



# MODELLING THE EFFECT OF OXYGEN CONCENTRATION ON NITRITE ACCUMULATION IN A BIOFILM AIRLIFT SUSPENSION REACTOR

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## ABSTRACT

For an integrated nitrification-denitrification process, nitrite formation in the aerobic stage leads to big savings. Recently experimental observations (Garrido *et al.*, 1996) have shown that it is possible to obtain full ammonium conversion with approximately 50 % nitrate and 50 % nitrite in the effluent of a biofilm air-lift suspension reactor. With oxygen concentrations between 1 and 2 mg/l a maximum nitrite accumulation of 50 % was reached. Here we give a simple diffusion-reaction model describing these results. All the kinetic and mass transfer parameters were taken from the literature, except the mass transfer coefficient around the biofilm surface, which was fitted. The proposed model describes very well the measured data, despite the assumptions made. Using this model the influence of operational parameters was evaluated in order to establish ways to affect the  $\text{NO}_2^-$  concentration. None of these (surface loading,  $k_f$ , pH) had a significant effect on the maximal  $\text{NO}_2^-$  accumulation. Controlling the oxygen concentration seems to be the most practical method to obtain optimal nitrification to nitrite in BAS reactors.

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## KEYWORDS

Air-lift reactor; biofilm; mathematical modelling; nitrification; nitrite accumulation; orthogonal collocation.

## INTRODUCTION

Biofilm reactors are more efficient for nitrification than the classical activated sludge process since immobilisation of biomass enables to operate the reactor at a high sludge age and at high volumetric loading rate. However, in biofilters and in the fluidized bed reactors clogging usually occurs. One way to overcome these problems is the use of a biofilm air-lift suspension reactor (Tijhuis *et al.*, 1992). Full ammonia conversion to nitrate is possible for loading rates up to 5 kg N/m<sup>3</sup> reactor/day.

In an integral nitrogen removal process (nitrification + denitrification), intermediate nitrite formation is of interest because:

- (1) 25 % lower oxygen consumption in the aerobic stage implies 60 % energy savings,
- (2) in the anoxic stage the electron donor requirement is lower, and

(3) nitrite denitrification rates are 1.5 to 2 times higher than with nitrate.

Three main strategies for obtaining nitrite as the main product were proposed. The use of pure *Nitrosomonas* cultures immobilised in gels was proposed by Kokufuta *et al.* (1988) and Santos *et al.* (1992), but it is expensive and not very practical. Suthersand and Ganczarzyk (1986) used an activated sludge system in which pH is raised to get *Nitrobacter* inhibition. The system is however difficult to control because adaptation of the bacteria to high free ammonia concentrations occurs. The third possibility would be to control the dissolved oxygen level in the reactor. At lower oxygen concentrations the nitrite oxidation rate decreased more than the ammonia oxidation rate (Tanaka and Dunn, 1981; Tanaka and Dunn, 1982), but without full ammonia conversion.

Recently experimental observations (Garrido *et al.*, 1996) have proved that it is possible to obtain full ammonium conversion with approximately 50 % nitrate and 50 % nitrite in the effluent. With oxygen concentrations between 1 and 2 mg/l a maximum nitrite accumulation was reached. Moreover, the nitrite formation was stable during several months, but also during a short term experiment. Therefore, it was suggested that the influence of dissolved oxygen on nitrite accumulation is an intrinsic characteristic of the biofilms in the BAS reactor.

Here we try to give a simple model formulation describing these results. Using this model the influence of operational parameters was evaluated in order to establish ways to affect the  $\text{NO}_2^-$  concentration.

#### BASIS FOR MODEL

- The BAS reactor is continuously operated at a constant flow rate and loading rate. After each switch in dissolved oxygen concentration a steady-state in the concentration of soluble compounds was achieved.
- The average particle diameter and biofilm thickness are constant, because the time scale of the experiment with variable dissolved oxygen is much smaller than the characteristic time of growth (biofilm in “frozen state”). Moreover, the biomass concentration profile in the biofilm depth is also considered constant.
- Retention time distribution experiments showed that the liquid phase of the biofilm airlift suspension reactor is completely mixed (Garrido *et al.* 1996).
- The proportion between ammonia and nitrite oxidisers in the biofilm was assumed based on a review by Wiesmann (1994).
- For the processes occurring inside the biofilm we used the classical model with diffusion and reaction in a catalyst bead.
- External mass transfer resistance was taken into account for all components.
- Decay and endogenous processes have been neglected because they have no large influence on ammonia conversion.

All parameters of the model are in the **Table 1**.

#### Stoichiometry

We can write the global stoichiometry for growth of ammonium oxidising bacteria as:

$$-\frac{1}{Y_{XN1}}(\text{NH}_4^+) - \frac{1}{Y_{XO1}}(\text{O}_2) - \dots + (\text{C}_5\text{H}_7\text{NO}_2) + \frac{1}{Y_{XN1}}(\text{NO}_2^-) + \dots = 0 \quad (1)$$

and for growth of nitrite oxidising bacteria as:

$$-\frac{1}{Y_{XN2}}(\text{NO}_2^-) - \frac{1}{Y_{XO2}}(\text{O}_2) - \dots + (\text{C}_5\text{H}_7\text{NO}_2) + \frac{1}{Y_{XN2}}(\text{NO}_3^-) + \dots = 0 \quad (2)$$

#### Kinetics

The substrate conversion kinetics can be expressed as:

$$r_{S1} = -q_{S1} C_{X1} \quad (3)$$

$$r_{S2} = -q_{S2} C_{X2} \quad (4)$$

where specific rates are modelled by Monod equations with double limitation of donor species (ammonia  $S_1$  or nitrous acid  $S_2$ ) and acceptor (oxygen). If the nitrogen substrates are present in high concentrations, both ammonia and  $\text{HNO}_2$  could be inhibitors and this is represented by the following equations:

$$q_{S1} = q_{m,S1} \cdot \frac{C_{NH_3}}{K_{S1} + C_{NH_3} + \frac{C_{NH_3}^2}{K_{i1}}} \cdot \frac{C_O}{K_{O1} + C_O} \quad (5)$$

$$q_{S2} = q_{m,S2} \cdot \frac{C_{HNO_2}}{K_{S2} + C_{HNO_2} + \frac{C_{HNO_2}^2}{K_{i2}}} \cdot \frac{C_O}{K_{O2} + C_O} \quad (6)$$

Concentrations of nitrogen neutral species can be related to the concentrations of ionic species with the equilibrium equations:

$$C_{NH_3} = \frac{C_{NH_4} 10^{pH}}{K_a} \quad \text{with} \quad K_a = \exp\left(\frac{6344}{273+t}\right) \quad (7, 8)$$

$$C_{HNO_2} = \frac{C_{NO_2}}{K_b 10^{pH}} \quad \text{with} \quad K_b = \exp\left(\frac{-2300}{273+t}\right) \quad (9, 10)$$

### Component balances

- *in bulk liquid* - the mass balance over the liquid phase for soluble components:

$$\frac{V_R}{Q_{in}} = \frac{C_{i,in} - C_{i,l}}{\Phi_i a_v} \quad (11)$$

with  $\Phi_i$  the mass flux between liquid and biofilm.

- *in biofilm* - for each relevant component  $i$  we can write a mass balance, taking into account diffusion processes in a spherical geometry (with an effective diffusion coefficient  $D_{i,eff}$ ) and an overall reaction rate in the catalyst particle ( $r_i$ ):

$$D_{i,eff} \cdot \left( \frac{d^2 C_i}{dr^2} + \frac{2}{r} \frac{dC_i}{dr} \right) + r_i = 0 \quad (12)$$

where the net reaction rates are:

$$\text{NH}_4 : \quad r_{NH_4} = -q_{S1} C_{X1} \quad (13)$$

$$\text{O}_2 : \quad r_{O_2} = -\frac{Y_{XN1}}{Y_{XO1}} q_{S1} C_{X1} - \frac{Y_{XN2}}{Y_{XO2}} q_{S2} C_{X2} \quad (14)$$

$$\text{NO}_2 : \quad r_{NO_2} = q_{S1} C_{X1} - q_{S2} C_{X2} \quad (15)$$

$$\text{NO}_3 : \quad r_{NO_3} = q_{S2} C_{X2} \quad (16)$$

The system of second order differential equations needs two sets of boundary conditions, one at the inner limit of the biofilm (carrier-biofilm interface), and the biofilm surface (biofilm-diffusion film):

$$\left. \frac{dC_i}{dr} \right|_{r=r_c} = 0 \quad (\text{zero fluxes at the carrier surface}) \quad (17)$$

and

$$D_{i,eff} \left. \frac{dC_i}{dr} \right|_{r=r_c+\delta} = k_l (C_{i,l} - C_i|_{r=r_c+\delta}) = \Phi_i \quad (\text{equality of fluxes at the biofilm outer surface}) \quad (18)$$

## MODEL SOLVING

## Method

The mathematical model of the biofilm air-lift suspension reactor consists of a system of coupled parabolic differential equations with double boundary condition (species in biofilm balances) and a system of algebraic equations (species in liquid balances).

In order to solve the model we have applied an orthogonal collocation method, as described in Finlayson (1972). Firstly, the model was made dimensionless by a series of substitutions. Concentrations were related to a reference concentration, for instance concentration of ammonium in the inlet flow,  $C_{NH_4, in}$ , so that  $c_i = C_i / C_{NH_4, in}$ . Dimensionless coordinate was  $X = (r - r_c) / \delta$  so that collocation points could have values between 0 and 1.

Discretisation of the dimensionless model results in a non-linear system of  $N-2$  equations for each component balance in the collocation points:

$$\left( \sum_{k=1}^N B_{j,k} c_{i,k} + \frac{2}{r_c / \delta + X_j} \sum_{k=1}^N A_{j,k} c_{i,k} \right) + \varphi_i^2 R_{i,j} = 0 \quad \text{for } j = 2 \dots N-1, \quad i = 1 \dots 4 \quad (19)$$

and boundary conditions:

$$X = 0 \quad \sum_{k=1}^N A_{1,k} c_{i,k} = 0 \quad \text{for } i = 1 \dots 4 \quad (20)$$

$$X = 1 \quad \frac{1 - c_{i,l}}{\sum_{k=1}^N A_{N,k} c_{1,k}} - Da_1 = 0 \quad \text{for } i = 1 \text{ (ammonium)} \quad (21)$$

$$\frac{-c_{i,l}}{\sum_{k=1}^N A_{N,k} c_{i,k}} - Da_i = 0 \quad \text{for } i = 2 \dots 4 \quad (22)$$

where external diffusion resistance leads to :

$$c_{i,l} = c_{i,N} + \frac{1}{Bi_i} \sum_{k=1}^N A_{N,k} c_{i,k} \quad (23)$$

The system is now described by a set of dimensionless numbers: Thiele modulus  $\varphi^2 = \delta^2 \frac{q_{sm} C_X}{D_{eff} C_{NH_4, in}}$ ,

Biot number  $Bi = \frac{k_l \delta}{D_{eff}}$ , a Damkohler-like number  $Da = \frac{V_R}{Q_{in}} \cdot \frac{a_v D_{eff}}{\delta}$ , dimensionless affinity and inhibition

constants, and a geometric parameter which is the  $r_c / \delta$  ratio.

For  $O_2$  dissolved in the liquid phase we have a constant (and measured) value,  $C_{2,l} = C_{dissolved\ oxygen}$ . The calculation of the steady-state profiles of dissolved components consists finally in the solution of a non-linear system of coupled algebraic equations. The system was solved by a classical Newton-Raphson iterative procedure. The main difficulty was to choose an adequate set of start values, because the procedure diverges if the starting profiles are too far from the solution profiles. In order to avoid this problem, an analytic continuation method was chosen (Reichert *et al.*, 1989). For high diffusion coefficients and low retention times the problem is solved easily, since the concentrations in the biofilm are constant and have the same values with the concentrations in liquid. Furthermore, these are equal with the inlet concentrations since the low retention times maintain constant values in the bulk liquid. This simplified problem is solved firstly and then hypotheses are gradually relaxed: the effective diffusion coefficient is decreased and the residence time is increased till they reach their correct values. Solved profiles for the first dissolved oxygen concentration are used then as initial approximation for another level of dissolved oxygen, till the entire domain is scanned.

Our program runs showed that a grid of 8 to 10 collocation points is sufficient for an accurate solution of this problem. Calculations were performed with a source written in BPascal 7.0 on an IBM 486 computer.

The computing time necessary for determination of bulk concentrations in the range of 0 to 4 mg dissolved  $O_2$  /L was approx. 60 seconds with 9 collocation points in the biofilm and 250 steady-states function of dissolved oxygen.

### Input parameters

All parameters used are summarised in table 1.

**Table 1.** Model parameters

Parameter	Symbol	Value	Units	Reference
<b>Kinetics</b>				
Maximal activity of ammonia-oxidisers	$q_{m1}$	13.4	mg N-NH <sub>3</sub> / mg VSS / day	(a)
Maximal activity of nitrite-oxidisers	$q_{m2}$	44.74	mg N-HNO <sub>2</sub> / mg VSS / day	(a)
Monod saturation constant of NH <sub>3</sub> for ammonia-oxidisers	$K_{S1}$	$2.8 \cdot 10^{-2}$	mg N-NH <sub>3</sub> / L	(a)
Monod saturation constant of HNO <sub>2</sub> for nitrite-oxidisers	$K_{S2}$	$3.2 \cdot 10^{-5}$	mg N-HNO <sub>2</sub> / L	(a)
Inhibition constant of NH <sub>3</sub> for ammonia-oxidisers	$K_{i1}$	540	mg N-NH <sub>3</sub> / L	(a)
Inhibition constant of HNO <sub>2</sub> for nitrite-oxidisers	$K_{i2}$	0.26	mg N-HNO <sub>2</sub> / L	(a)
Monod saturation constant of oxygen for ammonia-oxidisers	$K_{O1}$	0.3	mg O <sub>2</sub> / L	(a)
Monod saturation constant of oxygen for nitrite-oxidisers	$K_{O2}$	1.1	mg O <sub>2</sub> / L	(a)
Growth yield of ammonia-oxidisers on ammonia	$Y_{XN1}$	0.147	mg VSS / mg N-NH <sub>3</sub>	(a)
Growth yield of nitrite-oxidisers on HNO <sub>2</sub>	$Y_{XN2}$	0.042	mg VSS / mg N-HNO <sub>2</sub>	(a)
Growth yield of ammonia-oxidisers on oxygen	$Y_{XO1}$	0.046	mg VSS / mg O <sub>2</sub>	(a)
Growth yield of nitrite-oxidisers on oxygen	$Y_{XO2}$	0.039	mg VSS / mg O <sub>2</sub>	(a)
<b>Mass transfer</b>				
Diffusion coefficient of NH <sub>4</sub> <sup>+</sup> in water	$D_1$	$1.86 \cdot 10^{-9}$	m <sup>2</sup> /s	(b)
Diffusion coefficient of O <sub>2</sub> in water	$D_2$	$2 \cdot 10^{-9}$	m <sup>2</sup> /s	(b)
Diffusion coefficient of NO <sub>2</sub> <sup>-</sup> in water	$D_3$	$1.7 \cdot 10^{-9}$	m <sup>2</sup> /s	(b)
Diffusion coefficient of NO <sub>3</sub> <sup>-</sup> in water	$D_4$	$1.7 \cdot 10^{-9}$	m <sup>2</sup> /s	(b)
Liquid-solid mass transfer coefficient	$k_l$	3.5 .. 7.0	m / d	(b), (d), (e)
Porosity of biofilm	$\epsilon$	0.7	m <sup>3</sup> water / m <sup>3</sup> biofilm	assumed
<b>Geometry and operation</b>				
Total reactor volume	$V_R$	3	L	measured, (c)
Liquid input flowrate	$Q_{in}$	72	L / d	measured, (c)
Ammonium concentration in input wastewater	$C_{NH_3,in}$	200	mg N-NH <sub>4</sub> / L	measured, (c)
Specific area	$a_v$	3000	m <sup>2</sup> interface / m <sup>3</sup> reactor	calculated, (c)
Carrier diameter	$d_c$	0.3	mm	measured, (c)
Biofilm thickness	$\delta$	0.2	mm	measured, (c)
pH		7		measured, (c)
Temperature	$t$	30	°C	measured, (c)
Ammonia-oxidisers concentration in biofilm	$C_{X1}$	54	g VSS / L	(a)
Nitrite-oxidisers concentration in biofilm	$C_{X2}$	16	g VSS / L	(a)
Total density of biomass in biofilm	$C_X$	70	g VSS / L	measured

(a) Wiesmann (1994), (b) Tjihuis (1994), (c) Garrido *et al.* (1996), (d) Hunik *et al.* (1994), (e) Wijffels *et al.* (1991)

**Kinetic coefficients.** There are several reports which describe the kinetic behaviour of nitrifying bacteria. Wiesmann (1994) presented some averaged coefficients of widely scattered data collected from literature, together with own recent measurements. In this paper we have used their kinetic model and parameters, with the sole assumption that the decay and endogenic processes are not significant for the overall conversion of N-compounds. Maximum specific growth rates were corrected for 30 °C.

Because differentiation of the two nitrifying species is difficult we have assumed a constant ratio between ammonia and nitrite oxidisers, calculated supposing maximum conversion rates of substrates and complete conversion of nitrite intermediate: (Wiesmann, 1994)

$$\frac{C_{X2}}{C_{X1}} = \frac{q_{m1}}{q_{m2}} = 0.29 \quad (24)$$

Since the measured density of biomass after 120 days was 70 g/l biofilm, 54 g/l biofilm  $\text{NH}_4$ -oxydizers and 16 g/l biofilm  $\text{NO}_2$ -oxydizers were assumed as densities in the biofilm during the experiment.

*Mass transfer coefficients.* Data for diffusion coefficients in gel beads and biofilms ( $D_{i,eff}$ ) are also scattered (Hunik, 1994). We assume the gel porosity of biofilm equal to 0.7 (very probable according with our light microscopic photographs and the high biomass density in the biofilm), diffusion coefficients of  $\text{O}_2$  and ionic species in water should be halved.

The mass transfer coefficient between liquid phase and solid biocatalyst in a BAS reactor was calculated as described by Tjihuis (1994). Assuming the liquid velocity around the particle estimated as the settling velocity of a single particle in a stagnant liquid, we can obtain a rough approximation of oxygen mass transfer coefficient  $k_l = 8.7 \cdot 10^{-5} \text{ m/s} = 7.5 \text{ m/d}$ . This value was chosen as starting point for the data fitting procedure. Another value, calculated as described by Wijffels *et al.* (1991) for an air-lift reactor was 3.5 m/d. Due to the close diffusion coefficients of nitrogen species and oxygen,  $k_l$  was set equal for all components and fitted at the value 4.4 m/d.

*Geometric and operation parameters.* All these parameters were determined experimentally in our laboratory. Ammonium loading rate was 5 kg  $\text{N-NH}_4/\text{m}^3\text{d}$ . The hydraulic retention time, determined experimentally with a pulse signal of Blue Dextran (not diffusable in biofilm), was 0.5 h. The volume fraction of solids was 0.41 and gas fraction about 0.07 (Garrido *et al.*, 1996), the overall retention time is approx. 1 h. The temperature was 30 °C and the acidity was kept around pH 7.

## RESULTS

### Model evaluation

Figures 1a-d show oxygen, ammonium, nitrite and nitrate profiles calculated along the dimensionless radius of the biofilm. An arbitrary length of stagnant film (0.2 units) was chosen only for representation of external profiles on these graphs.

It can be seen that oxygen is indeed the rate limiting substrate (fig. 1a). The penetration depth of  $\text{O}_2$  in the biofilm rises with an increasing oxygen concentration in the bulk volume, allowing the respiration of a greater amount of biomass and consequently a faster ammonia conversion. For  $C_{\text{O}_2} = 3 \text{ mg/l}$  in the bulk liquid, the penetration depth  $\delta_{\text{O}_2} > 100 \mu\text{m}$  is sufficient so that all nitrite produced by ammonia-oxidisers can be converted by the slower metabolism of nitrite-oxidisers.

As figures 1b-d show, neither internal nor external diffusional resistance affects significantly the penetration of ammonia in the biofilm or the release of products in the bulk liquid. These results revealed that mass transfer affects mainly the oxygen conversion which is dissolved at low level in the reactor.

Experimental and calculated steady-states are shown in figure 2. It appears that the proposed model fits very well the measured data, despite the assumptions made.

A better validation of the model would be possible if the spatial distribution of two microorganisms in biofilms were determined with available techniques (Hunik *et al.*, 1993).

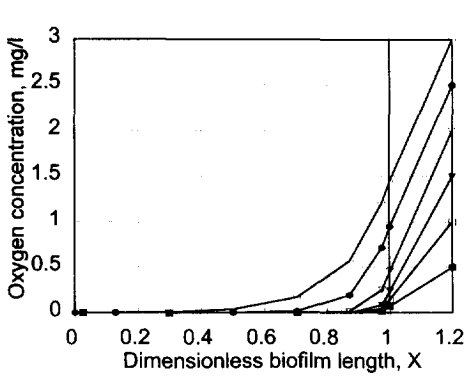


Figure 1a. Calculated profiles of oxygen in biofilm

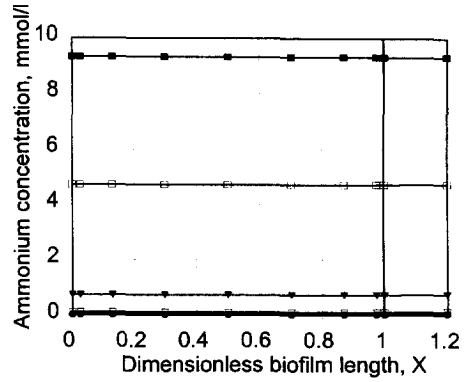


Figure 1b. Calculated profiles of ammonium in biofilm

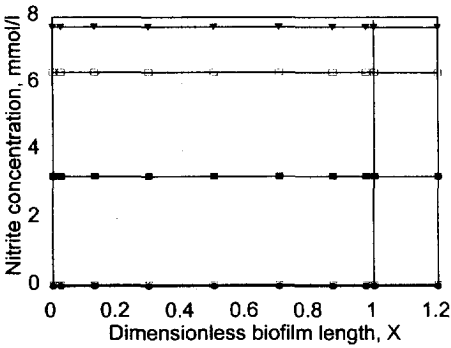


Figure 1c. Calculated profiles of nitrite in biofilm

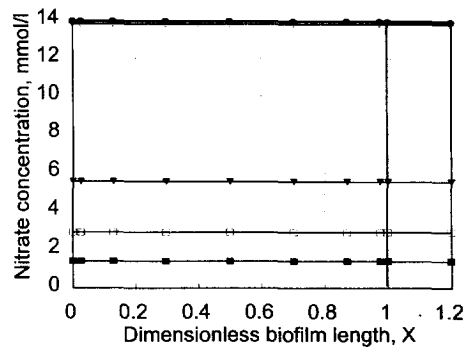


Figure 1d. Calculated profiles of nitrate in biofilm

Symbols: dissolved oxygen concentration

■ 0.5, □ 1.0, ▼ 1.5, ▽ 2.0, ● 2.5, ○ 3.0 mg/l

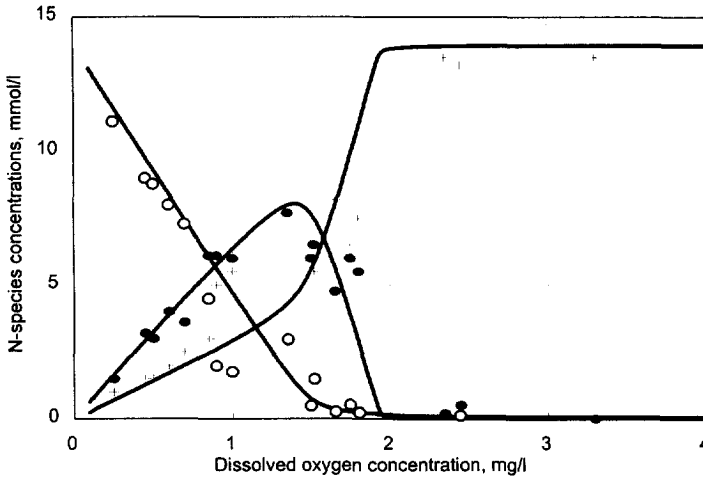


Figure 2. Experimental and calculated steady-states at different dissolved oxygen concentrations:

○ ammonium , ● nitrite , + nitrate .

### Evaluation of process conditions

The influence of operational parameters was evaluated in order to establish procedures to maximise the  $\text{NO}_2^-$  production.

Firstly the influence of ammonia specific loading rate (ALR,  $\text{kg N/m}^2_{\text{biofilm}} \text{ d}$ ) was studied. The amount of ammonia converted per biofilm area depends on the hydraulic retention time, specific area of biofilm, nitrogen concentration in the influent and the liquid hold-up and is defined as:

$$ALR = \frac{C_{\text{NH}_3, \text{in}} Q_{\text{in}}}{V_R a_v} \quad (25)$$

An increased ammonium loading rate has no effect on the maximum achievable  $\text{NO}_2^-$  accumulation. The only effect is that the maximum accumulation is reached at higher DO values (figure 3). This is in line with the expectation that oxygen diffusion becomes more limiting at higher ammonium loading rates.

Changing the operating pH affects the ammonium oxidation rate since  $\text{NH}_3$  is the substrate. Diffusion itself is marginally affected. An increasing pH has indeed a positive effect on nitrite formation (figure 4).

Finally the effect of mass transfer coefficients (reactor turbulence) is evaluated. In fact, a similar effect as for ammonium loading rate is observed. At high mass transfer rates a lower DO can be maintained for optimal  $\text{NO}_2^-$  accumulation.

Overall it can be concluded that changing operational conditions will not directly result in a changed accumulation of nitrite. Only the DO at which this maximum is reached can be affected.

### CONCLUSIONS

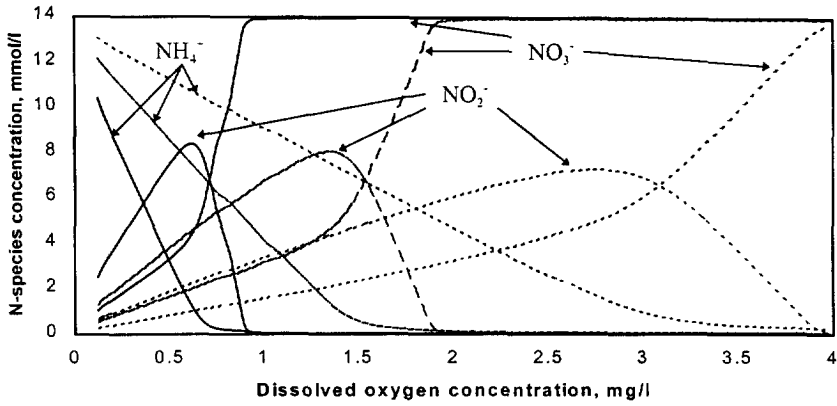
We could say that the proposed model describes the measured data properly, despite the assumptions made. All the kinetic and mass transfer parameters were taken from the literature, except the mass transfer coefficient around the biofilm surface which was fitted.

Neither internal nor external diffusional resistance affects significantly the penetration of ammonia in the biofilm or the release of products in the bulk liquid. These results confirmed that mass transfer of oxygen is the main limiting factor.

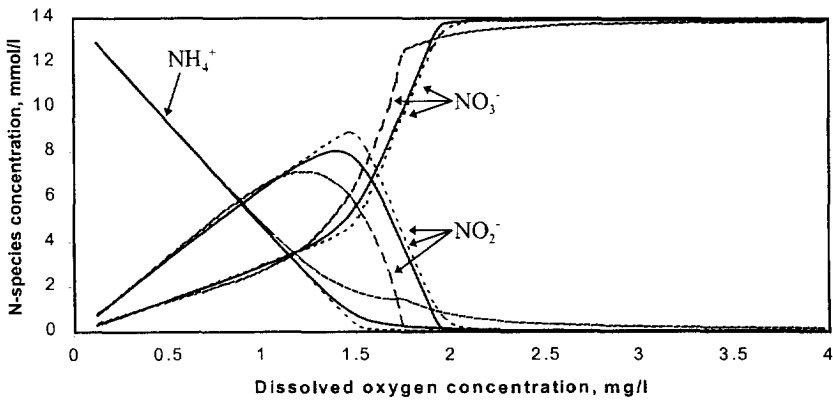
It was not needed to assume a distribution of nitrite oxidisers and nitrate oxidisers into the biofilm even if this could occur. Distribution of bacteria over biofilm depth depends on the concentration field of rate limiting substrate (oxygen in this case) and a better validation of the model would be possible if the spatial distribution of the two populations were determined with appropriate techniques.

The influence of operational parameters was evaluated in order to establish ways to improve  $\text{NO}_2^-$  production. The model was used to appreciate the effect of loading rate, mass transfer coefficient and pH on the nitrite concentration in the reactor. Influencing the oxygen concentration seems to be the most practical method to obtain partial nitrification to nitrite in BAS reactors since this can be done by varying the superficial gas velocity or by partial recirculation of the off-gas.

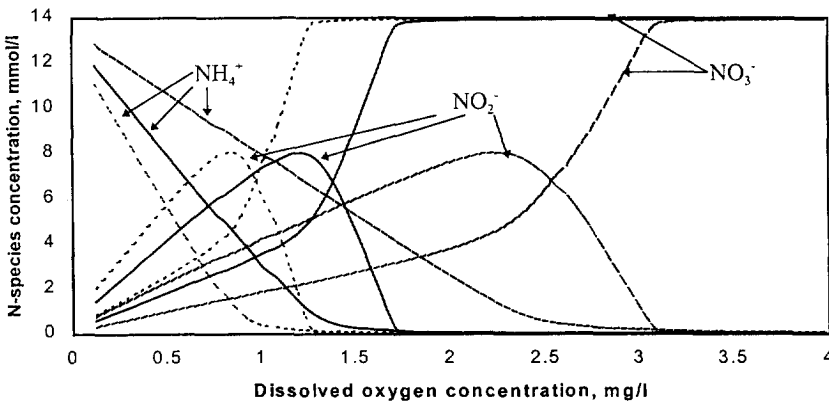




**Figure 3.** Calculated concentrations of N-species for different ammonia surface loading rates (influent concentration is 14 mmol N-NH<sub>4</sub><sup>+</sup>/l): 0.8 10<sup>-3</sup> \_\_, 1.6 10<sup>-3</sup> ---, 3.2 10<sup>-3</sup> ... kg N/m<sup>2</sup> biofilm day



**Figure 4.** Calculated concentrations of N-species for different medium pH: 6 ---, 7 \_\_, 8 ... pH



**Figure 5.** Calculated concentrations of N-species for different mass transfer coefficients: 2.5 ---, 5.0 \_\_, 7.5 ... m/day

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