is a fundamental difference. Wide-spread water recycling would further reduce earnings of already underfinanced municipalities who are usually owner of the centralized water supply and waste water treatment plants. Secondary, there are also concerns about reduced capacities of sewer systems in regard to the transport of solids, increased salinity and reduced biodegradability of waste water caused by missing low polluted GW from showers, baths and hand basins. These problems can be solved but in the end it will further increase costs of sewage transport and treatment. Less earning of municipalities and resulting higher water fees would cause again better pay-back times of decentralized GW recycling plants. In the next round there will be again more plants, even less income for central treatment and so on and on. In respect to above considerations and the fact that central supply and treatment never can be omitted, it would be more reasonable if there will be an extra charge on GW treatment instead of subsidies.

High water price regions, like the Northern European countries where people already have a strong water saving mentality, could be also a double-edged sword for water recycling. On the one hand, high fresh and waste water costs would be favourable for water recycling but on the other hand they also reduce fresh water demand. Lower production of recyclable water negatively affects payback time of a treatment unit. In other regions the supply of sufficient fresh water is a political issue and is highly subsidized by the state. Lower costs will lead to a higher consumption but also prevent people from reusing water. Only more and more water shortages and exhausted non-renewable water resources will sooner or later change this mentality and could open the doors for more water recycling.

The current market of GW recycling, especially for high-end treatment which was the topic of this thesis, is limited and strongly dependent on financial feasibility. It would be already cost-effective for bigger communities in newly built or renovated hotels, schools or multi-storey building. In these applications, the awareness of people, architects and designer in view of the already existing, long-term approved technologies of water recycling need to be generally increased. Concerns about health issues or possible cross connections of fresh and reuse water pipes have to be cleared out. If people are used to water recycling in public, they may tend to accept it also at home. Anyhow, the same issue about shorter pay-back times is also effective for the small on-site treatment plants. If the positive financial feasibility cannot be guaranteed, they will probably remain only niche products for ecology-minded people.

2.3. ALTERNATIVE APPLICATIONS OF PRESENTED PROCESSES

GW recycling with presented processes has been successfully tested for over two years. In regard to the currently rather limited demand of GW recycling, other applications have been figured out to expand the short-term range of the low pressure membrane filtration technology.

In cooperation with company Intewa (Germany), which is a specialist in rain water harvesting and house installation equipment, the product Aqualoop was created. Aqualoop has been developed on the basis of the moving bed biofilm membrane bioreactor (MBBMR, *Chapter 3*) and was transformed into a more or less universal product of in-house water treatment. It is a modular concept which basically includes pre-filtering of coarse particles, moving bed biofilm process, membrane filtration, heat reclamation and clear water storage (**Figure 2**). Depending on raw water characteristics, the one or other treatment step can be omitted while membrane filtration is always the key element.

Apart from GW treatment, following applications could be feasible and are currently in pilot testing status:

1. <u>Surface water treatment</u>

   Surface water of nearby springs, rivers or lakes can be used for toilet flushing or gardening. High turbidity and microbiological contamination of surface water will be removed by direct membrane filtration so that reuse guidelines will be met. Basically, the acceptance of using surface water for in-house applications is higher than for treated GW and it would not require separate pipe installation for GW collection. Thus, there are less investment and operating costs. The disadvantage is a limited range of users who have surface water source available and that overall sewage fees will not be reduced.

2. Potable water from rainwater harvesting

   As mentioned before, rainwater harvesting has become popular nowadays. The reason is not economically feasibility but eco-minded people who do it without considering payback times. The reuse possibilities with current technology are limited. Harvested and collected rainwater will be pre-filtered and used for toilet flushing and gardening. In an on-going pilot study Intewa has shown that rainwater can be also treated up to potable water quality by direct membrane filtration. Then it would be also a source for showering, bathing or washing machine. Positive side effect is less scaling of pipes and armatures and better cleaning effect of washing machine because of the lower TDS compared to ground water. People also report from more pleasant feeling after showering because of the soft water. The mains connection from city water will only be used in the kitchen.

3. Tertiary treatment of on-site sewage treatment plants

   In case a conventional on-site activated sludge treatment plant is already in operation, there is the possibility to treat the effluent once again with direct membrane filtration and recycle the water for toilet flushing. The effluent of small waste water treatment plants usually will be drained to a recipient or seeped away. The reuse of the water would directly safe fresh water fees without increasing the waste water treatment costs. A separate waste water collection compared to GW would not be necessary. An additional reuse pipe installation is still mandatory.

4. Heat reclamation from GW

   Remaining heat in GW from showering or bathing will be transferred from treated water by a heat exchanger and used to pre-heat mains water arriving at the boilers. The innovation about this application would be the uncomplicated heat transfer from treated water directly to the mains water. In other applications costly pre-treatment is necessary to protect the heat exchanger from clogging. Biologically treated and membrane filtered water do not contain any suspended solids. Danger of blocking of heat exchanger in the proposed process is almost excluded. Maintenance and operating costs of heat transfer are reduced to its minimum.

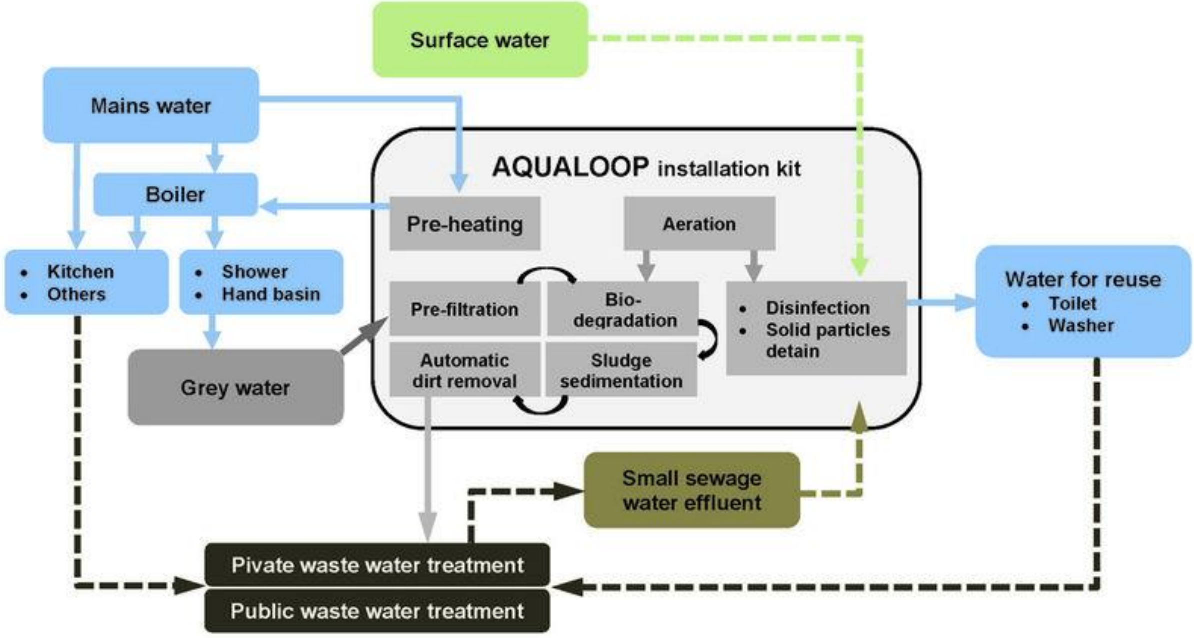

Figure 2: Flow scheme of Aqualoop treatment system from company Intewa

The mentioned applications are not fundamentally new and common membrane filtration itself has also shown to be technically feasible and reliable for over 30 years. Nevertheless, investment and operating costs of low pressure household membrane filtration plants could be significantly reduced with the new approaches discussed in this thesis. In-house treatment of water may become more economically viable and as a result more successful on the market in future.

2.4. OUTLOOK FOR FURTHER RESEARCH

Further work and necessary future research projects to address performance and costing of proposed GW treatment processes are as follows:

1. Two processes have been proposed for solving the problem of high operating costs of conventional grey water treatment with MBR: (i) a moving bed biofilm reactor with membrane filtration in one reactor (MBBMR) and (ii) fixed-fibre biofilm membrane system. A detailed comparison of both and additional long-term performance testing

over several years is pending. Especially the principles of flux stabilization need to be confirmed and investigated in more detail.

2. A cost analysis of currently available systems has been made. Based on the data of pilot tests of newly proposed processes, a payback comparison study with currently available MBR systems for on-site treatment is pending.
3. Preliminary studies for extending the range of application for proposed processes are considered. Possible applications would be on-site sewage treatment, drinking water treatment and rainwater harvesting.

CURRICULUM VITAE

1. **Family name:** **Jabornig**

2. **First names:** Simon Raffael

3. **Date of birth:** 15.08.1983

4. **Nationality:** Austrian

5. **Education:**

Institution	University of Innsbruck Austria
Date: from (month/year): to (month/year):	10/2011 12/2014
Degree(s) or Diploma(s) obtained:	Dr. tech.

Institution	University of Applied Sciences Upper Austria
Date: from (month/year): to (month/year):	10/2003 07/2007
Degree(s) or Diploma(s) obtained:	Dipl.-Ing. (FH)

6. **Language skills:** (from 1 to 5, depending on level of knowledge; use 5 as fluent)

Language	**Reading**	**Speaking**	**Writing**
English	5	5	5
German	5	5	5

7. **Membership of professional bodies:**

None.

8. **Other skills:**

Experience as Project Engineer:	Several years experience in water related projects for private consultants, responsible for process design, tender preparation, evaluation of tenders, supervision of works, start-up and commissioning. - sewage treatment (collection, treatment) - Extensive experience with M & E & ICA equipment and O & M of wastewater treatment plants

9. **Present position:** Project Manager

10. **Professional Experience Record:**

Date: from (month/year) to (month/year)	1.8.2013 - to date
Location	Salzburg, Austria
Company	Environmental Service of Federal Government Salzburg
Position	Environmental Consultant
Description	Focus on Water Treatment and Reuse

Date: from (month/year) to (month/year)	July 2006 - to date
Location	Salzburg, Austria
Company	SFC Umwelttechnik GmbH
Position	Project Manager
Description	Process Design, R&D

11. **Others:**

11a. **Research Projects:**

- Membrane Development and Optimization

FFG Research Project 1.11.2010 – 31.10.2012

- Optimization of the C-MEM Membrane Filtration Technology regarding Flux, Life Time and Operation Costs

  FFG Research Project 1.10.2009 – 30.4.2012

- Intelligent Production, Increase of Efficiency in Membrane Production

  FFG Research Project 1.1.2013 – to date

- Eco-Innovation, Development and Serial Production of Grey Water Plants

  CIP-Eco EU Research Project 1.8.2013 – to date

14b. **Product Development:**

- Grey Water Use C-MEM (Cooperation with Intewa Co.)

  1.11.2011 – to date

- Swimming Pool Water Treatment with C-MEM (Cooperation with Eder Bau Co.)

  1.7.2012 – to date

- Koi Pond Water Treatment with C-MEM (Cooperation with Teichlogistik Co.)

- 1.8.2012 – to date

14c. **Publications, Lectures, Seminars, Conferences:**

- Jabornig S. (2014). Overview and feasibility of advanced grey water treatment systems for single households. Urban Water J. 11(5) 361-369

- Jabornig S. and Favero E. (2013). Single household grey water treatment with a moving bed biofilm membrane reactor (MBBMR). Journal of Membrane Science 446 277–285.

- Jabornig S. (2013). Flux Stabilization in On-site MBR Plants with Reduced Fouling Control. Conference Proceedings MBR Asia 2013.

14d. **Selected References:**

Project:	**Stary Las Landfill WWTP**
Client:	Municipality
Country:	**Poland**
Period:	2014 – to date

Services:	Process Specialist

Project:	**Audi Service Car Wash WWTP**
Client:	Beijing Xing Yue Yang Guang EPE
Country:	**China**
Period:	2014
Services:	Process Specialist

Project:	**Holzwärme Flachau WWTP**
Client:	Holzwärme Flachau GmbH
Country:	**Austria**
Period:	2013 - 2014
Services:	Process Specialist

Project:	**Al Ain WTP**
Client:	EET
Country:	**UAE**
Period:	2011
Services:	Process Specialist

Project:	**PETP MBR**
Client:	AMEC Mining / W2R
Country:	**United Kingdom**
Period:	2009 - 2010
Services:	Project Manager

Project:	**Tidel Park WWTP**
Client:	Tidel Park
Country:	**India**

Period:	2008
Services:	Project Engineer

Project:	**Sarajevo Landfill WWTP**
Client:	Department of Reconstruction
Country:	**Bosnia-Herzegovina**
Period:	2007 - 2008
Services:	Project Engineer

Project:	**Bosnalijek WWTP**
Client:	Bosnalijek d.o.o.
Country:	**Bosnia-Herzegovina**
Period:	2007 - 2008
Services:	Project Engineer

Project:	**Moura River WTP**
Client:	Queensland Nitrates / Cleanteq
Country:	**Australia**
Period:	2007 - 2008
Services:	Project Engineer

Project:	**Yeganeh Kazar WWTP**
Client:	Yeganeh Kazar
Country:	**Iran**
Period:	2007
Services:	Project Engineer