Optimal Operating Strategies of a Submerged Membrane Bioreactor for Wastewater Treatment

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Submerged membrane bioreactors (SMBR) are hybrid systems that couple a biological reactor, where the biodegradable substrates are consumed by the active biomass, and a membrane module, which prevents the solids to pass into the effluent, replacing the classical separation and recirculation unit.

An improved version of an available mathematical model describing a SMBR was used to search for the optimum operating strategies of a semi-continuous system for wastewater treatment. The model takes into account two simultaneous processes: biological chemical oxygen demand (COD) removal and the cake layer formation/removal onto the membrane surface.

Based upon this model, an optimization was done, searching for the operating parameters which maximize the final global conversion of ammonia plus ammonium nitrogen. These optimal values were then used to study the behavior of the system for two operating strategies: discontinuous and continuous permeate recycling. The results showed that the former strategy is the most effective.

Key words:
ASM1, wastewater treatment, submerged membrane bioreactor, fouling, membrane bioreactor modeling, optimal operating policy

Introduction

Fresh water is being called the "blue gold" of the 21st century. It is a natural resource already in short supply and will become even scarcer with increased urbanization and population, climate change, and industrial pollution, turning it into humanity's most precious resource and one of the major environmental issues of this century.

In this context, there are two important ways of coping with water scarcity: the first is to increase the water usage efficiency through either the reduction of fresh water consumption, or the increase of water reuse through a better optimization of water networks; the second is to increase constantly the importance of the wastewater treatment, finding ways to cut energy consumption and to foster system's efficiency.

One of the most widely used waste removal methods is biological treatment. The two main processes utilized for secondary (biological) treatment are the trickling filtration – lesser and lesser important – and elimination with activated sludge.[1]

All activated sludge systems include a suspended growth reactor, in which the wastewater, recycled sludge, and molecular oxygen transported from air bubbles are mixed, a separator and pump station for sludge recycle and a wasting line. The latter is intended to limit the accumulation of dead cells and, at the same time, to keep the activated sludge concentration into some predefined limits. The oxygen, coming from air – the usage of pure oxygen is discouraged due to the high operating and safety costs – must be transferred into wastewater, hence the need for an energy-intensive air supply system, which is also responsible for the energy dissipation at molecular level through mixing. The separator is usually a sedimentation tank that is designed to have a dual function as both clarifier and thickener.[2]

The microorganisms found in activated sludge could be lumped into four generic categories: bacteria, fungi, protozoa and rotifers.[3] The growth and predominance of each of these types are controlled by a number of feed and operating conditions including type of waste-organic matter (food), metabolic rate, size, nature of the process. In activated sludge, bacteria are the most frequent microorganisms. They exist as both individuals and colonies. Some are strictly aerobic while others are anaerobic. This ability to perform in the presence and absence of oxygen is an important asset which helps maintaining an acceptable sludge activity at low concentrations of oxygen.[4]

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The bacteria are classified in heterotrophs and autotrophs and the former are predominant in activated sludge processes. The autotrophic bacteria reduce the oxidized compounds with carbon (from i.e. carbon dioxide) for the cellular growth, on behalf of nitrification, which is a two-step process of oxidizing the ammonia to nitrate.

The heterotrophs oxidize the carbonaceous organic matter to gain energy and are responsible for the denitrification process. Denitrifying bacteria represent approximately 80 % of all bacteria flocculated and dispersed in the activated sludge process.[4]

Submerged membrane bioreactors

The SMBRs are hybrid systems, which group in the same unit the reaction and the separation steps,[5,6] according to the process intensification principles. The gains are twofold: increase of biomass concentration inside the bioreactor and the mass transfer area, accordingly, to obtain a higher flux of substrate to be processed and raise of the mass transfer driving force, keeping the product concentration as low as possible, removing it from the vicinity of the activated sludge.

The main advantages of the SMBRs, besides the higher effluent quality, are the smaller overall reactor volume, compared to the aeration and sedimentation basins used in the conventional activated sludge plants, and the reduced footprint and sludge production through maintaining a high biomass concentration in the bioreactor.[7–10] The long sludge retention time (SRT) achieved in the SMBRs, which can be manipulated independently from hydraulic retention time (HRT) due to membrane separation, also allows the retention of the microorganisms with low growth rate, such as the nitrifying bacteria, which is important for the efficiency of the nitrification process.[7,8,11] As such, SMBRs are nowadays being accepted increasingly as the technology of choice, as suggested by a number of indicators.[12]

The major drawback of SMBRs, which constitutes the motif for the slow acceptance of this technology, is the high investment costs, compared to other processes.[13] While the activated sludge plants have average cost to high value ratios, and the biological aerated filters have low-average cost to average value ratios, SMBRs are perceived as having high cost to high value ratios. Unless a high output quality is required for the discharged water, unreachable by conventional methods of treatment, investing large amounts of money in an SMBR will not be a priority.[10,12] Other drawbacks of SMBRs are the high oxygen demands caused by the long SRT, leading to a high aeration costs[8,13] and the membrane fouling.[10,14]

Many studies have been performed for the optimization of membrane modules used in gaseous mixtures separations (e.g. CO_2 removal from natural gas), methanol synthesis, etc., but few attempts were made for optimizing membrane bioreactors used in wastewater treatment and they are all experimental. Schoeberl *et al.* studied the influence of suction time, backwash time and aeration intensity on fouling, using a complete factorial experiment.[15] According to their paper, suction time was the most important parameter affecting the fouling rate, followed by aeration intensity and backwash time, which enhanced the fouling only at high suction times.

In this context, our study aims to find the optimal operating conditions and strategies for a semi-continuous SMBR system for wastewater treatment, using an improved version of a mathematical model previously developed and calibrated by Di Bella *et al.*[9]

The physical model

The physical model is an abstraction of the experimental set-up used by Di Bella *et al.*[9] to develop the mathematical model. The system is schematically presented in Fig. 1. The whole system works semi-continuously, which renders the mathematical treatment more complex. The raw wastewater passes through a 2 mm screen to remove hair, debris, rags, sand, etc. prior to be fed into the reservoir (1) (Fig. 1 for details). Thereafter, the waste-
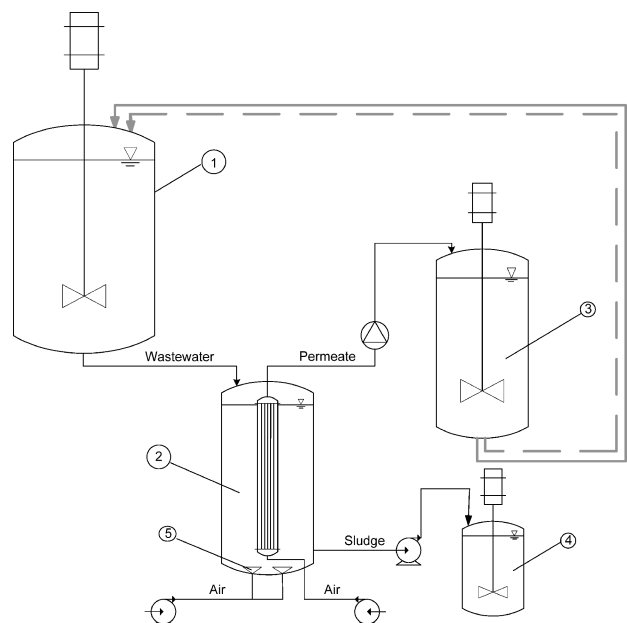

F i g . 1 – *Set-up of the system subject to optimization: 1 – wastewater tank; 2 – membrane bioreactor; 3 – permeate tank; 4 – sludge tank; 5 – air sparger. The continuous and the dashed grey lines symbolize the continuous and discontinuous recirculation of the permeate, respectively.*

water feeds the SMBR (2), where the organic matter and the nitrogen based compounds are removed. The permeate is withdrawn through the membrane module and stored in the tank (3). After a while, the feeding of wastewater and the permeate suction are interrupted and a fraction of the latter is pumped back through the membrane for a period of time to clean the pores and to remove the cake from the surface. The wasted sludge, which should prevent the accumulation in excess of dead cells, is continuously withdrawn from the bottom of the bioreactor and stored in the tank (4).

The bioreactor is aerated and stirred by two fine bubble air spargers located at its bottom, outside the membrane module. Another air sparger, placed under the membrane module, produces constant bubble swarms which generate strong turbulences when flowing tangentially with respect to the membrane surface; the air bubbles entrain continuously a part of the biomass stuck there, thus cleaning at least partially the surface of the membrane. The process takes place until the total consumption of the wastewater in the feed tank (1).

The mathematical model

The mathematical models developed for biological wastewater treatment in MBRs are generally classified in three categories:[9] biomass kinetic models, membrane fouling models and integrated models.

The kinetic models are mainly based on the activated sludge models[9,16–18] (ASMs) which have been modified to take into account the formation and degradation of the soluble microbial products (SMP) in the SMBR. The SMP category lumps all the soluble organic matters produced by mixed bacteria in the bioreactor. They are of crucial importance for biological wastewater treatment systems because of their significant impact on both effluent quality and treatment efficiency.[9] They are formed, on one hand, as a result of biological degradation of the initial substrate (the *utilization associated products*, UAP), and, on the other, as a result of biomass decay (the *biomass associated products*, BAP).[19]

The membrane fouling models are based upon solid-liquid separation (in which case the filtration process is replaced by an ideal settler with unitary efficiency) or upon the resistance-in-series representation. Mostly, they are empirical models.

The integrated models basically couple the kinetic and the fouling models (such as the resistance-in-series model) and they often consider the formation and degradation of SMP.[9]

For this study, the system presented in Fig.1 is abstracted into an integrated model consisting of two sub-models: one for the biological removal of the pollutants from the wastewater and the other for the physical removal of the solids from the permeate, due to cake deposition upon the surface of the membrane. The organic matter is lumped into two groups: the soluble matter formed by particles that can pass a 0.45 μm filter and the particulate matter, with particles larger than 0.45 μm, which are retained in the bioreactor by the membrane acting like an ideal filter with a homogenous distribution of the pores.[9] The microorganisms are classified in heterotrophs and autotrophs and the lag phase appearing when the operating conditions are changed is disregarded.

The dynamic equations of the biological process are based upon mass balances around the whole SMBR for all wastewater components; the Monod type kinetics applies for the different substrates.[9,17,20] The mass balances for two generic substrates are given by eq. (1), for the soluble case, and by eq. (2) for the insoluble case, respectively:

$$\frac{\mathrm{d}S_i^R}{\mathrm{d}t} = \frac{Q^{IN} \cdot S_i^{IN} + Q^B \cdot \overline{S_i^P} - Q^W \cdot S_i^R - Q^P \cdot S_i^R}{V^R} +$$
$$+ \sum \alpha_{S,i} \cdot \rho_{S,i}^{cons.} + \sum \beta_{S,i} \cdot \rho_{S,i}^{prod.}, \quad i = 1..M \tag{1}$$

$$\frac{\mathrm{d}X_i^R}{\mathrm{d}t} = \frac{Q^{IN} \cdot X_i^{IN} - Q^W \cdot X_i^R}{V^R} +$$
$$+ \sum \alpha_{X,i} \cdot \rho_{X,i}^{cons.} + \sum \beta_{X,i} \cdot \rho_{X,i}^{prod.}, \quad i = 1..N \tag{2}$$

where M is the number of soluble species, N is the number of particulate species, α and β represent the coefficients of the reaction rates, $\rho^{prod.}$ and $\rho^{cons.}$ are the formation and consumption rates of the soluble or insoluble species present in the wastewater. The complete stoichiometry and kinetic expressions are presented in Table 1.

When the permeate is withdrawn through the membrane during the filtration period, the soluble components (denoted by S_i in the previous equations) enter the permeate tank (3, in Fig. 1) with the concentration from the bioreactor, C_i^R, considered perfectly mixed. Here, the fresh permeate is mixed with the existing one collected in the previous filtration periods and a new concentration, $\overline{C_i^P}$, is reached. This average concentration results from a mass balance around the permeate tank (3), which is also assumed perfectly mixed:

$$\frac{\mathrm{d}\overline{C_i^P}}{\mathrm{d}t} = \frac{Q^P}{V^P} \cdot (C_i^R - \overline{C_i^P}), \quad i = 1..M \tag{3}$$

The mean concentration in the waste tank, $\overline{C}_i^W$, is obtained the same way:

$$\frac{\mathrm{d}\overline{C}_i^W}{\mathrm{d}t} = \frac{Q^W}{V^W} \cdot (C_i^R - \overline{C}_i^W), \quad i = 1..P \quad (4)$$

where P is the total number of species in the wastewater, soluble and insoluble.

The two periods, treatment and cleaning/backwash, differ only with respect to the values of the inlet flow, Q^{IN}, the permeate flow, Q^P, and the backflow for membrane cleaning, Q^B. During the treatment period, the backflow is zero, while throughout the cleaning period, which starts when cake resistance becomes too high, the inlet and the permeate flows are zero; the waste inlet flow, Q^W, is always constant and different from zero no matter the interval. These periodic changes cause the continuous variation of the average concentrations in the waste tank, while the permeate concentration remains constant during the backwashing.

The membrane sub-model describes the formation of cake layer on the surface of the membrane during the filtration period and the detachment during the backwash period.

Two opposite forces[21] act on a sludge particle approaching the membrane and affect its tendency of attachment: a drag force by suction which leads to attachment and a lifting force given by the turbulent flow of the bubble swarm, carrying the particle away from the membrane surface. In these conditions, the rate of biomass accumulation during the filtration period can be written as:[22]

$$\frac{\mathrm{d}M_{sf}}{\mathrm{d}t} = \frac{24 \cdot C_{SS} \cdot J^2}{24 \cdot J + C_d \cdot d_p \cdot G} - \frac{\beta \cdot (1-\alpha) \cdot G \cdot M_{sf}^2}{\gamma \cdot V_f \cdot t_f + M_{sf}} \quad (5)$$

T a b l e 1 – *Kinetics' coefficients and stoichiometry for the biological processes*

Process	Component							
	1	2	3	4	5	6	7	8
	S_S	X_S	X_{BH}	X_{BA}	S_{SMP}	S_O	S_{NO}	S_{NH}
Aerobic growth of heterotrophs	$-\dfrac{1}{Y_H}$		1		γ_{UAPH}	$-\dfrac{1-Y_H}{Y_H}$		$-i_{XB}$
					$\gamma_{UAPH} - \dfrac{1-Y_{SMP}}{Y_{SMP}}$	$-\dfrac{1-Y_{SMP}}{Y_{SMP}}$		$-i_{XB}$
Anoxic growth of heterotrophs	$-\dfrac{1}{Y_H}$		1		γ_{UAPH}		$-\dfrac{1-Y_H}{2.86 \cdot Y_H}$	$-i_{XB}$
			1		$\gamma_{UAPH} - \dfrac{1-Y_{SMP}}{Y_{SMP}}$		$-\dfrac{1-Y_{SMP}}{2.86 \cdot Y_{SMP}}$	$-i_{XB}$
Aerobic growth of autotrophs				1	γ_{UAPA}	$-\dfrac{4.57-Y_A}{Y_A}$	$\dfrac{1}{Y_A}$	$-i_{XB} - \dfrac{1}{Y_A}$
Particulate formation by decay of heterotrophs		$1-f_p$	-1					
BAP formation by decay of heterotrophs			-1		$1-f_B$			
Particulate formation by decay of autotrophs		$1-f_p$		-1				
BAP formation by decay of autotrophs				-1	$1-f_B$			
Ammonification of soluble organic nitrogen								1
Hydrolysis of entrapped organics	1	-1			1			
Hydrolysis of entrapped organic nitrogen								

T a b l e 1 – *continue*

Component				Reaction rates, ρ_i/g m^{-3} h^{-1}
9	10	11	12	
S_{ND}	X_{ND}	S_I	X_I	
				$\rho_{1\_1} = \mu_H \cdot \dfrac{S_S}{K_S + S_S} \cdot \dfrac{S_O}{K_{O,H} + S_O} \cdot X_{BH}$
				$\rho_{1\_2} = \mu_{SMP} \cdot \dfrac{S_{SMP}}{K_{SMP} + S_{SMP}} \cdot \dfrac{S_O}{K_{O,H} + S_O} \cdot X_{BH}$
				$\rho_{2\_1} = \mu_H \cdot \dfrac{S_S}{K_S + S_S} \cdot \dfrac{K_{O,H}}{K_{O,H} + S_O} \cdot \dfrac{S_{NO}}{K_{NO} + S_{NO}} \cdot \eta_g \cdot X_{BH}$
				$\rho_{2\_2} = \mu_{SMP} \cdot \dfrac{S_{SMP}}{K_{SMP} + S_{SMP}} \cdot \dfrac{K_{O,H}}{K_{O,H} + S_O} \cdot \dfrac{S_{NO}}{K_{NO} + S_{NO}} \cdot \eta_g \cdot X_{BH}$
				$\rho_3 = \mu_A \cdot \dfrac{S_{NH}}{K_{NH} + S_{NH}} \cdot \dfrac{S_O}{K_{O,A} + S_O} \cdot X_{BA}$
	$i_{XB} - f_P \cdot i_{XP}$		f_P	$\rho_4 = b_H \cdot X_{BH}$
$i_{XB} - f_B \cdot i_{XP}$		f_B		$\rho_5 = b_{BAP,H} \cdot X_{BH}$
	$i_{XB} - f_P \cdot i_{XP}$		f_P	$\rho_6 = b_A \cdot X_{BA}$
$i_{XB} - f_B \cdot i_{XP}$		f_B		$\rho_7 = b_{BAP,A} \cdot X_{BA}$
-1				$\rho_8 = k_a \cdot S_{ND} \cdot X_{BH}$
				$\rho_9 = k_h \cdot \dfrac{X_S}{k_X \cdot X_{BH} + X_S} \cdot \left(\dfrac{S_O}{K_{O,H} + S_O} + \eta_h \cdot \dfrac{K_{O,H}}{K_{O,H} + S_O} \cdot \dfrac{S_{NO}}{K_{NO} + S_{NO}} \right) \cdot X_{BH}$
1	-1			$\rho_{10} = \rho_9 \cdot \dfrac{X_{ND}}{X_S}$

During the cleaning process, the cake is not compressed further due to the lack of permeate suction. The compression coefficient, γ, may then be presumably reduced to one-tenth of its initial value and the biomass detachment is:[22]

$$\frac{\mathrm{d}M_{sf}}{\mathrm{d}t} = -\frac{\beta \cdot (1 - \alpha) \cdot G \cdot M_{sf}^2}{0.1 \cdot \gamma \cdot V_f \cdot t_f + M_{sf}} \qquad (6)$$

Based upon eq. (6), the backwashing period ends when there are no particles on the membrane surface, i.e. M_{sf} is zero.

The main drawback of this approach is that the permeate flux is assumed constant. The mathematical model has been solved using an in-house Matlab™ program based upon *ode15s* integration function for systems of stiff ordinary differential equations.

Measuring the performance of the system

In order to assess the completeness of the treatment, a performance criterion should be used, valid irrespective of the working strategy. The dynamic nature of the system given by the periodic change of the input and output flows, with consequences upon the biological transformation, renders more difficult the computation of such a criterion. For the present study, global conversion, defined as the ratio between the quantity of substrate consumed by microorganisms during a certain amount of time and the quantity initially fed in the supplying vessel and the bioreactor (1 and 2, Fig. 1), is a suitable metric in assessing the performance of the MBR at any moment. Noting with V_0 the initial volume of wastewater in the feed tank (1), with V the same volume corresponding to the time when the global conversion is evaluated and with V^R the volume of

the bioreactor, the amount of the generic substrate, m^{IN}, that has to be processed by the biomass is:

$$m^{IN} = (V_0 + V^R) \cdot c^{IN} \qquad (7)$$

where c^{IN} is the initial generic substrate concentration in the feed tank (1) and in the bioreactor (2).

At any given moment τ, the untransformed generic substrate is the sum between what was collected in the permeate and waste tanks and what remained untransformed in the bioreactor and the feeding vessel:

$$m^{OUT}(\tau) = \int_0^\tau Q^P \cdot c(t)dt + \int_0^\tau Q^W \cdot c(t)dt +$$

$$+ V^R \cdot c(\tau) + V(\tau) \cdot c^{IN} = \qquad (8)$$

$$= V^P(\tau) \cdot \overline{c}^P(\tau) + V^W(\tau) \cdot \overline{c}^W(\tau) + V^R \cdot c(\tau) + V(\tau) \cdot c^{IN}$$

where c is the concentration of generic substrate in the reactor, while $\overline{c}^P$ and $\overline{c}^W$ are its means in permeate and wasted sludge, respectively.

Using (7) and (8) the global conversion of the generic substrate is:

$$C(\tau) = \frac{m^{IN} - m^{OUT}(\tau)}{m^{IN}} = \qquad (9)$$

$$= 1 - \frac{V^P(\tau) \cdot \overline{c}^P(\tau) + V^W(\tau) \cdot \overline{c}^W(\tau) + V^R \cdot c(\tau) + V(\tau) \cdot c^{IN}}{(V_0 + V^R) \cdot c^{IN}}$$

Taking into account their importance into the economy of the treatment process and the quality of the discharged water, two were the pollutants for which the global conversion was computed: ammonia (C_{NH}) and soluble biodegradable substrate (C_S) whose formulas are obtained replacing the corresponding notations in (9).

Results and discussions

The base case

The concentration profiles for the state variables that describe the dynamic of the wastewater system including the SMBR were obtained by solving the mathematical model using the recommended values for the operating parameters:[9] $t_F = 9$ min and $k_L \cdot a_V = 2.5$ h^{-1}. The parameters of the model[9,17] are listed in Table 2 and the initial values for the state variables[9,17] are presented in Table 3.

The dynamic of the concentrations of the readily biodegradable substrates in the bioreactor and permeate are shown in Fig. 2, for a ratio of permeate to waste flows equal to two.

Table 2 – *The parameters of the mathematical model*

Para-meter	Value	Measuring units	Para-meter	Value	Measuring units
μ_H	0.25	h^{-1}	η_g	0.8	–
μ_A	0.0333	h^{-1}	η_h	0.4	–
μ_{SMP}	0.3458	h^{-1}	$\gamma_{UAP,H}$	0.82	–
b_H	0.092	h^{-1}	$\gamma_{UAP,A}$	1.03	–
$b_{BAP,H}$	0.0138	h^{-1}	Y_H	0.67	g$_{COD}$ g$_{COD}^{-1}$
b_A	0.0021	h^{-1}	Y_A	0.24	g$_{COD}$ g$_{NH4+}^{-1}$
$b_{BAP,A}$	0.0058	h^{-1}	Y_{SMP}	0.84	g$_{COD}$ g$_{COD}^{-1}$
k_a	0.0033	m^3 g$_{COD}^{-1}$ h^{-1}	i_{xb}	0.086	g$_N$ g$_{COD}^{-1}$
k_h	0.1250	g$_{COD}$ g$_{COD}^{-1}$ h^{-1}	i_{xp}	0.06	g$_N$ g$_{COD}^{-1}$
K_S	20	g m^{-3}	f_p	0.08	–
$K_{O,H}$	0.2	g m^{-3}	f_b	0.006	–
K_{SMP}	132.68	g m^{-3}	α	0.47	–
K_{NO}	0.5	g m^{-3}	β	0.01	–
K_{NH}	1	g m^{-3}	γ	0.09	kg m^{-3} h^{-1}
$K_{O,A}$	0.4	g m^{-3}	G	720000	h^{-1}
k_X	0.03	g$_{COD}$ g$_{COD}^{-1}$	S_O^*	10	g m^{-3}
V_0	0.3	m^3	V^R	0.19	m^3
S	0.93	m^2	J	0.0215	m^3 m^{-2} h^{-1}

Table 3 – *The initial values of the state variables*

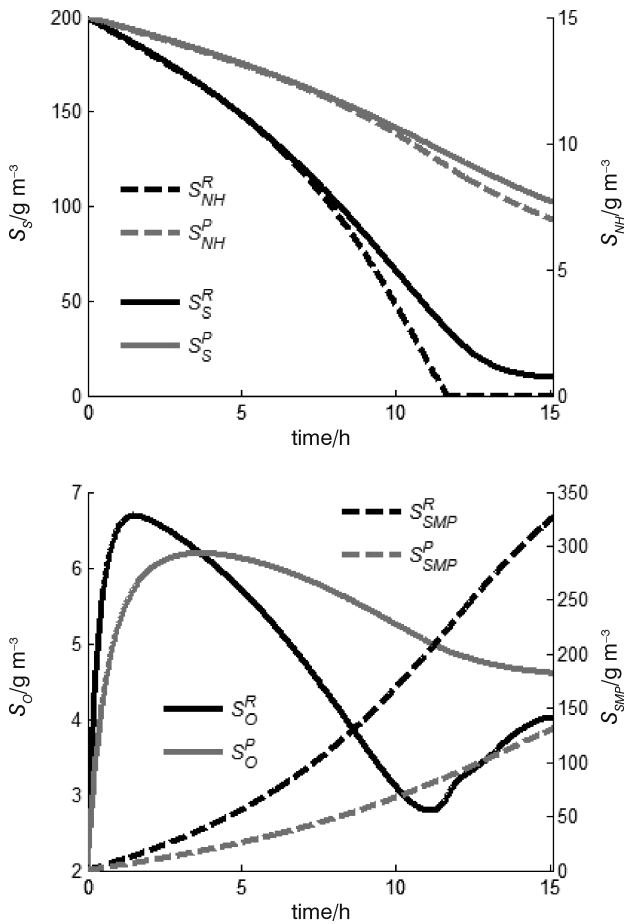| Component	Value	Measuring units
S_S	200	g m^{-3}
X_S	100	g m^{-3}
X_{BH}	25	g m^{-3}
X_{BA}	8	g m^{-3}
S_{SMP}	0	g m^{-3}
S_O	2	g m^{-3}
S_{NO}	1	g m^{-3}
S_{NH}	15	g m^{-3}
S_{ND}	9	g m^{-3}
X_{ND}	0	g m^{-3}
S_I	2	g m^{-3}
X_I	3	g m^{-3}
M_{sf}	0	kg m^{-2}

Fig. 2 – *Temporal profiles of the readily biodegradable substrates for the proposed values of the operating parameters[9] (S_S – carbonaceous substrate, g m^{-3}; S_{NH} – ammonia plus ammonium nitrogen, g m^{-3}; S_O – dissolved oxygen, g m^{-3}; S_{SMP} – soluble microbial products, g m^{-3}; superscript R – reactor, superscript P – permeate)*

Fig. 3 – *Concentration profiles of the heterotrophs (X_{BH}, g m^{-3}) and autotrophs (X_{BA}, g m^{-3}) for the proposed values of the operating parameters[9] (t_F = 9 min, $k_L a_V$ = 2.5 h^{-1})*

– **the second period** is characterized by an intensification of the biological processes as a result of both the augmentation of the heterotrophs concentration (Fig. 3) and the relatively high DO concentration. Consequently, the DO starts decreasing to a minimum value at which the flow of oxygen transferred from the gaseous phase equals the microorganisms' consumption. The oxygen consumed by heterotrophs increases continuously because they grow faster, thus diminishing the concentration of the readily biodegradable substrate, S_S, as well. In contrast, the reduction of both autotrophs and ammonia concentrations determines a decrease of oxygen used for nitrification;

– **the third period** is characterized by the resumed growth of the DO concentration. Although the concentration of heterotrophs is more than three times higher than in the first period (Fig. 3), implying higher oxygen consumption, the decrease in the concentration of carbonaceous substrate is such that it becomes rate limiting. At the same time, the ammonia reaches a minimum plateau (Fig. 2) sufficient for a process rate which allows the autotrophs to consume all the ammonia flow fed to the MBR. A change in the carbonaceous substrate profile can be observed (Fig. 2), from an accelerated decrease to an asymptotic one, as the kinetic passes from zero to first order as the substrate concentration drops.

The concentration of the soluble microbial products increases continuously, but a lower slope can be observed after the moment at which the ammonium and carbonaceous substrates decrease to very low levels in the reactor. The SMP are formed as a result of the activity or the degradation of the heterotrophs and autotrophs and become substrate

The combined analysis of the four profiles in the SMBR (Fig. 2) reveals three distinct regions throughout the whole working time:

– **the first period** is characterized by a pronounced increase in the concentration of the dissolved oxygen (DO); the mass transfer driving force (the difference between the saturation concentration of DO and the actual concentration) is high and as such the oxygen flux coming from air overcomes the microorganisms consumption by metabolic activities. The increase in the DO in the liquid phase causes a decrease of the mass transfer driving force. At the same time, however, as the concentration of DO increases, so does the consumption rate of some metabolic processes – see the reaction rates column in Table 1, lines 1–5, 11 and 12. Thus, there is a moment when the flow of oxygen consumed in the metabolic processes equals the flow of oxygen provided by the air and its concentration profile reaches a maximum;

for the heterotrophs. When the concentration of the ammonia (S_{NH}) and soluble substrate (S_S) reach their limits, the heterotrophs concentration continue to grow, but slower than before, causing a reduction in the amount of produced SMP.

The dynamic of heterotrophs and autotrophs is presented in Fig. 3, where a divergent behavior is observed. The growth of heterotrophs is accelerated until the concentration of readily biodegradable substrate, S_S, decreases to values for which the growth rate changes from zero to first-order. Simultaneously, the growth of heterotrophs diminishes also because of the DO low level.

The concentration of autotrophs decreases continuously with a constant rate given by the linear reduction of ammonia concentration. Once this substrate reaches its lowest limit, the reduction of autotrophs concentration caused by the imbalance between growth rate and death rate intensifies. For operating periods larger than 15 h, stabilization to a value ensuring balanced growth and death would be expected. It is worth noting that both heterotrophs and autotrophs vanish from the system through wastage line flow, which could be assimilated to cells death with respect to its effects.

The performance of the system, expressed as the global conversions of ammonia and carbonaceous substrates, is presented in Fig. 4. While the global conversion of ammonia is a sigmoid curve, in the case of the carbonaceous substrate the final asymptote corresponding to large operating periods does not exist, although the inflexion point had been already reached around 10 h. The behavior on the last part of the ammonia curve is the result of the decrease of the conversion caused by reaching the limit concentration of this substrate in the reactor, which is not the case for the readily biodegradable substrate.
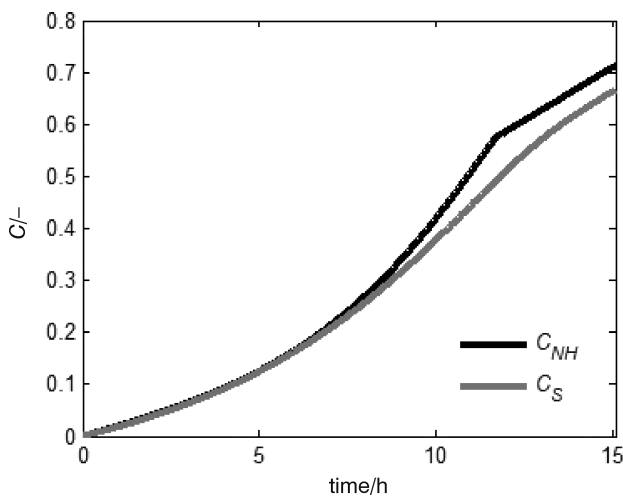
Throughout the operating period, the concentration of solids deposited on the surface of the membrane has periodic variations (not shown here). In the filtration step, the cake thickness increases linearly to a certain value, after which the cleaning period follows and it starts decreasing, also linearly, to almost zero at the end of the cleaning period. Although some studies[21,23–25] suggest that a thin cake layer could act as a secondary filter, thus preventing the membrane from a more severe fouling, our model doesn't provide any information about the characteristics of the cake, making it impossible for us to evaluate its contribution to the membrane fouling and filtration performance.

Sensitivity analysis against two candidate control variables

The parameters whose influence upon the behavior of the system was studied were *the filtration period, the volumetric mass transfer coefficient* and *the permeate to waste ratio*.

While Di Bella *et al.* chose two fixed values for the filtration and backwashing periods, we considered that the latter is not an independent variable, but it depends upon the quantity of solids deposited on the membrane surface, which increases with the length of the filtration period. Considering this, it may seem that the longer the cleaning period, the better. But during backwashing, the process is less efficient because the feeding is interrupted and the concentration of the microorganisms lowers due to the decreasing availability of the substrates and continuous waste sludge removal; therefore washing should last as little as possible.

For the filtration period, the interval of choice was between 5 and 20 minutes considering that at higher values the hypothesis of constant flux through the membrane could not be valid anymore due to the thickness of the cake.

The air plays an important role in the activity of the activated sludge and the good mixing of the liquid, but an air flow higher than necessary leads to unjustified aeration costs; the interval of choice for the volumetric mass transfer coefficient, which is a measure of the air flow, was between 1 and 6 h^{-1}.

The results have been expressed as the profiles of the global conversions of the ammonia and carbonaceous substrates at the end of the operating period against each of the two variables and are represented in Fig. 5.

The analysis shows that for both substrates the final global conversion has a maximum depending on the duration of filtration, situated at approximately 11 min, which is not far from the experimental value used by Di Bella.[9]
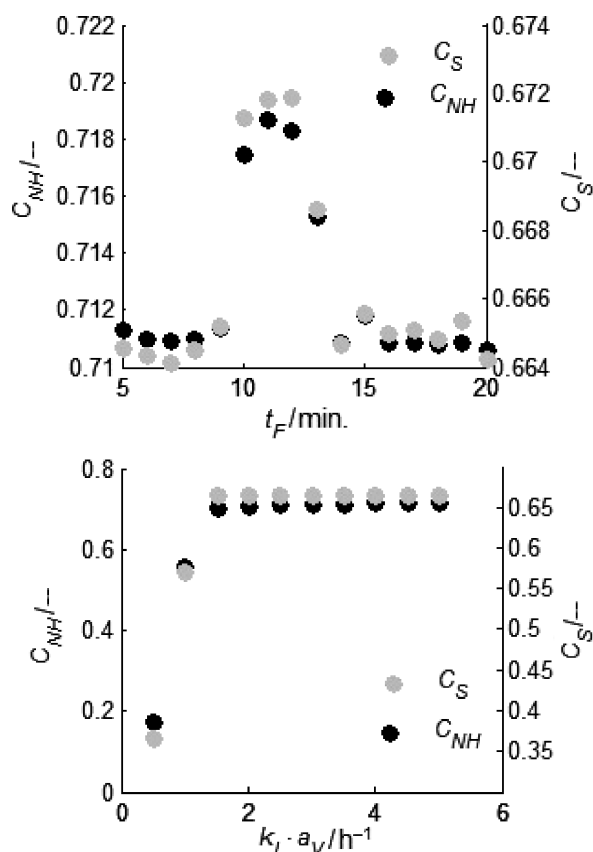

F i g . 4 – *Global conversions of the ammonia plus ammonium nitrogen and carbonaceous substrates for the proposed values of the operating parameters[9](t_F = 9 min, $k_L a_V$ = 2.5 h^{-1})*

F i g . 5 − *The variation of the final conversions of the ammonia plus ammonium nitrogen and carbonaceous substrates with the filtration period and volumetric mass transfer coefficient*

The volumetric mass transfer coefficient has a significant influence on the final global conversions only for values smaller than 1.5 h^{-1} after which both global conversions reach a plateau. This may suggest that after this point the value of this parameter becomes less important and that it would be ideal – from an economically point of view – to work closer to the beginning of this plateau. Nevertheless, the filtration period and the volumetric mass transfer coefficient are not independent. The influence of the filtration period on the global conversion was studied for $k_L \cdot a_V$ equal to 2.5 h^{-1}, while the influence of air flow-rates upon the global conversion was studied for the filtration period of 9 min. Therefore, both variables were chosen as commands in the optimization procedure.

Another component of the sensitivity analysis was the response of the system to the variations of the ratio between the permeate and the waste flows. The preliminary simulations for ratios between 1 and 10 showed that the performance of the system has increased asymptotically with this ratio. Consequently, ten was considered as working value and this ratio was left unchanged for all further simulations.

The optimization of the treatment process

The results of sensitivity analysis showed that the process is prone to optimization, using two control variables – the filtration period and the air flow, the latter with direct implications upon the availability of DO for the microorganisms. The length of the period in which the bioreactor is working at high substrate concentrations in the feed depends not only upon the biomass concentration, but also upon the air flow, which affects both the local turbulence near the surface of the membrane and the rate of oxygen mass transfer to microorganisms, represented by the volumetric mass transfer coefficient, $k_L \cdot a_V$.

The system at hand is discontinuous; therefore the optimization is restricted to the operating period which ends when the feed tank becomes empty. For this reason the objective function to be minimized is given by eq. (10) computed for the end of the working period (the ammonia was chosen as reference since it is the most aggressive pollutant).

$$f_{ob} = 1 - C_{NH}^{end} \tag{10}$$

The optimization has been carried out using the genetic algorithm toolbox from Matlab™, MathWorks, Natick, MA. The filtration period was varied within the range 5–20 min and the air flow within 0.18–0.72 m^3 h^{-1}. The number of chromosomes was 20 and the number of generations was set to l00. The following values for the operating parameters were obtained after optimization: t_F = 13.8 min and Q_{air} = 0.7 m^3 h^{-1}, corresponding to a specific mass transfer coefficient, $k_L \cdot a_V$, equal to 5.8 h^{-1}.

The performance of the system improves when working with the optimized values of the operating parameters. The final concentrations for the readily biodegradable and ammonia substrates are smaller, but still significantly higher than zero (Fig. 6). These residual concentrations are the consequence of the discontinuous operating mode of the system; immediately after the start-up the effluent concentrations are high and only after a rather long period they decrease to acceptable levels. This is caused by the fact that the pollutant flow consumed by the bacteria in the bioreactor is small at the beginning because the initial concentration of the cells is low.

As the quantity of substrate in the bioreactor decreases (the concentration diminishes through evacuation in the effluent and consumption in the biological process) the permeate is diluted to the final values in Fig. 6.

The improvement produced by the optimization can also be observed in the case of the global conversions (Fig. 6), when the values at the end of
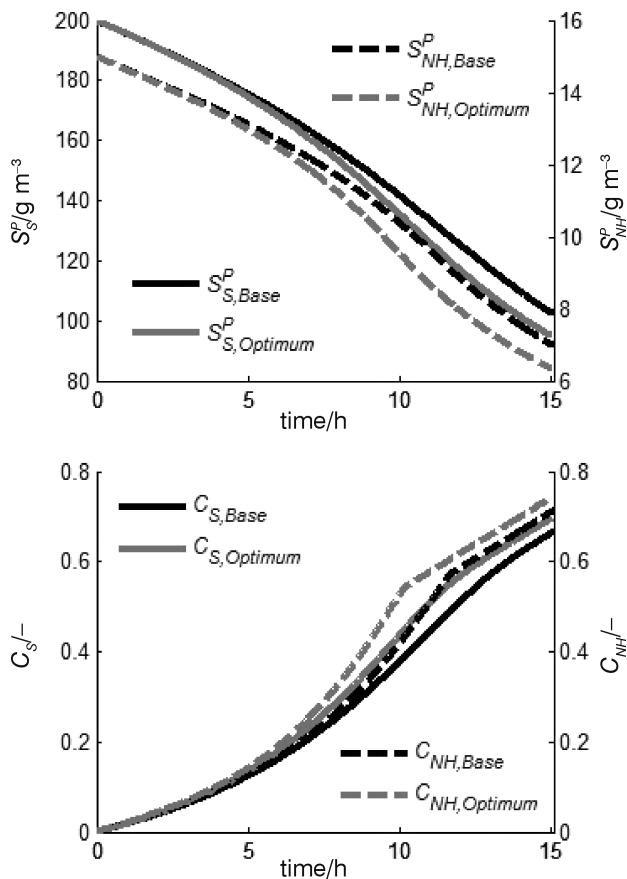
F i g . 6 – *Temporal profiles of the ammonia plus ammonium nitrogen and carbonaceous substrates in the permeate and their global conversions after optimization – comparison with the basic profiles*

the operating period are higher than those in the non-optimized case.

A promising strategy of increasing the global performance of the system is to recycle the permeate collected during the operating period for a supplemental treatment. This way, the unreacted substrates will be consumed by the biomass which reached higher concentrations in the bioreactor at the final of the working time.

The recycling of the permeate can be done two ways: either at the end of the operating period, by refilling the feed tank (1) and rerunning the process until the consumption of the recycled permeate (referred to as discontinuous recirculation), or by continually recycling a fraction of permeate in the feed tank (referred to as continuous recirculation). This fraction must be chosen so that the permeate tank (3) still holds the necessary amount of liquid for the cleaning of the membrane at all time.

The discontinuous recirculation is justified by the high biomass concentration at the end of the first operating period compared to the beginning of the process. The microorganisms, being now

in a much higher concentration, are capable of further reducing the residual pollutants in the permeate.

The continuous recirculation finds its justification in the need of reducing the high concentration of the pollutants in the effluent right from the beginning of the treatment process.

The effect of recycling the permeate

The dynamic of the system in response to the discontinuous recirculation of permeate is shown in Fig. 7. For the soluble biodegradable substrate, S_S, and ammonia plus ammonium nitrogen substrate, S_{NH}, respectively, the curves are typical, meaning their concentrations are diminishing during the second treatment. On the contrary, the DO profile has a specific evolution; in the bioreactor it has small variations given by the interplay between its level and the feeding flow of substrates into the bioreactor. At the beginning of the second period, DO sharpens its increase, due to the positive imbalance between the transfer rate from air bubbles and the rate of consumption into the biological processes. After a while, the consumption rate, which rises
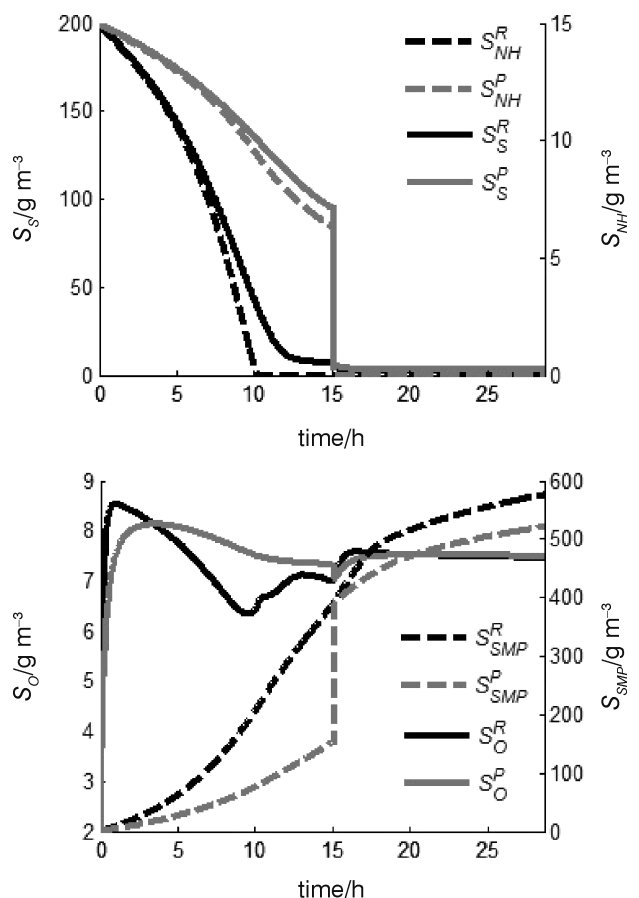

F i g . 7 – *Temporal profiles of the readily biodegradable substrates for the discontinuous recirculation (the notations are the same as in Fig. 2)*

based upon the higher DO, overcomes the supply rate, decreased by the decline of the mass transfer driving force, and therefore the DO concentration reaches a peak then starts falling. The drop is eventually counterbalanced by the raise of the mass transfer rate.

The SMP (Fig. 7) profile for the first operating cycle is the same as for single operated period; then, the concentration in the permeate increases sharply because the initial concentration in the permeate tank for the second period is the concentration in the reactor at the end of the first one, which is higher than at the beginning of the treatment process. Due to the fact that in the new operating period the heterotrophs are subjected to very low substrates levels (Fig. 7, ammonia and soluble substrate profiles, after 15 h), their concentration reaches a plateau (Fig. 9) and the increase in SMP is now much lower than before.

F i g . 9 – *Concentration profiles of the heterotrophs and autotrophs for the discontinuous recirculation (the notations are the same as in Fig. 3)*

F i g . 8 – *Global conversions of the ammonia plus ammonium nitrogen and carbonaceous substrates for the discontinuous recirculation (the notations are the same as in Fig. 3)*

Both conversions experience an important supplemental growth despite the lower concentrations of substrates in the new feeding and the operating conditions, which now are not optimal (Fig. 8). The significant drop in the slope of the conversions' increase witnesses these changes. The raise, although slower, is the result of the activity of the heterotrophs and autotrophs, whose concentrations are much higher than during the first period, for the former, and still sufficiently high, for the latter (Fig. 9).

The temporal profiles for the continuous recirculation of a fraction of the permeate are presented in Fig. 10. The operating period can be divided in two different regions, the first characterized by a significant reduction of the substrates entering the membrane bioreactor, when the process is
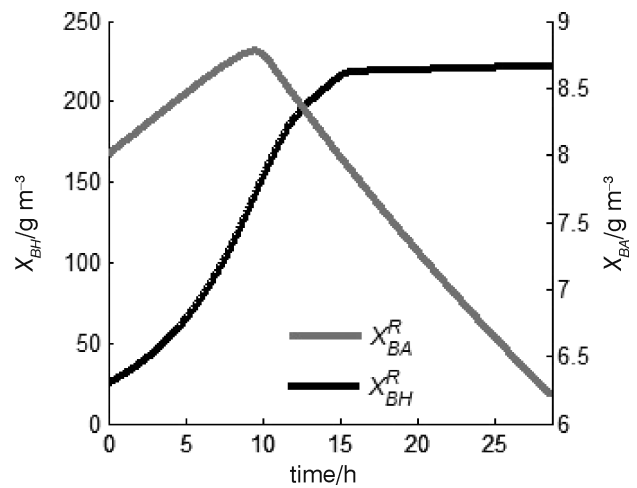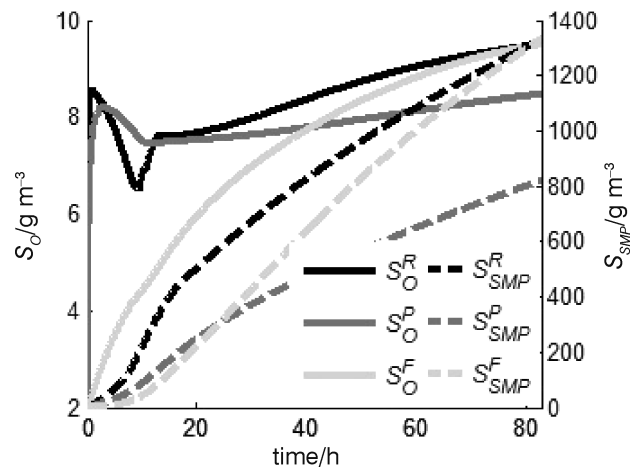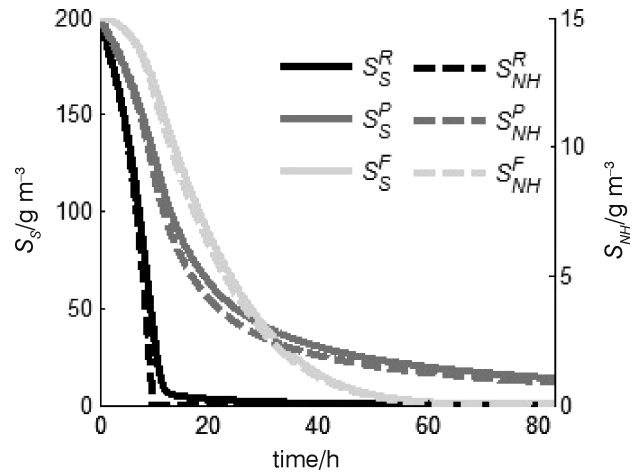

F i g . 1 0 – *Temporal profiles of the readily biodegradable substrates for the continuous recirculation (superscript F – feeding; the rest of the notations are the same as in Fig. 2)*

carried out at relatively high rates of oxygen consumption (Fig. 10, time up to 60 h), and the second, when the feeding concentrations become equal to the levels in the bioreactor (including SMP, whose

level at the end of the operating period is higher than in the two previous operating strategies). This leads to the reduction of the oxygen consumption so the system approaches saturation, which means that the aerobic biological processes are carried out at a minimum level. It should also be noted that after approx. 8 h, the concentration of ammonia in the bioreactor decreases to very low levels resulting in a reversal of autotrophs behavior – their concentration starts falling slowly but irreversibly, in average, to very low values (Fig. 11). Due to the minimum concentration of ammonia in the bioreactor, reached after 10 h, the growth rate of the autotrophs is smaller than the death rate for the current concentration of oxygen in the bioreactor, which makes the consumption of the oxygen smaller than the flux coming from the gaseous phase; thus, its concentration increases. This effect causes an enhancement of the growth rate of the heterotrophs which consume increasingly more oxygen but this consumption is not high enough to diminish the oxygen concentration (Fig. 10).
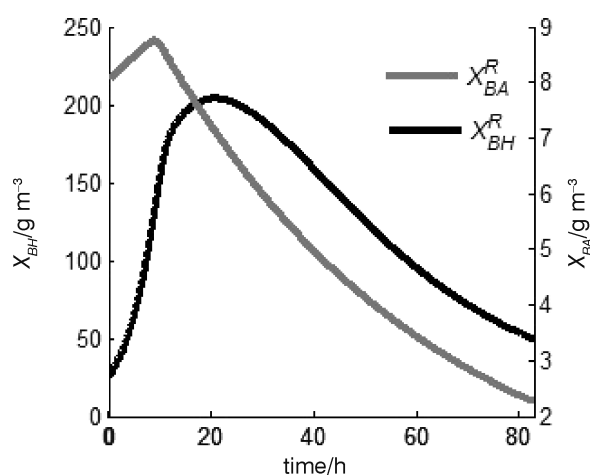

F i g . 1 1 – *Concentration profiles of the heterotrophs and autotrophs for the continuous recirculation (the notations are the same as in Fig. 3)*

The concentrations of microorganisms have typical profiles for growth in abundance and lack of substrate, respectively (Fig. 11). During the first period, when the ammonia has high values, both types of microorganisms grow as a result of substrate abundance. This growth stops when the concentration of the substrate becomes very small and the feed flow is not sufficient to ensure conditions for further growth. From this moment, the concentrations of the microorganisms start to decline, the autotrophs reaching very low levels.

It can be noted that the process already attained its high conversion limit, reached when the concentration of the readily biodegradable substrate
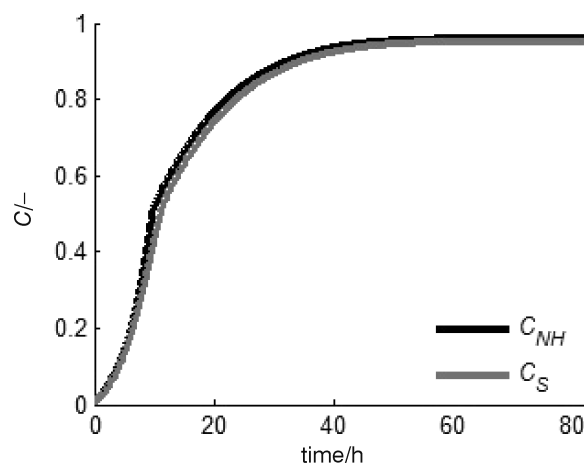

F i g . 1 2 – *Global conversions of the ammonia plus ammonium nitrogen and carbonaceous substrates for the continuous recirculation*

in the feed tank becomes very small (Fig. 10, after 60 h and Fig. 12). Thus, the further operation of the MBR becomes economically inefficient, the more so as the concentration of the ammonia drops to negligible values after 60 h of operation.

Conclusions

An improved version of a mathematical model describing a submerged membrane bioreactor was used to simulate the behavior of a discontinuous system for ammonia and carbon-based substrate removal from water, using a set of published values for the parameters. A sensitivity analysis was carried out with respect to the filtration period, air flow (which affects the specific mass transfer coefficient, $k_L \cdot a_V$) and permeate to wastage flow ratio, showing that the latter influences the process for uneconomic values only (lower than 3); for higher values, the effect is asymptotic. Consequently, the other two variables were chosen for an optimization study and can be seen as commands for better controlling the wastewater treatment with this new hybrid bioreactor. The optimal values found using the genetic algorithm routine implemented in Matlab™ are 13.8 min for the filtration period and 0.7 m³ h⁻¹ for the air flow.

The results corresponding to these operating conditions showed that the concentration of the pollutants (ammonia and carbonaceous substrates) in the permeate were smaller than in the un-optimized case, but not small enough to discharge the processed water into the environment. Two strategies were used to improve the performance of the biological wastewater treatment system already working under the optimal operating conditions, namely

the continuous recycling of a fraction of the permeate and the reuse of the permeate after the first cycle of treatment. In both cases we seek to break down to a convenient level the ammonia past through the bioreactor in the first half of the discontinuous working period, when the concentration of the microorganisms was too low to insure a convenient degradation.

Of these two methods, the discontinuous recycling proves to be the most efficient, lowering the water pollution under the level suitable for discharge into environment; at the matter of fact, the discharge water concentration of ammonia becomes zero.

A clearer picture of how the two optimal strategies (since they work with the optimal found parameters) of recycling the permeate improved the performance of the system at hand is shown in Fig. 13. As expected, there is a perfect match between the first period of the discontinuous recycling and the single period operated system. Then, in the former's case, the rest of the substrates are processed during the second period, profiting from availability of the biomass at high concentrations. The second strategy, the continuous recycling of a fraction of the permeate, gives the same performance till S_{NH} vanishes. After that moment, there is a slow decline in the performance, compared to the previous two operations, because the concentrations of the substrates in the feeding tank diminishes more and more, with respect to the values at which the biomass was exposed in the other two cases (Fig. 13, the lightest color line). Although this strategy gives close conversion compared to the discontinuous recycling of the permeate, the working time renders it less attractive.
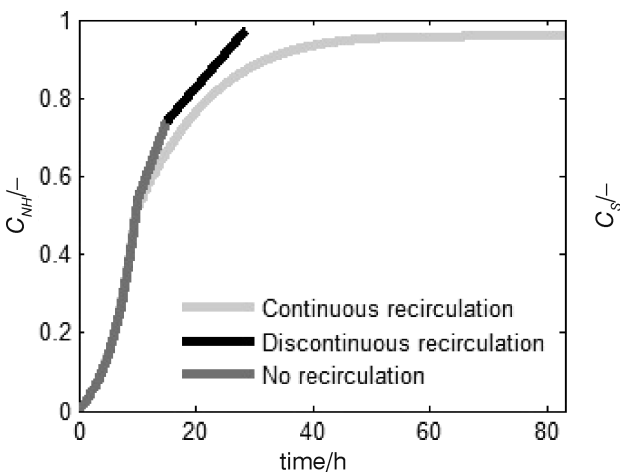

F i g . 13 – *Global conversions of the ammonia plus ammonium nitrogen and carbonaceous substrates for the three operating strategies*

Notations

b_A – autotrophic decay coefficient for formation of particulate, h^{-1}

b_H – heterotrophic decay coefficient for formation of particulate, h^{-1}

$b_{BAP,A}$ – autotrophic decay coefficient for formation of BAP, h^{-1}

$b_{BAP,H}$ – heterotrophic decay coefficient for formation of BAP, h^{-1}

c – concentration of generic substrate in the reactor, $g\ m^{-3}$

$\bar{c}^P$ – mean concentration of generic substrate in permeate tank, $g\ m^{-3}$

$\bar{c}^W$ – mean concentration of generic substrate in the wasted sludge tank, $g\ m^{-3}$

c^{IN} – the initial generic substrate concentration in the feed tank and in the bioreactor, $g\ m^{-3}$

C_d – lifting force coefficient, –

$\bar{C}_i^P$ – mean concentration of the component i in the pearmeate tank, $g\ m^{-3}$

C_i^R – the concentration of the component i in the reactor, $g\ m^{-3}$

$\bar{C}_i^W$ – mean concentration of the component i in the waste tank, $g\ m^{-3}$

COD – chemical oxygen demand, $g_{COD}\ m^{-3}$

C_{SS} – suspended solid concentration of biomass sludge, $kg\ m^{-3}$

d_p – particle size, μm

f_B – inert fraction of biomass leading to soluble products, –

f_P – inert fraction of biomass leading to particulate products, –

G – local shear intensity, h^{-1}

i_{xb} – ammonia fraction in biomass, $g_N\ g_{COD}^{-1}$

i_{xp} – ammonia fraction in particulate products, $g_N\ g_{COD}^{-1}$

J – permeate flux, $m^3\ m^{-2}\ h^{-1}$

k_a – ammonification coefficient, $m^3\ g_{COD}^{-1}\ h^{-1}$

k_h – hydrolysis coefficient, h^{-1}

K_{NH} – ammonia half-saturation coefficient for autotrophic biomass, $g_N\ m^{-3}$

K_{NO} – nitrate half-saturation coefficient for denitrifying heterotrophic biomass, $g_N\ m^{-3}$

$K_{O,A}$ – oxygen half-saturation coefficient for autotrophic biomass, $g_{O_2}\ m^{-3}$

$K_{O,H}$ – oxygen half-saturation coefficient for heterotrophic biomass, g_{O_2} m^{-3}

K_S – substrate half-saturation coefficient for heterotrophic biomass, g_{COD} m^{-3}

K_{SMP} – SMP half-saturation coefficient for heterotrophic biomass, g_{COD} m^{-3}

k_X – half-saturation coefficient for hydrolysis of particulate biodegradable substrate, –

M – the number of soluble species in the system

M_{sf} – dynamic sludge film cake on the membrane, kg m^{-2}

N – the number of particulate species in the system

P – the total number of species in the system, soluble and particulate

Q_{air} – air flow, m^3 h^{-1}

Q^B – backwashing flow rate, m^3 h^{-1}

Q^{IN} – influent flow rate, m^3 h^{-1}

Q^P – permeate flow rate, m^3 h^{-1}

Q^W – wasted flow rate, m^3 h^{-1}

S – surface of the membrane, m^2

S_I – soluble inert organic matter, g_{COD} m^{-3}

SMP – Soluble Microbial Products

S_{ND} – soluble biodegradable organic nitrogen, g_N m^{-3}

S_{NH} – ammonia plus ammonium nitrogen, g_N m^{-3}

S_{NO} – nitrate plus nitrite nitrogen, g_N m^{-3}

S_O – dissolved oxygen, g_{O_2} m^{-3}

S_S – soluble biodegradable substrate, g_{COD} m^{-3}

S_{SMP} – total soluble microbial product, equals to S_{BAP} plus S_{UAP}, g_{COD} m^{-3}

t – time, h

t_F – filtration period, min

V – actual volume of wastewater in the feeding tank, m^3

V_f – volume of permeate produced in a filtration period, m^3 m^{-2}

V^R – reactor volume, m^3

V_0 – initial volume of the wastewater, m^3

X_{BA} – active autotrophic biomass, g_{COD} m^{-3}

X_{BH} – active heterotrophic biomass, g_{COD} m^{-3}

X_I – particulate inert organic matter, g_{COD} m^{-3}

X_{ND} – particulate biodegradable organic nitrogen, g_N m^{-3}

X_S – particulate biodegradable organic matter, g_{COD} m^{-3}

Y_A – autotrophic yield coefficient, g_{COD} $g_{NH_4}^{-1}$

Y_H – heterotrophic yield coefficient from substrate, g_{COD} g_{COD}^{-1}

Y_{SMP} – heterotrophic yield coefficient from SMP, g_{COD} g_{COD}^{-1}

Greek letters

α – stickiness of biomass, –

β – erosion rate coefficient of dynamic sludge film, –

γ – compression coefficient for dynamic sludge layer, kg m^{-3} h^{-1}

$\gamma_{UAP,A}$ – UAP formation constant of autotrophs, –

$\gamma_{UAP,H}$ – UAP formation constant of heterotrophs, –

η_g – correction factor for anoxic growth of heterotrophs, –

η_h – correction factor for anoxic hydrolysis, –

μ_A – maximum specific growth rate for autotrophs, h^{-1}

μ_H – maximum specific growth rate of substrate for heterotrophs, h^{-1}

μ_{SMP} – maximum specific growth rate of SMP for heterotrophs, h^{-1}

ρ – reaction rate, g m^{-3} h^{-1}

Indices and superscripts

B – backwash

IN – inlet

P – permeate

R – reactor

W – waste

prod. – production

cons. – consumption

‾ – mean value

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