0273-2289/03/109/0003/$20.00

Combining Mixing Regimes for Optimized Anaerobic Wastewater Treatment

Robbert Kleerebezem

Delft University of Technology,
Kluyver Laboratory of Biotechnology, Julianalaan 67, 2628 BC Delft,
The Netherlands, E-mail: r.kleerebezem@tnw.tudelft.nl

Abstract

Operational practice of high-rate anaerobic bioreactors such as upflow anaerobic sludge bed (UASB) reactors is generally based on maximization of the biomass concentration and, in the case of more than one reactor compartment, operation in parallel. In this article, a modeling approach is used to postulate that the treatment performance of anaerobic bioreactors can be improved by simple operational measures. To achieve minimized effluent soluble substrate concentrations, operation of two reactors in series combined with active exchange of biomass between both reactors is suggested. In this way, substrate concentrations lower than the minimum achievable concentration in a completely mixed reactor can be achieved. It is furthermore suggested that maximized biomass concentrations (and solid retention times [SRTs]) do not necessarily lead to minimized effluent concentrations of organic material. At elevated SRTs, the soluble microbial products resulting from biomass turnover are shown to represent the main fraction of soluble organic material in the effluent of the reactor, limiting treatment efficiency.

Index Entries: Anaerobic wastewater treatment; operation optimization; soluble microbial products; staged reactor concepts; upflow anaerobic sludge bed reactor.

Introduction

Although anaerobic treatment can typically reduce the chemical oxygen demand (COD) in wastewaters by 80–90%, effluent concentrations are normally too high to meet the discharge standards. Consequently, some form of posttreatment is required to enable discharge on surface waters. For waste streams that contain ammonia and phosphate in concentrations exceeding the discharge limits, nutrient removal also needs to be achieved in a posttreatment plant.

Posttreatment is in general disproportionally expensive compared to anaerobic pretreatment. Minimization of the effluent COD concentrations may simplify posttreatment methods or allow for more specific posttreatment methods aimed at removal of specific compounds. For example, for nitrogen removal from wastewaters rich in ammonium, the Sharon Anammox process may represent an interesting alternative for traditional nitrification-denitrification methods *(1)*. However, a strict requirement for implementation of the Sharon Anammox process is the absence of significant amounts of organic substrates in the wastewater owing to the extremely slow growth rate of the bacteria responsible for anaerobic ammonium oxidation. Consequently, the feasibility of this process strictly depends on the effluent qualities of the anaerobic pretreatment applied.

The effluent soluble COD material of anaerobic bioreactors consists of nonconverted substrate, intermediates of degradation (e.g., volatile fatty acids [VFA]), and soluble microbial products (SMP). It has been demonstrated that under normal operational conditions, SMP represents the bulk fraction of effluent soluble COD material *(2–5)*. Furthermore, experimental evidence is available that the concentration of SMP increases at increasing solid retention times (SRTs) *(3,4,6,7)*. Typically, upflow anaerobic sludge bed (UASB)-type reactors are operated without discharging biomass from the reactor compartment. Despite the absence of sludge discharge from the reactor, effluent solid concentrations are normally low owing to strongly reduced biomass yields as a result of the turnover of biomass and substrate consumption for maintenance purposes. Consequently, the SRT values are very high, and effluent SMP concentrations may represent a significant fraction of the influent COD concentration.

One may wonder if the described operational procedure is optimal for minimization of effluent soluble COD concentrations. In this article, a theoretical approach based on strongly simplified microbial kinetics is applied to study whether simple operational measures may decrease effluent soluble COD concentrations. First, minimization of effluent substrate concentrations is described and subsequently the influence of SMP formation and the impact of different types of microorganisms with strongly different kinetic characteristics are discussed.

Results

Minimum Substrate Concentration

The minimum substrate concentration that sustains microbial growth (S_{min}) corresponds to the minimum effluent concentration that can be achieved in a completely mixed reactor. The specific microbial growth rate according to Monod can be described as

$$\mu = Y_{X/S} \cdot q^{max} \frac{S}{K_S + S} - b \qquad (1)$$

in which q^{max} represents the maximum substrate uptake rate, S is the substrate concentration, X is the biomass concentration, K_S is the half-velocity

(Monod) constant for substrate uptake, $Y_{X/S}$ is the yield for biomass X on substrate S, and b is the decay rate constant. Solution of this equation for $\mu = 0$ gives the minimum substrate concentration that can be achieved in a completely mixed reactor:

$$S_{min} = K_S \cdot \frac{b}{Y_{X/S} \cdot q^{max} - b} \quad (2)$$

To achieve effluent substrate concentrations lower than S_{min}, a plug-flow pattern for substrate needs to be achieved in the biological reactor. A plug-flow pattern can be achieved either by discontinuous operation (sequencing batch reactor), or by placing completely mixed bioreactors in series.

Note that the actual model structure depends on the implementation of Eq. 1. The mass balance for active biomass (X) can either be coupled to a balance for inactivated biomass (X_i) or not. When coupled to an inactive biomass balance, $Y_{X/S}$ equals the total biomass yield since it expresses the formation of X as well as X_i per amount of S converted. This approach is typically used in models of environmental bioreactors with long biomass retention times such as Anaerobic Digestion Model No. 1 *(8)*. If no mass balance for inactive biomass is used and μ is positive, Eq. 1 suggests that b describes substrate uptake for growth-independent maintenance purposes ($b = Y_{X/S} \cdot m_S$, in which m_S is the specific substrate uptake rate for maintenance purposes). In this case, $Y_{X/S}$ represents the maximum active biomass (X) yield, and no inactive biomass is assumed to be formed in the system. This approach is typically used in biotechnological fermentations in which growth rates are high, and biomass inactivation can be neglected *(9)*.

The reaction compartment of high-rate anaerobic bioreactors such as the UASB reactor and derivatives thereof approach a completely mixed regime regarding the liquid phase. Because these types of reactors are biomass retention systems, effluent biomass concentrations are several orders of magnitude lower than the reactor concentrations. In fact, effluent biomass concentrations are typically very low as a result of biomass turnover in the reactor and substrate consumption for growth-independent maintenance purposes. If effluent biomass concentrations are negligible, operation of two or more UASB-type reactors in series will not enable the achievement of lower effluent substrate concentrations since Eq. 2 remains valid for the individual reactors. Consequently, to enable effluent substrate concentrations lower than S_{min} in a continuously operated anaerobic system, operation in series should be combined with the exchange of biomass between the individual reactor units, as discussed next.

Effluent Substrate Concentrations Lower Than S_{min} in Two-Stage Reactors

To predict the theoretical steady-state effluent substrate concentrations for the reactor setups RS1–RS3 shown in Fig. 1 a mathematical model

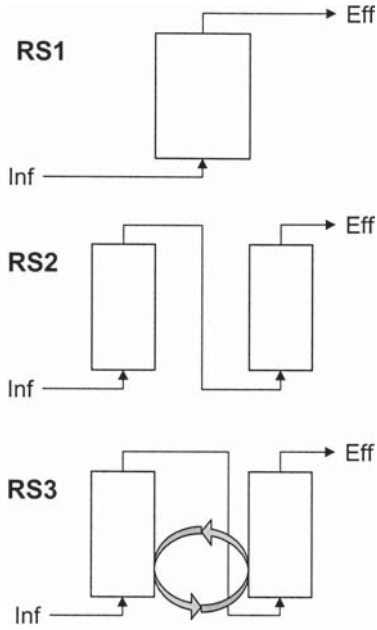

Fig. 1. One-stage reactor (RS1), two-stage reactor (RS2), and two-stage reactor setup with biomass exchange between the two reactors (RS3).

was developed. Microbial kinetics were based on the equations for substrate uptake and microbial growth as shown in Eq. 1. The description of the biomass dynamics in the reactors was based on the assumption that the total biomass concentration (X_{tot}) in the reactor is constant. This rather blunt assumption is based on the practical observation that the biomass concentrations in well-functioning UASB-type reactors appear to be limited by a physical maximum. The value of 30 g VS-COD/L corresponds to approx 20% of the total reactor volume. Taking into account both gas holdup and the volume of the three-phase separators in the UASB-type reactor, the actual concentration of wet biomass in the reaction compartment is approx 25–35%. Much higher biomass concentrations are unlikely to occur.

Biomass is assumed to consist of both active (X) and inactivated (X_i) biomass. For the two-stage systems, biomass washout from the first reactor and subsequent introduction in the second reactor is neglected. Minimized effluent concentrations of biomass are assumed to be achieved by sludge discharge from the reactor compartment. As an example, the mass balance for active biomass in RS3 is as follows:

$$\frac{dX_2}{dt} = \frac{-X_2}{SRT} + \frac{Q_e}{V_{R2}}(X_1 - X_2) + \left(Y_{X/S} \cdot q^{max} \cdot \frac{S_2}{K_S + S_2} - b\right) \cdot X_2 \qquad (3)$$

in which the first term corresponds to the influent-effluent term, the second term describes biomass exchange between both reactors (with flow rate Q_e,

Table 1
Operational and Kinetic Parameter Values

Parameter	Unit	Value
q^{max}	g S-COD g X-COD^{-1}d^{-1}	4
$Y_{X/S}$	g X-COD g S-COD^{-1}	0.1
b	d^{-1}	0.05
K_S	g S-COD/L	0.2
X_{tot}	g X-COD/L	30
Q_l	L/d	Var.
Q_e	L/d	0.2
$V_R(V_{R1} + V_{R2})$	L	2
S_{inf}	g S-COD/L	5

q^{max}, maximum specific uptake rate; $Y_{X/S}$, biomass yield on substrates; b, biomass decay rate constant; K_S, half velocity constant for substrate uptake; X_{tot}, total biomass concentration; Q_l, influent liquid flowrate; Q_e, liquid flow rate for biomass exchange between R1 and R2 in RS3 (Fig. 1); V_R, total reactor volume; S_{inf}, substrate concentration in the influent.

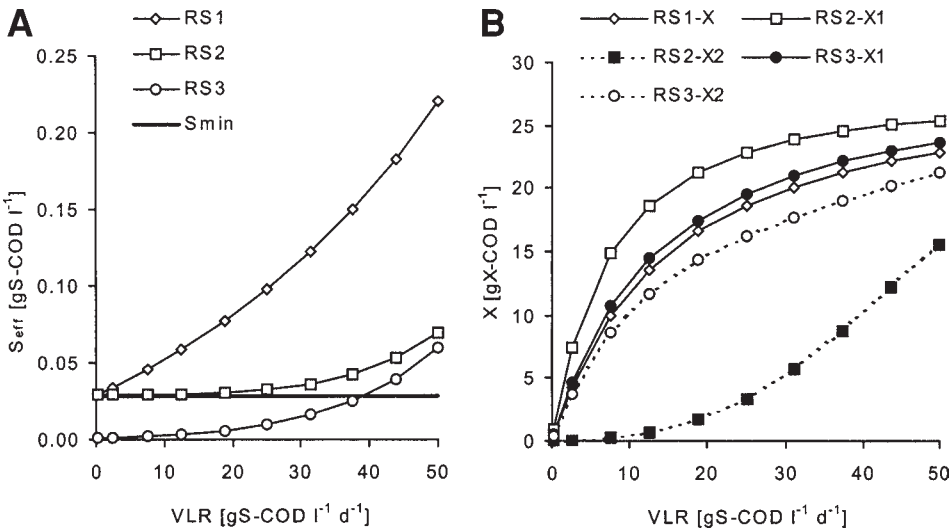

Fig. 2. Effluent substrate concentrations **(A)** and concentration of active biomass **(B)** as function of volumetric loading in RS1, RS2, and RS3. X1 and X2 stand for the active biomass concentrations in the first and second stages of RS2 and RS3, respectively.

RS3), and the third term describes microbial growth. Subscripts 1 and 2 stand for the first and second reactor in RS3. RS2 is obtained by setting Q_e to zero. Parameter values and operational parameters used for the calculations are given in Table 1. These values represent typical values but were primarily used to enable the generation of numerical output.

Figure 2 shows the steady-state effluent dissolved COD concentrations and the active biomass concentrations in RS1–RS3 as a function of the

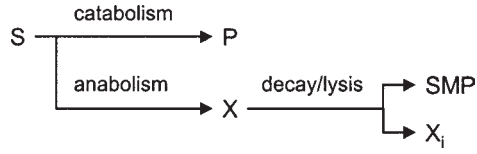

Fig. 3. Schematic representation of fate of substrates in anaerobic bioreactors. Carbon atoms originating from substrate are proposed to end up in product(s) (*P*), biomass (*X*), nonbiodegradable soluble products (*SMP*), or inert solids (X_i).

volumetric loading rate (VLR). The results show that in RS1 and RS2 the minimum effluent substrate concentration at low loading rates is restricted by S_{min}, as can be calculated using Eq. 2: 0.029 g S-COD/L. As proposed, the effluent substrate concentrations for RS3 are significantly lower than for RS1 and RS2, mainly at lower loading rates. At higher loading rates, biomass in both reactors from RS2 and RS3 consists mainly of active biomass, resulting in a decrease in the difference between RS2 and RS3. The lower effluent substrate concentrations for RS2 compared with RS1 are explained by the much higher SRT values in the second stage of RS2.

Soluble Microbial Products

As described in the Introduction section, the soluble organic material in the effluent of high-rate anaerobic bioreactors typically consists primarily of SMP. The concentration of SMP in glucose-fed chemostat reactors may amount up to 10% of the influent COD concentration *(2,10)*. It has furthermore been demonstrated that the SMP concentration increases at increasing biomass concentrations *(2,4,11–13)*.

Recent modeling efforts to describe SMP formation have been based on the division of SMP into two categories: biomass decay–associated products (BAP), and substrate uptake–associated products (UAP). Rittmann et al. *(7)* have proposed a complex method for modeling of both UAP and BAP production and consumption. This method was later used in other studies to describe the formation and degradation of SMP in anaerobic bioreactors *(4,11)*. Herein a simplified approach is used based only on BAP formation from active biomass. The approach assumes that decayed biomass is susceptible to cell lysis resulting in the formation of inert solids and nonbiodegradable SMP (Fig. 3). These simplifications compared to previous modeling efforts are justified by the following observations: First, UAP is rapidly formed and degraded in transient-state experiments but represents an insignificant fraction of the effluent concentration of SMP in steady state *(4,11)*. This suggests that UAP actually is an unidentified intermediate in the complex substrate degradation studied by Kuo et al. and Barker and Stuckey *(4,11)*. Second, BAP degradation rates are small compared with formation rates *(4,11)*. In fact, Barker and Stuckey *(11)* obtained an optimized system description when BAP degradation rates were assumed zero.

A 20% fraction of decayed biomass is assumed to end up in SMP, whereas all other parameters and operational variables are the same as in

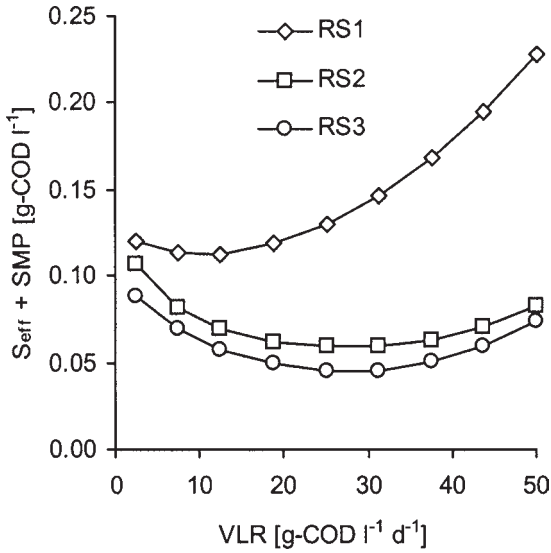

Fig. 4. Effluent soluble COD concentration as function of VLR, taking into account formation of SMP.

the previous paragraph. Herewith the specific SMP production rate equals 0.01 g SMP-COD g X-COD^{-1}d^{-1}, which is a low value compared with the value of 0.05 suggested by Kuo et al. *(4)*. The system of equations was solved and the total soluble COD concentration ($S + BAP$) in the effluent as a function of the volumetric loading rate is shown in Fig. 4. The results demonstrate that a volumetric loading rate exists in which the soluble COD concentration is minimized. This is due to the occurrence of two influences on the effluent COD concentration in opposite direction: At increasing VLR values the effluent substrate concentrations will increase due to limitations in SRT, and at decreasing VLR values the high SRT values result in the formation of increasing concentrations of SMP. Note that the different reactor configurations (RS1–RS3) have a limited impact on SMP production, and differences are mainly determined by differences in the effluent substrate concentration (Fig. 2).

Complex Substrate Degradation

In methanogenic environments, complex substrates are degraded in a network of different bacteria and archea. Kinetic and stoichiometric parameters of the different bacteria involved in these networks can be strongly different, limiting the usefulness of the described single-substrate approach.

A two-species model was developed according to the scheme shown in Fig. 5. In this scheme, a substrate S is fermented by X_a to an intermediate product IP that is converted by X_m to the end product P. For both organisms, the same kinetic parameters were used as described before, except for the biomass yield values that were assumed to amount to 0.069 g of Xa-COD

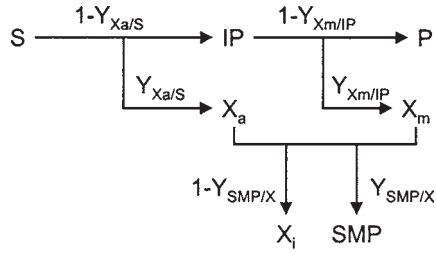

Fig. 5. Schematic representation of degradation of substrate in two-organism mechanism.

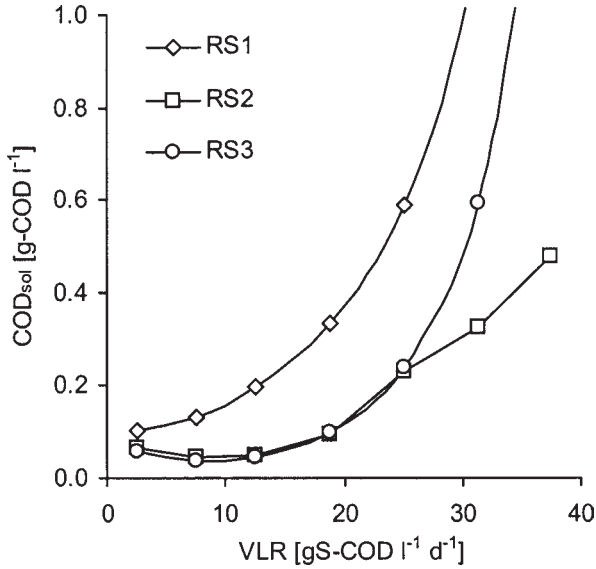

Fig. 6. Effluent soluble COD concentration as function of VLR in RS1–RS3 as calculated with two-species model.

g of S-COD^{-1}, and 0.030 g of Xm-COD g of IP-COD^{-1}. Herewith the total biomass yield of the mixed culture remains the same as in the previous examples.

Steady-state simulation results for the effluent soluble COD concentration as a function of the VLR are shown in Fig. 6. It is evident that both of the two-reactor setups (RS2 and RS3) are strongly superior to the single-reactor approach (RS1). Even though the effluent soluble COD concentrations for RS2 and RS3 are comparable at loading rates less than 25 g of S-COD/(L · d), the distribution of biomass is completely different. Because of the recirculation of biomass in RS3, the concentrations X_a and X_m are comparable in both reactors. In RS2 a clear spatial separation is achieved of X_a and X_m in the first and second reactor, respectively. The improved performance of RS2 compared with RS1 is therefore the result of biomass separation, whereas for RS3 this is the direct result of the plug-flow pattern for

the substrate, combined with complete mixing for biomass. At loading rates exceeding 25 gS-COD/(L · d) full spatial separation of the biomass is achieved in RS2. This explains the lower effluent concentrations at elevated loading when compared to RS3. In RS3 the fraction of X_m strongly decreases.

Discussion

Anaerobic UASB-type reactors and derivatives thereof are typically operated as completely mixed reactors, and at maximized SRTs and biomass concentrations. The work described herein suggests that both approaches can be improved.

Optimized treatment as reflected in minimized effluent (soluble) COD concentrations is achieved in a system that achieves a plug-flow regime regarding the liquid, and a completely mixed regime for the biomass. This intuitively contradicting objective can be achieved relatively easily in UASB-type reactors. A simple approach that enables the required regime is shown as RS3 in Fig. 1. This approach includes two UASB-type reactors that are operated in series, whereas concentrated biomass is exchanged between the reaction compartments of the first and second stage. A basically comparable approach, based on cyclic operation of two anaerobic filter reactors, has been suggested by Young and colleagues *(14,15)*.

The presence of biomass in a granular form in UASB-type reactors may accentuate the advantages of RS3 compared with RS1 in Fig. 1. In completely mixed bioreactors, the substrate concentration is kept as low as possible to minimize effluent concentrations. At these low substrate concentrations, the methanogenic biofilm will only partly be penetrated, and, consequently, only part of the catalytic capacity of the granule is used. In RS3 the substrate concentration in the first stage will be (relatively) high, and biofilms will be fully penetrated, resulting in highly active granular biomass. On introduction in the second stage, the granular biomass will be able to effectively remove COD material left over by the first stage.

Only in the case of strongly different biomass yields of organisms involved in the degradation of complex substrates (e.g., methanogenic glucose degradation) does a complete spatial separation of the different microbial groups appear to be beneficial (RS2). Operation in series without biomass exchange between both stages may also enhance the startup of bioreactors treating mixtures of rapidly and slowly degradable substrates owing to maximization of the SRT in the second stage *(16)*.

Until now optimization efforts for anaerobic treatment have been aimed at minimizing the effluent concentration of substrate and intermediates (VFA) while maximizing the treatment capacity of the systems. These treatment objectives are met when the biomass concentrations and SRT values are maximized. The models presented herein suggest that maximization of the biomass concentration may, however, lead to significant accumulation of SMP in the reactor effluent. The treatment objective could therefore be reformulated to minimization of the effluent soluble COD

concentration by operation at SRT values that maximize substrate removal while minimizing SMP production. Evidently, the optimal operational strategy depends on the specific treatment objective. If anaerobic treatment is aimed at combined, minimized sludge production and minimized effluent soluble COD concentrations, the optimized operational strategy may consist of maximization of biomass turnover in the reactor by maximization of the SRT. In this case, the accumulation of SMP (relatively low concentrations) is taken for granted.

The models described herein represent strong simplifications of the practice of anaerobic wastewater treatment. Wastewaters consist of mixtures of substrates, and in methanogenic environments these substrates are degraded in a complex network of bacteria and archea. The dynamics of the concentrations of the individual organisms are the result of substrate conversion for growth and maintenance purposes, bacterial decay, cell lysis, and subsequent formation (and degradation) of SMP and inert solids. Only very limited quantitative information is available on these latter subjects, even though they play an important role in the actual active biomass concentrations that can be achieved in anaerobic bioreactors and, consequently, the treatment performance of anaerobic bioreactors.

In general, a treatment performance of about 80% COD removal in anaerobic bioreactors treating wastewaters containing readily biodegradable substrates is considered an acceptable value. One may wonder on what the general acceptance of these relatively low treatment efficiencies is based. Evidently, unfavorable kinetics or limitations in biomass retention generally cannot explain the limited treatment efficiency. It is suggested that significant improvements in treatment performance can be achieved by introduction of a plug-flow pattern combined with an adequate sludge management strategy. This may enable the application of more specific posttreatment methods or, in specific cases, may exclude the necessity for posttreatment of the effluent of anaerobic bioreactors.

Acknowledgment

I gratefully acknowledge Biothane Systems International for financial support for attending the 7th FAO/SREN workshop on anaerobic digestion in Moscow, Russia, May 19–22, 2002.

References

1. Dongen, L. G. J. M. V., Jetten, M. S. M., and Loosdrecht, M. C. M. (2001), *The Combined Sharon/Anammox Process, a Sustainable Method for N-Removal from Sludge Waters, Water and Wastewater Practitioner Series*, IWA, London, p. 64.
2. Barker, D. J. and Stuckey, D. C. (1999), *Water Res.* **33(14),** 3063–3082.
3. Kuo, W. C. and Parkin, G. F. (1996), *Water Res.* **30(4),** 915–922.
4. Kuo, W. C., Sneve, M. A., and Parkin, G. F. (1996), *Water Environ. Res.* **68(3),** 279–285.
5. Barker, D. J., Mannucchi, G. A., Salvi, S. M. L., and Stuckey, D. C. (1999), *Water Res.* **33(11),** 2499–2510.
6. Pribyl, M., Tucek, F., Wilderer, P. A., and Wanner, J. (1997), *Water Sci. Technol.* **35(1),** 27–34.

7. Rittmann, B. E., Bae, W., Namkung, E., and Lu, C. J. (1987), *Water Sci. Technol.* **19(5),** 517–525.
8. Batstone, D., Keller, J., Angledaki, I., Kalyuzhnyi, S. V., Pavlostathis, S. G., Rozzi, A., Sanders, W. T. M., Siegrist, H., and Vavilin, V. A. (2002), Scientific and Technical Report No. 13, IWA, London, p. 80.
9. Heijnen, J. J. (1999), in *Encyclopedia of Bioprocess Technology: Fermentation, Biocatalysis, and Bioseparation*, Flickinger, M. C. and Drew, S. W., eds., John Wiley & Sons, pp. 267–291.
10. Aquino, S. F. and Stuckey, D. C. (2001), in *Anaerobic Digestion 2001, Anaerobic Conversion for Sustainability*, Van Velzen, L. and Verstraete, W., eds., Technologisch Inistituut VZW, Antwerp, Belgium, pp. 109–116.
11. Barker, D. J. and Stuckey, D. C. (2001), *Water Environ. Res.* **73(2),** 173–178.
12. Noguera, D. R., Araki, N., and Rittmann, B. E. (1994), *Biotechnol. Bioeng.* **44(9),** 1040–1047.
13. Fang, H. H. P. and Jia, X. S. (1998), *Biores. Technol.* **66,** 235–239.
14. Howerton, D. E. and Young, J. C. (1987), *J. WPCF* 59(8), 788–794.
15. Young, J. C. (1990), US patent no. 4,975,196.
16. Kleerebezem, R. (1999), in *Anaerobic Treatment of Phthalate Isomers: Microbiological and Technological Aspects*, Wageningen University, Wageningen, The Netherlands, p. 184.