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Work Report to obtain the credits of investigation in the Program of

Doctorate in Industrial Engineering

First approach in Automatic Control for sludge treatment systems. A case of study: ATAD technology

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Content

1. Introduction (Background)

In general, Wastewater Treatment Plants (WWTPs) are complex systems subject to large perturbations in wastewater flow rate, load and composition. Within WWTP, the sludge is a by-product derived from the water treatment line and must be treated before its final disposal. And, as in the water line, the sludge final disposal has to meet strict regulations.

Traditionally, conventional anaerobic digesters have been used as sludge treatments. However, in the last years, there are new technologies that have been developed, like the Autothermal Thermophilic Aerobic Digesters (ATAD). The ATAD is a promising alternative to conventional digestion systems; the performance of these reactors depends on the applied aeration.

Many control strategies have been proposed in the literature for an improved and more efficient operation of wastewater treatment plants, all of them have been concentrated in the water treatment line, but less effort has been done to enhance the operation of the sludge treatment line. In this respect, standard evaluations and extended studies have been developed to improve the operation and evaluation of controllers for water treatment line. But at present, the realisation of automatic control in ATADs is just begun, and evidently, a standard evaluation doesn't exist for sludge treatment line.

Moreover the relatively recent technology of ATADs, the lack of robust sensors also makes the control of these processes difficult, only ORP and temperature sensors can operate in full-scale plants. Based in this problematic, to start possible control strategies to improve the ATADs performance, it is necessary to develop a standard evaluation for sludge treatment systems. Such an evaluation needs to be based on a rigorous methodology including a simulation model, plant layout, controllers, sensors, performance criteria and test procedures, i.e. a complete benchmarking protocol.

A brief description of the ATAD technology and a definition of a benchmark and its characteristics will be presented as follows.

The ATAD Technology

The ATAD (Autothermal Thermophilic Aerobic Digester) process was introduced some 30 years ago as a means of retrofitting conventional digesters in order to deal with more stringent regulations for the disposal of municipal wastewater sludge. The operating principle of ATAD process relies on conserving the heat energy released during the aerobic biodegradation of the organics substances contained in the sludge so as to reach and maintain thermophilic temperatures (50 - 70º C). Amongst other advantages, working in the thermophilic range means that: (1) sludge stabilization can be achieved in shorter detention time (SRT = 6 - 12 days),

resulting in smaller volumes; and (2) sludge pasteurization is also feasible, due to high pathogen destruction efficiencies. To achieve sludge pasteurization, ATAD are operated in batch-mode (sewage feeding; digestion period; sludge withdrawal) to avoid the mixture of the treated pasteurized sludge with the incoming sludge.

The operation of full-scale ATAD facilities in different countries has confirmed the suitability of this technology to provide a final product which meets all the standards for unrestricted application and reuse of municipal sludge (USEPA, 1990; Layden et al., 2007).

Aeration in all aerobic biological systems is one of the most important considerations since it affects both the quality of the effluent and the total operating costs. The ATAD system is no an exception, the influence of aeration being even more marked. Over-aeration increases costs without leading to a significantly better quality of treated sludge. Moreover, in the case of air-based aerating systems, over-aeration involves a cooling effect on the slurry with the consequent risk of pasteurization temperatures not being reached. On the other hand, under-aeration limits the efficiency for stabilization and heat generation. Also, under-aeration promotes anaerobic conditions in the reactor, which increases the potential for undesired odours in the outlet off-gas (Staton et al., 2001).

Automatic control for ATAD technology is limited to the use of very elemental strategies for aeration. In fact, the first ATAD generation did not consider any capability for the regulation of the aeration system and it was not until the advent of the second ATAD generation that these systems began to be equipped with more sophisticated devices where the flow-rate of the injected gas stream could be automatically manipulated (Scisson, 2003). Even so, the lack of robust and reliable online sensors is still an important drawback that has hindered the deployment of monitoring and control tools for these systems. Industrial sensors for dissolved oxygen or suspended solids, for example, are intended for use in the secondary treatment of water, where temperatures and solid concentrations are not excessive. However, these sensors are not ready to work under thermophilic temperatures and very high solid concentrations.

At present, only Oxidation-Reduction Potential (ORP) and temperature sensors fulfil the technical features required for operation in full-scale ATADs. This is the reason why aeration control in ATADs has so far been based on either ORP or temperature sensors. Staton et al. (2001) stated that with an appropriate processing of the ORP signal, the depletion of biodegradable organic substrate in the digester can be automatically detected (end of aerobic oxidation transformations); thus, on the basis of this observation, control strategies for external aeration can be implemented to save aeration costs. Breider and Drnevich (1981) probably hold the first patent on real-time control of ATADs. They propose a very simple strategy that automatically modifies the inlet gas flow-rate as a function of the sludge temperature, with the objective of maintaining it within a predetermined range.

Simulation Benchmarks for Control of WWTPs

In the environmental engineering, simulation benchmarks have been used for evaluating and quantitative comparing of activated wastewater treatment control strategies. Thus, it is established that a simulation benchmark must be independent of any commercial simulation software employed (GPS-X, WEST, Matlab/Simulink, BioWin, etc.). Moreover, it is not just a model for simulations, but a complete protocol that includes all the steps for evaluation of controllers.

The Benchmark Simulation Model nº.1 (BSM1) was the first benchmark for WWTP (Copp, 2002). The BSM1 was designed to evaluate control strategies for wastewater treatment with nitrogen removal. This protocol consists of a plant-layout comprised in five reactors in series with a secondary setting tank. The evaluation protocol consists of a prior controller initialization followed by an evaluation period of 7 days. In this respect, an important point for controller's evaluation, besides the quantifying of control responses (errors, deviations, etc.), is the index process definition, from which it is possible to quantify: (i) effluent quality, (ii) operating costs and (iii) sludge production.

(i) The effluent quality index is referred to the quantification of the effluent pollution load, expressed in kg/d.
(ii) The operational costs take into account both the required energy for pumping operations and the aeration, expressed in kWh/d.
(iii) The sludge production is quantified in two terms: sludge outflow and the total sludge production.

Later on, advances and modifications have been developed in benchmarks, such as the work presented by Vanrolleghem and Gillot (2002) who proposed to group the BSM1 indexes in two general indexes: (1) TCI (Total Cost Index), it summarizes the effluent quality costs and the energy costs; and (2) RI (Robustness Index), it evaluates the sensitivity of TCI between a range of variation for some selected process parameters (influent flow-rate, COD and N load, temperature, etc.). Samuelsson et al. (2007) and Stare et al. (2007) demonstrate how depending on the defined function costs related to the addition of external carbon and the effluent nitrate requirements, the optimal operation criteria for internal recirculation and for addition of methanol can fluctuate significantly.

Another propose to extend the BSM1 includes controllers evaluation indexes in which their values are obtained from expert reasoning modules included in the benchmark (Comas et al., 2006). The idea is to quantify the amount of risk that these control strategies could offer, for instance, foams, bulking, increasing sludge in secondary settlers, etc.

2. Objectives

In this work, the ATAD technology has been selected as the unitary process for sludge digestion. Considering the innovation of this kind of processes nowadays and the lack of control studies that have been developed in these reactors, this work presents two main objectives: (1) the definition of a benchmarking simulation protocol for evaluation of controllers in ATAD systems operated as single treatments, and (2) the introduction of new control strategies for ATAD aeration. Indeed, the development of this work allow us to achieve these two objectives that can complement to each other; because the benchmark presented in this work is designed as a tool to evaluate the control strategies proposed.

These two main objectives have been developed through the following specific objectives:

- The upgrading of the mathematical model for the ATAD reactor.
- The programming of an appropriate simulation tool that allows modelling and control tests in a simple and flexible way.
- The design of an appropriate scenario of simulations that allows to explore different kinds of control strategies.

These specific objectives as well as a discussion of results in terms of the benchmark proposed will be presented in this work.

3. Materials & Methodology

The mathematical model

The ATAD modelling has been improved in the last few years. Gómez et al. (2007) bring together existing formulations on biochemical reactions, physical and chemical transformations and thermal energy balances, in order to develop a comprehensive model for the ATAD which includes dynamic prediction of liquid and gas compounds as well as temperature.

The ATAD reactor has been modelled as a completely-stirred tank; but, in this case, biological and heat effects have also been added. The biochemical model is based on the standard ASM1 with slight modifications to make it consistent with observations from thermophilic aerobic digesters.

The mass balance describes changes in the liquid and gaseous components within a reactor; for a generic model component X_i, the basic mass balance equation is:

$$\frac{d(V_k X_i)}{dt} = \sum_j v_{ij} \rho_j V_k + Input(X_i) - Output(X_i) \quad (1)$$

, where v_{ij} is the stoichiometric coefficient for the component X_i in the transformation *j;* ρ_j is the kinetic expression of the transformation *j* and V_k is the volume of the phase in which the component is defined. *Input*(X_i) and *Output*(X_i) are the mass transport terms associated with the input and output streams. The representation of this balance is drawn in **Figure 1**, where the reactor is modelled as two volumes that interact to each other, one represents the soluble-particulate components and the other one represents the gaseous components.

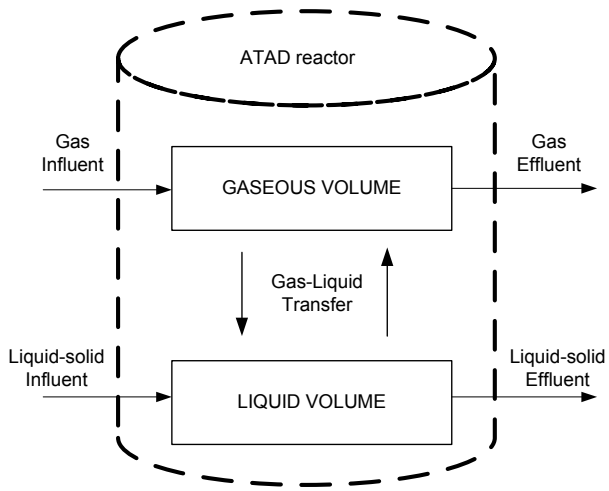

Figure 1. Mass balance in the digester

The thermal energy balance allows predicting the temperature variation both in the liquid and gaseous phase of the reactor, by means of the following equation:

$$\Delta H = \Sigma\ (Heat\ generation) - \Sigma\ (Heat\ loss) \quad (2)$$

In the specific case of ATAD process, there are diverse elements that have been included to achieve a good prediction of the sludge temperature such as: heat added by the biological activity; heat added by the liquid and gas influent; heat lost by the liquid and gas effluent; heat transfer between volumes by convection; heat lost in the liquid volume due to water evaporation; heat lost through the reactor walls by conduction; heat added by the mixing and pumping systems. In this respect, two main effects that influence the temperature dynamics are:

- The heat added by the biological activity. It is produced during the aerobic oxidation of organic matter by the heterotrophic bacteria.
- The heat lost by the water evaporation.

All the factors that take part in the energy balance of the ATAD reactor are represented in the **Figure 2**.

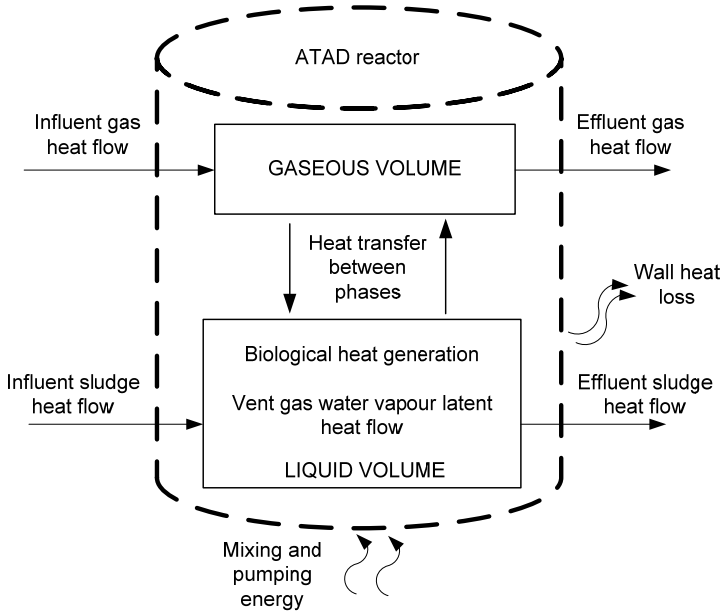

Figure 2. Thermal energy balance in the digester

Previous model of ATAD reactors have incorporated evaporation and convection equations taking the oxygen transfer coefficient (*Kla*) as the manipulated variable. In this work, evaporation, convection and *Kla* equations have been modelled to be dependent on the air flow-rate variation.

Additionally, previous works related to dynamic prediction of temperatures in biological tanks (Vismara, 1985; Messenger et al., 1990) form the basis of the heat model. A detailed definition of the ATAD model used in the AT_BSM can be found in Gómez et al. (2007).

The Simulator

The Matlab/Simulink® platform has been selected to program and simulate the ATAD model. Matlab/Simulink is an appropriate platform oriented for numerical resolution of equations (in our case are differential equations), and is also suitable for control studies. In this respect, the S-functions C-Mex type included in Simulink have been selected as the programming tools within Simulink because, besides the possibility of programming complex functions within blocks structures, this platform offers low simulation times compared with same programs developed in m-files, and with models implemented in commercial tools like WEST.

Study of specific Benchmarks

Mathematical modelling and simulation have become important tools for supporting not only the design and operation of WWTPs but also the analysis and synthesis of automatic controllers. Particularly, the original BSM1, and then the Long-Term BSM1 (BSM1-LT) and the BSM nº 2 (BSM2) are three important simulation benchmarks aimed at providing researchers with standard methodologies for the objective comparison of control strategies (Coop, 2002). However, ATAD technology is not supported by any of the existing benchmarks and hence, and for this reason, nowadays studies on ATAD control cannot be undertaken with any of these protocols.

For these reasons, a new benchmark definition for sludge treatment by ATAD technology will be presented in the next chapter. As a starting point, this benchmark definition takes the advantages and procedures of the benchmarks mentioned above.

Preliminary simulation studies in ATAD

Several simulations done with the ATAD have established a correlation between the temperature trajectory during a digestion cycle with the lack of the biodegradable substrate (Gómez et al., 2007). This correlation is demonstrated with the appearance of a “knee” or “bend-point” in the temperature trajectory (the pass from the hydrolysis to lysis), which is an indirect indicator that the sludge is already stabilised. **Figure 3** shows an example of the typical temperature trajectories that take place in the ATAD when it is operated at both influent under-loads and constant air-flow rates. It is observed that the temperature trajectory has a bend-point during the reaction phase, which is related to the depletion of biodegradable substrate in the reactor (“bCOD knee”).

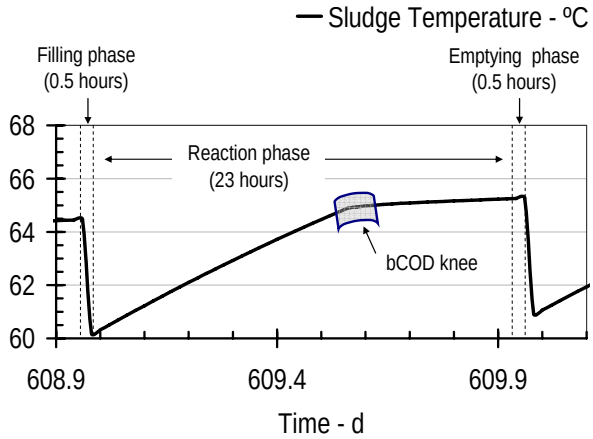

Figure 3. Temperature profile for an over-aerated batch

The control strategies proposed in this work are based on the effects observed in the sludge temperature profile by means of the air flow-rate applied.

The reasons of the performance index

The performance index, as a whole, is a set of independent indicators that synthesize all the output data into a small number of composite terms. These composite terms include, among other, a general effluent quality measurement, energy terms for pumping and aeration, and a measurement of withdrawal production.

These criteria aim at condensing the simulation output to a few indices or key variables that can be said to represent the system, controller or monitoring performance to allow for easy comparison of results. The criteria definition for control system performance and the equations needed to formulate these terms are part of the new benchmark, and will be presented later on.

4. Results & Discussions

Benchmark configuration

As referred previously, based on standard BSM protocols, an *ad hoc* benchmark for the ATAD process as unique treatment has been specifically defined (henceforth referred as: "AT_BSM"). Implementation of the AT_BSM protocol involves four major definitions: (i) Influent definition; (ii) Plant layout; (iii) Evaluation criteria; and (iv) Simulation procedure.

(i) Influent definition

To compile an historic data file with enough information (flow and composition) of typical sludge characteristics in a full-scale plant is a very difficult task. In this case it was decided to automatically generate the file by modelling a representative plant. Therefore, the virtual plant of BSM2 was chosen and simulated according to the BSM2 simulation procedure (Vrecko et al., 2006). Simulation results were saved every 15 minutes so as to create an influent file with the characteristics of sludge (both, primary and secondary sludge) for a 728-day period of plant performance. **Table 1** summarises some major features of the sludge obtained.

Table 1. Influent file in the *AT_BSM*

Sludge features	Min	Max	Average
Flow rate [m3/d]	39 / 17 / 56	544 / 102 / 646	148 / 40 / 188
COD [g/L]	2.2 / 85 / 14	74/ 95 / 77	35 / 86 / 47
TSS [g/L]	1.5 / 60 / 10.3	49 / 67 / 52	24 / 61 / 32
Temperature [ºC]	9.5	20.5	15

** Primary/Secondary/Mixed*

(ii) Plant layout

In our case of study to obtain sludge pasteurization the ATAD process has to operate in batch-mode, since it is advisable to avoid hydraulic short-circuits during digestion. For this reason, within the benchmark, a sequencing operation has been implemented for the ATAD operation. In particular, in compliance with EU requirements for sludge disinfection in batch processes, the following 24 hours cyclic sequence has been adopted by default: 0.5 hours for sewage feeding; 23 hours for reaction (aerated reaction phase); and finally 0.5 hours for sludge withdrawal. The non-continuous feeding in the ATAD makes the utilisation of pre-holding tank necessary. As a result, the proposed plant layout in the AT_BSM consists of a pre-holding tank and an ATAD single reactor (**Figure 4**). An analysis of the obtained data file influent described before was performed to obtain appropriate values for the effective volume of both the pre-holding tank (V_{max} = 2000 m^3) and the digester (V_{max} = 2400 m^3).

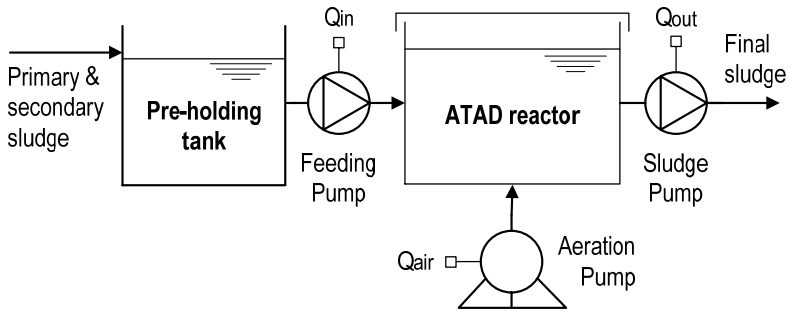

Figure 4. Plant-layout in the AT_BSM

In relation to the plant-model, the pre-holding tank has been modelled as a completely-stirred variable volume basin where only mass transport has been considered (neither biological reactions nor heat transformations have been included).

(iii) Evaluation criteria

In the same way like in standards benchmarks protocols, the AT_BSM has included three major performance indices in order to evaluate and compare control strategies: (1) the Operational Cost Index (OCI); (2) the Pasteurization Quality Index (PQI); and (3) the Stabilisation Quality Index (StQI). The OCI takes into account all the energy costs involved in the operation of the ATAD reactor and has been calculated in a similar way to that undertaken in the BSM2, but using a non-weighted sum:

$$OCI(kWh) = AE + PE + ME \quad (3)$$

, where the terms AE represents the energy for external aeration, PE is the pumping energy and ME refers to the mixing energy. PE covers the feeding of the raw sludge into the ATAD as well as the withdrawal of the treated sludge from the ATAD. For the mixing energy, only the energy required for mixing the ATAD has been considered.

PQI and StQI have been introduced to quantify the degree of disinfection and stabilisation of the sludge leaving the ATAD, respectively. Due to the lack of consensus, on the definition of general criteria for sludge pasteurization, the EU recommendation applicable to batch digesters has been adopted in this work (European Commission, 2000). It recommends: “thermophilic aerobic digestion at temperature above 55º C during batch time of 20 hours, without admixture or withdrawal during the treatment”. Accordingly, PQI (%) has been incorporate into the benchmark to represent the percentage of ATAD cycles in which the sludge leaving the treatment complies with the above definition. Since in the more general case the decant volume per cycle (V_{out}) might change from cycle to cycle, PQI has been formulated in terms of mass fluxes per cycle, as follows:

$$PQI(\%) = \frac{\sum_{i=1}^{N}\left[k_{paste.}^{(i)} \cdot V_{out}^{(i)} \cdot TSS_{out}^{(i)}\right]}{\sum_{i=1}^{N}\left[V_{out}^{(i)} \cdot TSS_{out}^{(i)}\right]} \cdot 100 \text{, where: } \quad k_{paste.}^{(i)} = \begin{cases} 0; & if \;\; PTime^{(i)} < 20\,hrs \\ 1; & if \;\; PTime^{(i)} > 20\,hrs \end{cases} \qquad (4)$$

, where N represents the total number of batch cycles, *i* is the *i*-th batch, and TSS_{out} is the total suspended solids concentration in the existing sludge. $PTime^{(i)}$ represents the total time in which the sludge has been at temperature above 55º C during the aerated reaction phase of the *i*-th batch.

A review of existing policies and guidelines for sludge management in different countries shows the diversity of criteria used to specify the requirements for sludge stabilisation. In fact, such requirements are strongly conditioned by the type of treatment used for digestion. For instance, depending on the degree that aerobic digestion is disturbed, the U.S. EPA regulation 40 CFR Part 503 (USEPA, 1993) establishes three options for compliance with the vector attraction reduction requirements (i.e. sludge stabilisation):

- Option 1: “At least 38% reduction in volatile solids during sewage”.
- Option 2: “Less than 15% additional volatile solids reduction during bench-scale aerobic batch digestion for 30 additional days at 20º C”.
- Option 3: “Specific Oxygen Uptake Rate (SOUR) at 20º C less than 1.5 mg O2/hr/g total sewage sludge solids”.

Option 3 is only applicable to mesophilic aerobic digesters; Option 2 is only valid for aerobically designed sewage sludge with 2% or less solids. Unlike Option 2 and 3, Option 1 is not restricted to any specific treatment technology; however, this option has certain limitation since it is not completely appropriate for treatment where the incoming sludge has been partially pre-stabilised (for example, sewage from secondary treatments operated at medium/large SRT). In these situations, Option 2 should be used instead. Since Option 1 and 2 are valid for aerobic thermophilic digestion, a combination of both has been adopted to formulate StQI:

$$StQI(\%) = \frac{\sum_{i=1}^{N}\left[k_{st}^{(i)} \cdot V_{out}^{(i)} \cdot VS_{out}^{(i)}\right]}{\sum_{i=1}^{N}\left[V_{out}^{(i)} \cdot VS_{out}^{(i)}\right]} \cdot 100 \text{, where } \quad k_{st}^{(i)} = \begin{cases} 1 & if \left\{\begin{matrix} Option\ 1^{(i)} \\ or \\ Option\ 2^{(i)} \end{matrix}\right\} is\ met \\ 0 & \text{Otherwise} \end{cases} \qquad (5)$$

Option $1^{(i)}$ and Option $2^{(i)}$ are Boolean variables, whose values result from the evaluation at the end of the i-th batch of the Option 1 and 2 statements referred above; VS_{out} represents the

volatile solids of the effluent sludge. Finally, three additional indicators have been introduced as complements of the PQI and StQI indices: (1) the total withdrawal volume (TWV_{out} – m^3); (2) the thermal energy in the treated sludge (ThE_{out} – cal/d); and (3) the biodegradable of the treated sludge ($bCOD_{out}$ – kgO_2/d):

$$TWV_{out} = \sum_{i=1}^{N} V_{out}^{(i)} \; ; \; ThE_{out} = \frac{\sum_{i=1}^{N}\left[C_{p,H_2O} \cdot \rho_{H_2O} \cdot V_{out}^{(i)} \cdot T_{out}^{(i)}\right]}{N \cdot T_{cycle}} \; ; \; bCOD_{out} = \frac{\sum_{i=1}^{N}\left[V_{out}^{(i)} \cdot bCOD_{out}^{(i)}\right]}{N \cdot T_{cycle}} \quad (6)$$

(iv) Simulation procedure

Like in the BSM2, this benchmark has four different events in a pre-defined 2-years simulation procedure, in order to validate the control strategies performance (see **Figure 5**). First, at T_{sim}=0, a constant influent sludge is applied to the process and simulated to reach a steady state regime; here the air flow-rate applied is constant. At T_{sim}=100*d*, these final values are the initial conditions for a simulation period with a dynamic influent. Next, at T_{sim}=182*d*, the dynamic influent is still applied and here the control strategy to be assessed is activated. Finally, from T_{sim}=364*d* to 728*d*, and keeping the monitor and control tasks, the performance indices in the benchmark evaluation are computed.

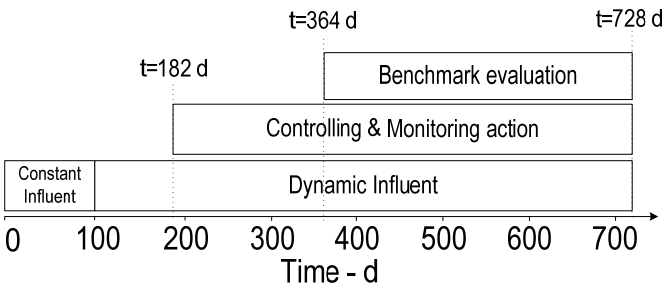

Figure 5. Simulation procedure

Table 2. OL strategy: performance results

PQI	%	100
StQI	%	100
TWV_{out}	m^3	63852
ThE_{out}	*Mcal/d*	12324
$bCOD_{out}$	*kg O_2/d*	623
OCI	*kWh/d*	5351
AE	*kWh/d*	2457

In addition, the AT_BSM includes an open-loop strategy (OL) aimed at providing a reference basis for the comparison of control strategies. The operating parameters for the ATAD reactor in the OL strategy are: the feeding volume = 200 m^3/cycle; T_{cycle} = 24 hours (feeding phase = 30 min.; reaction phase = 23 hours; decant phase = 30 min.); Q_{air} = 65000 m^3/d. The performance indices values, when the simulation procedure is applied to OL strategy, are summarises in **Table 2**.

Control strategies

Two different approaches for the automatic regulation of the external air flow-rate (Q_{air}) have been designed and then evaluated with the AT_BSM. The proposed control strategies rely on

the basic idea: to automatically detect the depletion of biodegradable organic substrate fed into the digester (bCOD knee). As it was observed in **Figure 3**, after the occurrence of the "bCOD knee", the lack of substrate to be oxidised makes external aeration unnecessary; therefore, it can be stopped until the next cycle, in order to save energy costs.

Strategy 1 (ST1): Automatic switching-off of external aeration

Control strategies based on the automatic detection of bend-points in signal trajectories have been implemented extensively. Indeed, many works on the control of sequencing batch reactors for nutrient removal make use of these techniques to optimise the length of anoxic and aerated phases (Puig et al., 2005). In a similar way, the ST1 has been designed to perform the following actions (see **Figure 6**):

- At the beginning of every reaction phase, the external aeration is switched on and fixed to a constant air flow-rate of Q_{air} = 65000 m^3/d (the same value as in OL strategy).
- During the reaction phase, a real-time signal processing algorithm collects the temperature trajectory in order to detect the occurrence of the "bCOD knee" bend-point.
- If the "bCOD knee" happens, the air supply is switched off until the next cycle.

Strategy 2 (ST2): ST1 combined with air flow-rate regulation from cycle to cycle

The ST1 strategy is effective at managing those cycles in which the air flow-rate is greater that that needed to oxidise the sludge fed into the reactor. Nevertheless, it has no provision for any corrective action on the air flow-rate in situations where "bCOD knees" repeatedly remain undetected. ST2 takes into account these situation and adapts the air flow-rate set-point from batch to batch depending on weather the "bCOD knee" is observed or not. As in ST1, the air flow-rate is constant during the reaction phase, but now this value increase or decrease from batch to batch according to the following algorithm (see **Figure 7**):

$$Q_{air}^{(i+1)} = Q_{air}^{(i)} + k_a^{(i)} \cdot \Delta Q \; ; \qquad k_a^{(i)} = \begin{cases} 1 & \text{IF "knee}^{(i)}\text{" NOT detected} \\ -1 & \text{otherwise} \end{cases} \tag{7}$$

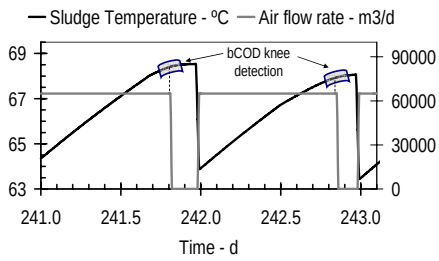

Figure 6. ST1 performance: switching-off of aeration after detection of the "bCOD knee"

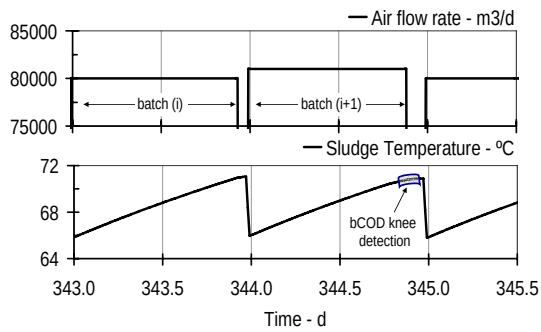

Figure 7. ST2 performance: air flow-rate reduction when a bCOD knee is detected

Although ST2 involves higher aeration costs compared to ST1, a maximum degree of stabilisation in the final sludge is achieved. Accordingly, with the ST2 algorithm, maximum heat is generated biologically which, a priori, pushes the ATAD to reach higher temperatures. If not appropriately controlled, these temperatures can exceed safe limits for thermophilic micro-organisms. Therefore, it is recommended that ST2 implements an upper limit value for the ATAD temperature (e.g., 65º C) so that increments in Q_{air} are allowed only if the temperature in the ATAD is below the upper limit.

Software Implementation

A multi-layer model architecture has been proposed for the software implementation of the AT_BSM plant-layout. The idea of introducing model layers is to establish a clear separation between the mathematical modelling of components related to the treatment process (tanks, reactors, hydraulic flows) and, the automation and control devices (sensors, actuators, information flows). In this respect, architecture includes four basics layers for: mass, sensors, controllers and actuators. The mass layer contains the mathematical model that describes the dynamics of different unit processes (in this case the Storage Tank and ATAD). The sensors and actuators layer contain elements that interact from/to the process (in this work they are considered ideals, no delays and noise). The controller's layer contains the algorithm of control strategies to be tested, its input signals come from sensors layer and its output signal are connected to the actuators layer. The interconnection between the layers is schematically described in **Figure 8**.

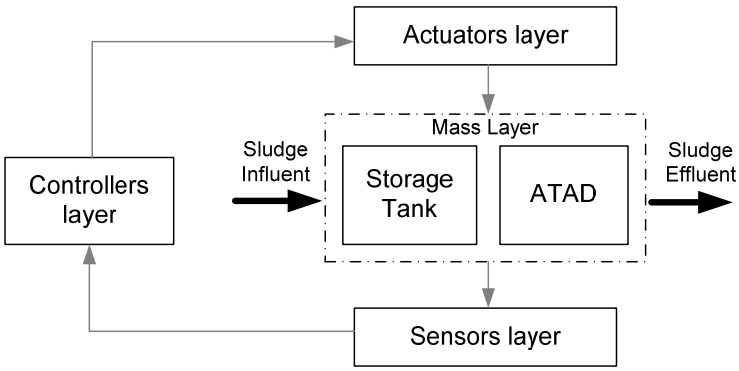

Figure 8. General schema of the multi-layer model architecture. Black arrows refer to flow data, gray arrows refer to information data.

Figure 9 shows how the AT_BSM plant-layout has been implemented in Matlab/Simulink. The blocks structure of Matlab/Simulink allows the interconnection between blocks; furthermore, any variable state from the unit processes can be plotted in Scopes and/or be recorded in vectors for further analysis.

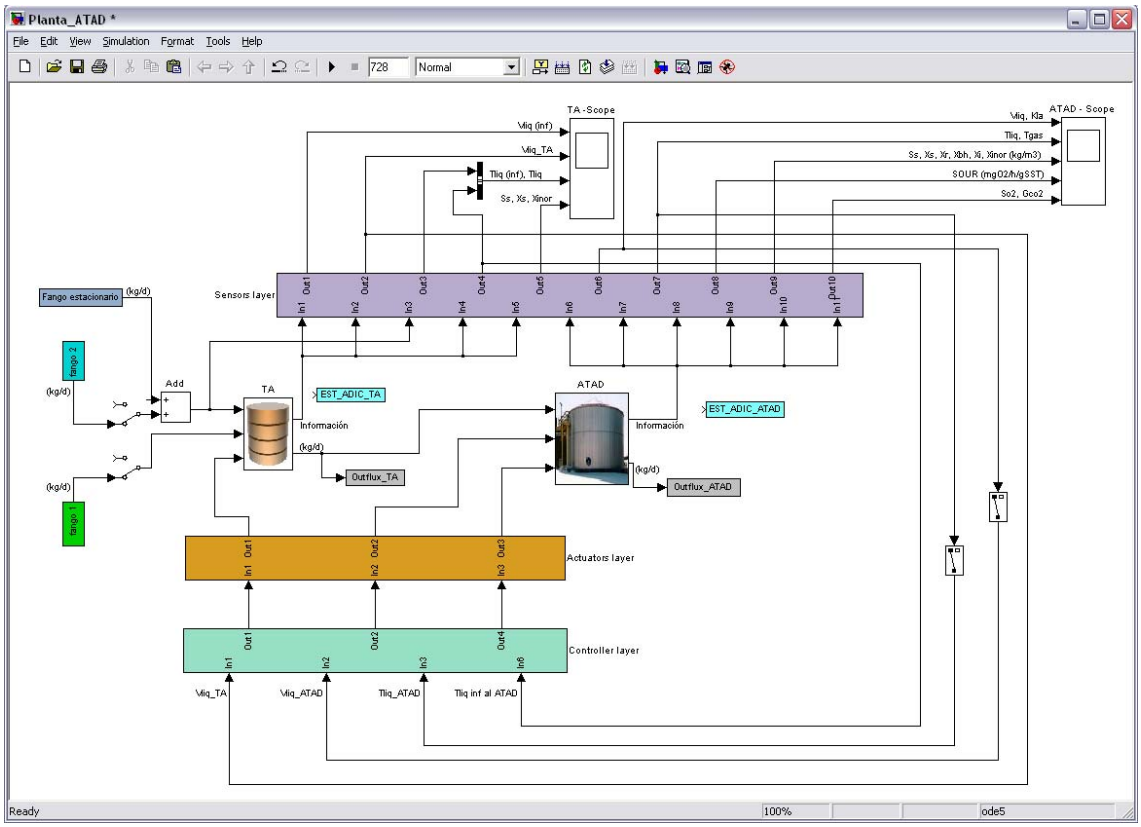

Figure 9. Software implementation of the ATAD layout in Matlab/Simulink.

Benchmark results

Table 3 shows the values corresponding to the performance indices for ST1 and ST2, taken, compared to the OL strategy. As expected, ST1 produces the same sludge quality as the OL strategy (PQI and StQI values). In contrast, savings in aeration are achieved with ST1 (≈-1.5%), but less than expected. Looking at the 141 "bCOD knees" observed in ST1 revealed that most of them occurred close to the end of the reaction phase, which explains the reason for such a small reduction in aeration costs.

In comparison to OL and ST1, ST2 leads to a significantly smaller value of bCODout (≈-17%), which means a more stabilised treated sludge. The increase in the number of "bCOD knees" (141 in ST1 versus 312 in ST2) confirms the above conclusion. Nevertheless, these results are achieved at expense of higher aeration (21%). Therefore, taking the minimisation of aeration costs as the premise of operation, ST1 is more effective than ST2. In contrast, if the degree of stabilisation in the treated sludge is the principal requirements, then the ST2 strategy becomes superior.

Table 3. Results of the performance indices for ST1 and ST2

Strategy	PQI	Knees	StQI	TWV_{out}	ThE_{out}	$bCOD_{out}$	OCI	AE
	%		*%*	*m^3*	*Mcal/d*	*Kg O2/d*	*kWh/d*	*kWh/d*
OL	100		100	63852	12324	623	5351	2457
ST1	100 (-)	141	100 (-)	63885 (0.05%)	12360 (0.3%)	627 (0.6%)	5314 (-0.7%)	2420 (-1.5%)
ST2	100 (-)	312	100 (-)	63370 (-0.8%)	12008 (-2.6%)	518 (-16.9%)	5857 (9.5%)	2963 (20.6%)

In brackets, performance results expressed as percentage with respect to that of the OL strategy

The ThEout index for the three strategies reveals that ST2 leads to lower temperatures in the ATAD than that obtained with OL or ST1. This result contradicts the initial assumption that the application of the ST2 strategy would increase the ATAD temperature. However, the reason for this behaviour lies in the heat losses associated with the air flow-rate and, in particular, with the water evaporation effect. As the ST2 strategy increases the air flow-rate from batch to batch in order to find a maximum stabilisation of the sludge, the evaporative heat losses also increase. Within the normal operating range (i.e., air flow-rates around 65000 m3/d), changes in the air flow-rate don't have significant effects on the evaporative losses. On the other hand, at very high air flow-rates, the cooling effect due to evaporation prevails over the heat generated biologically, and causes a decrease in the sludge temperature. In this respect, simulation results agree with experimental observations of this cooling effect under over—aerated conditions (Cheng and Zhu, 2008).

In order to overcome the ST2 limitation, two following approaches should be dealt with in future studies: (1) An air/oxygen based aeration; and (2) An automatic control of the feeding volume per cycle. About the air/oxygen aeration, an advantage of using pure-oxygen systems is that evaporative heat losses are considerably reduced. Thus, since air is not effective at very high flow-rates, ST2 combined with an automatic on/off of oxygen supply might be an effective solution to ensure maximum sludge stabilisation with no risk of cooling effects due to water evaporation. The air system would work alone (i.e., with the oxygen supply switched off), unless the air flow-rate set-point reached a preset upper limit value. In that case, ST2 would set the air flow-rate to its upper limit and, simultaneously, the pure oxygen supply would be switched on. Under those cases, any perturbation in the oxygen demand would be handled by regulating the oxygen flow-rate.

ST2 combined with automatic regulation of the feeding volume from batch to batch would also allow greater control on the evaporative losses. ST2 would work as it was referred in this work, unless the air flow-rate set-point reached a preset upper limit value. In that case, ST2 would set the air flow-rate to its upper limit and, simultaneously, the feeding volume would be reduced. Under these conditions, any perturbation in the oxygen demand would be handled by regulating the feeding volume.

5. Conclusions

Knowing the importance that the BSM1 protocol has taken in the developing of automatic control in the secondary treatment of WWTPs, the implementation of similar protocols for the sludge treatment would also seem necessary to meet the challenges of integrating enhanced controllers into sludge digestion technologies. In this respect, the proposed AT_BSM protocol has proved its usefulness at analysing the performance of control strategies for ATAD technology. In addition, with minor modifications, the AT_BSM can be easily adapted to enable control studies in different ATAD configurations, for instance, ATADs designed to operate as pre-treatment unit for anaerobic digesters.

The simulation study has concluded that with appropriate control strategies for aeration, either energy savings or enhanced sludge qualities in the effluent can be obtained. Further steps should address the experimental verification of simulation results. When requirements for maximum sludge stabilisation are prioritised, the proposed control strategy for aeration (ST2) promotes cooling effects due to high evaporation losses. In this respect, the design of more complex control strategies based on additional manipulated variables (such as pure-oxygen supply or the feeding volume per cycle) should be investigated in the future.

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Appendix

A. Notation and Abbreviations

AE	Aeration Energy
ASM	Activated Sludge Model
ATAD	Autothermal Thermophilic Aerobic Digestion
AT_BSM	ATAD Benchmark Simulation Model
bCOD	Biodegradable Chemical Oxygen Demand
BSM	Benchmark Simulation Model
C_p	Specific heat
H	Enthalpy
Kla	Oxygen transfer coefficient
k_{paste}	Indicator of pasteurized cycle
k_{st}	Indicator of stabilised cycle
ME	Mixing Energy
N	Total number of batch cycles
OCI	Operational Cost Index
OL	Open-Loop Strategy
ORP	Oxidation Reduction Potential
PE	Pumping Energy
PQI	Pasteurization Quality Index
PTime	Partial Time
Q_{air}	Volumetric air flow-rate
Q_{in}	Influent volumetric flow-rate to the plant
Q_{out}	Effluent volumetric flow-rate from the plant
RI	Robustness Index
SOUR	Specific Oxygen Uptake Rate
SRT	Solid Retention Time
ST1	Strategy 1
ST2	Strategy 2
StQI	Stabilisation Quality Index
t	Time
T	Temperature
T_{cycle}	Time of cycle duration
TCI	Total Cost Index
ThE	Thermal Energy
T_{sim}	Time of simulation
TSS	Total Suspended Solids
TWV	Total Withdrawal Volume
V	Volume
VS	Volatile Solids
WWTP	Waste Water Treatment Plant
X	Concentration of element

Greek letters

Δ	Difference
ν_{ij}	Stoichiometric coefficient for a specific component X_i in the transformation j
ρ_j	Kinetic of transformation j

Subscript

i	Refers to a specific model component
j	Refers to a specific transformation
k	Refers to the volume in which de component is defined

Superscript

i	Specific batch number respect to the total batches

B. Publications of this work

Zambrano J., Gil-Martínez M., García-Sanz M. and Irizar I. (2009). Benchmarking of control strategies for ATAD technology – a first approach to the automatic control of sludge treatment systems. Accepted for Oral presentation at the *10th IWA Conference on Instrumentation Control and Automation (ICA)*, Australia, June 14-17.

Zambrano J., Gil-Martínez M., García-Sanz M. and Irizar I. (2009). First approach in Automatic Control for sludge treatment Systems. A case study: ATAD technology. *Wat. Sci. Tech. (Submitted)*

Benchmarking of control strategies for ATAD technology - a first approach to the automatic control of sludge treatment systems

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Abstract
Autothermal Thermophilic Aerobic Digestion (*ATAD* technology) is a promising alternative to conventional digestion systems. Aeration is a key factor in the performance of these kinds of reactors, in relation to effluent quality and operating costs. At present, the realisation of automatic control in *ATAD*s is in its infancy. Additionally, the lack of robust sensors also makes the control of these processes difficult. Only Oxidation Reduction Potential (*ORP*) and temperature probes are reliable for operation in full-scale plants. Based on the existing *BSM* protocols for benchmarking of control strategies in wastewater treatment plants (*WWTP*), this paper presents the implementation of a protocol specifically adapted to the needs of *ATAD* technology. The implemented protocol has been used to validate two different control strategies for aeration (*ST1* and *ST2*). In comparison to an open-loop operation for the ATAD, simulation results showed that the *ST1* strategy was able to save aeration costs of around 2-4%. Unlike *ST1*, *ST2* achieved maximum sludge stabilisation, but at the expense of higher aeration costs.
Keywords
Automation; Benchmark; Real-time control; Sludge digestion.

INTRODUCTION

Autothermal Thermophilic Aerobic Digestion (*ATAD* technology) was introduced some 30 years ago as a means of retrofitting conventional digesters in order to deal with more stringent regulations for the disposal of municipal wastewater sludge. The operating principle of *ATAD* processes relies on conserving the heat energy released during the aerobic biodegradation of the organics substances contained in the sludge so as to reach and maintain *thermophilic* temperatures (50-70 ºC). Amongst other advantages, working in the *thermophilic* range means that: (1) sludge stabilization can be achieved in shorter detention times (*SRT* = 6-12 days), resulting in smaller volumes; and (2) sludge pasteurization is also feasible, due to high pathogen destruction efficiencies. The operation of full-scale *ATAD* facilities in different countries has confirmed the suitability of this technology to provide a final product which meets all the standards for unrestricted application and reuse of municipal sludges (USEPA, 1990; Layden *et al.*, 2007).

Aeration in all aerobic biological systems is one of the most important considerations since it affects both the quality of the effluent and the total operating costs. The *ATAD* system is no exception, the influence of aeration being even more marked. Over-aeration increases costs without leading to a significantly better quality of treated sludge. Moreover, in the case of air-based aerating systems, over-aeration involves a cooling effect on the slurry with the consequent risk of pasteurization temperatures not being reached. On the other hand, under-aeration limits the efficiency for stabilization and heat generation. Also, under-aeration promotes anaerobic conditions, which increases the potential for undesired odours in the outlet off-gas (Staton *et al.*, 2001).

Automatic control for *ATAD* technology is limited to the use of very elemental strategies for aeration. In fact, the first *ATAD* generation did not consider any capability for the regulation of the aeration system and it was not until the advent of the second *ATAD* generation that these systems began to be equipped with more sophisticated devices where the flow-rate of the injected gas stream could be automatically manipulated (Scisson, 2003). Even so, the lack of robust and reliable online

sensors is still a limitation that has hindered the deployment of monitoring and control tools for these systems. Industrial sensors for dissolved oxygen or suspended solids, for example, are intended for use in the secondary treatment, where temperatures and solid concentrations are not excessive. However, these sensors cannot withstand the aggressive environment within ATAD tanks, due to the corrosive nature of the digesting sludge and the high values of temperature and solid concentrations. At present, only *ORP* and temperature sensors fulfil the technical features required for operation in full-scale *ATAD*s. This is the reason why aeration control in *ATAD*s has so far been based on either *ORP* or temperature sensors. Staton *et al.* (2001) stated that with an appropriate processing of the *ORP* signal, the depletion of biodegradable organic substrate in the digester can be automatically detected (end of aerobic oxidation transformations); thus, on the basis of this observation, control strategies for external aeration can be implemented to save aeration costs. Breider and Drnevich (1981) probably hold the first patent on real-time control of *ATAD*s. They propose a very simple strategy that automatically modifies the inlet gas flow-rate as a function of variations in temperature, with the objective of maintaining it within a predetermined range.

Nowadays, mathematical modelling and simulation have become essential tools for supporting not only the design and operation of *WWTP*s but also the analysis and synthesis of automatic controllers. The original Benchmark Simulation Model no. 1 (*BSM1)* protocol and its sequels: Long-Term BSM1 (*BSM1_LT)* and the *BSM2*, are three important simulation benchmarks aimed at providing researchers with standard methodologies for the objective comparison of control strategies (Copp, 2002). Nevertheless, *ATAD* technology is not supported by any of the existing *BSM*s and, therefore, studies on *ATAD* control cannot be undertaken with any of these protocols. In contrast, the last few years have witnessed significant advances in dynamic modelling of *ATAD* systems. Gómez *et al.* (2007) bring together existing formulations on biochemical reactions, physico-chemical transformations and thermal energy balances, so as to develop a comprehensive model for the *ATAD* which includes dynamic prediction of liquid and gas compounds as well as temperature. Similarly, Kovács *et al.* (2007) proposes an extension of the standard Activated Sludge Model No 1 (*ASM1*) with the incorporation of thermophilic bacteria and their respective biochemical transformations. Thus, both these model approaches open the door for either the integration of the *ATAD* into current *BSM* benchmarks or even the definition of a new benchmark protocol specifically for *ATAD*s, such as proposed in this paper.

In the present work two major objectives have been pursued: (1) the definition of a benchmark protocol for evaluation of controllers in *ATAD* systems; and (2) the design and evaluation of new control strategies for aeration. The paper has been organised as follows: firstly, the definition of the benchmark protocol is described; next, two different control approaches for external aeration are proposed; finally, the performance of each control strategy is assessed in terms of the benchmark.

BENCHMARK PROTOCOL FOR THE ATAD PROCESS

On the basis of the standard *BSM* protocols, an *ad hoc* benchmark for the *ATAD* process has been specifically defined (henceforth referred to as "*AT_BSM*"). Implementation of the *AT_BSM* protocol has involved the following four major definitions: (1) Influent definition; (2) Plant layout and plant-model; (3) Evaluation criteria; and (4) Simulation procedure.

- **Influent definition**

Given the difficulty of compiling an historic data file with comprehensive information about the short-term and long-term characteristics (flow and composition) of a typical sludge in a full-scale plant, it was decided to automatically generate the file by selecting a representative virtual plant. In this respect, for the sake of simplicity, the virtual plant of the *BSM2* was chosen and simulated

according to the *BSM2* simulation procedure (Vrecko *et al.*, 2006). Simulation results were saved at regular intervals of 15 minutes so as to create an influent file (sludge_lt.txt) with the characteristics of the sludge (both, primary and secondary) for a 728-days period of plant performance. **Table 1** summarises some major features of the raw sludge obtained.

Table 1. Influent file in the *AT_BSM*

Sludge features	Min	Max	Average
Flow rate [m3/d]	39 / 17 / 56	544 / 102 / 646	148 / 40 / 188
COD [g/L]	2.2 / 85 / 14	74/ 95 / 77	35 / 86 / 47
TSS [g/L]	1.5 / 60 / 10.3	49 / 67 / 52	24 / 61 / 32
Temperature [°C]	9.5	20.5	15

** Primary/Secondary/Mixed*

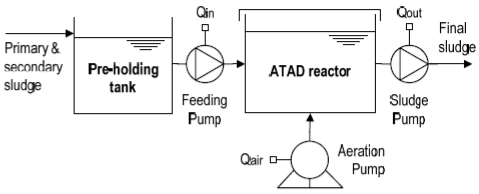

Figure 1. Plant-layout in the *AT_BSM*

- **Plant-layout and mathematical model**

*ATAD*s are generally operated in batch-mode, mainly in scenarios where sludge pasteurization is mandatory and where, therefore, it is advisable to avoid hydraulic short-circuits during digestion. Thus, within the benchmark, a sequencing operation has been imposed for the *ATAD*. In particular, in compliance with *EU* requirements for sludge pasteurization in batch processes, the following 24-hours cyclic sequence has been adopted by default: 0.5 hours for sewage feeding; 23 hours for reaction (aerated reaction phase); and, finally, 0.5 hours for sludge withdrawal. Non-continuous feeding in the *ATAD* makes the utilisation of pre-holding tanks necessary. Accordingly, the plant-layout in the *AT_BSM* has been made up of a pre-holding tank and an *ATAD* single reactor (**Figure 1**). An analysis of the influent data file was performed to obtain appropriate values for the effective volume of both the pre-holding tank (V_{max} = 2000 m^3) and the digester (V_{max} = 2400 m^3).

Concerning the plant-model, the pre-holding tank has been modelled as a completely-stirred variable-volume basin where only mass transport has been considered (neither biological reactions nor heat transformations have been included). The *ATAD* reactor has been modelled as a completely-stirred tank; but, in this case, biological and heat effects have also been added. The biochemical model is based on the standard *ASM1* with slight modifications to make it consistent with observations from *thermophilic* aerobic digesters. Additionally, previous works on the dynamic prediction of temperatures in biological tanks (Vismara, 1985; Messenger *et al.*, 1990) form the basis of the heat model. A detailed definition of the *ATAD* model used in the *AT_BSM* can be found in Gómez *et al.* (2007).

- **Evaluation criteria**

Analogously to the standard benchmark protocols, three major performance indices have been incorporated into the *AT_BSM* in order to compare control strategies: (1) the Operational Cost Index (*OCI*); (2) the Pasteurisation Quality index (*PQI*); and (3) the Stabilisation Quality index (*StQI*). The *OCI* takes into account all the energy costs involved in the operation of the *ATAD* reactor and has been calculated in a way similar to that undertaken in the *BSM2,* but using a non-weighted sum:

$$OCI(\text{kWh}) = AE + PE + ME \qquad (1)$$

, where *AE* represents the energy for external aeration, *PE* is the pumping energy and *ME* refers to the mixing energy. *PE* covers the feeding of the raw sludge into the *ATAD* as well as the withdrawal of the treated sludge from the *ATAD*. For the mixing energy, only the energy required for mixing the *ATAD* has been considered.

PQI and *StQI* have been introduced to quantify the degree of pasteurization and stabilisation of the sludge leaving the *ATAD,* respectively. Due to the lack of consensus on the definition of general

criteria for sludge pasteurisation, the *EU* recommendation applicable to batch digesters has been adopted here. It recommends: *"thermophilic aerobic digestion at a temperature of at least 55°C for 20 hours as a batch, without admixture or withdrawal during the treatment"*. Accordingly, *PQI* (%) has been incorporated into the benchmark to represent the percentage of *ATAD* cycles in which the sludge leaving the treatment complies with the above definition. Since in the more general case the decant volume per cycle (V_{out}) might change from cycle to cycle, *PQI* has been formulated in terms of mass fluxes per cycle, as follows:

$$PQI(\%) = \frac{\sum_{i=1}^{N}\left[k_{paste.}^{(i)} \cdot V_{out}^{(i)} \cdot TSS_{out}^{(i)}\right]}{\sum_{i=1}^{N}\left[V_{out}^{(i)} \cdot TSS_{out}^{(i)}\right]} \cdot 100 \text{, where: } k_{paste.}^{(i)} = \begin{cases} 0; & if\ PTime^{(i)} < 20\,hrs \\ 1; & if\ PTime^{(i)} > 20\,hrs \end{cases} \quad (2)$$

, N being the total number of batch cycles, i the i-th batch, and TSS_{out} the total suspended solids concentration in the exiting sludge. $PTime^{(i)}$ represents the total time in which the sludge has been at a temperature greater than 55 °C during the aerated reaction phase of the i-th batch.

A review of existing policies and guidelines for sludge management in different countries shows the diversity of criteria used to specify the requirements for sludge stabilisation. In fact, such requirements are strongly conditioned by the type of treatment used for digestion. For example, as far as aerobic digestion is concerned, the *U.S. EPA* regulation 40 CFR Part 503 (USEPA, 1993) establishes three options for compliance with the vector attraction reduction requirements (*i.e.*, sludge stabilisation). *Option 1* refers to: "*at least 38% reduction in volatile solids during sewage*"; *Option 2* to: "*less than 15% additional volatile solids reduction during bench-scale aerobic batch digestion for 30 additional days at 20°C*"; and *Option 3* to: "Specific Oxygen Uptake Rate (*SOUR) at 20°C less than 1.5 mg O_2/hr/g total sewage sludge solids*". *Option 3* is only applicable to *mesophilic* aerobic digesters; *Option 2* is only valid for aerobically digested sewage sludge with *2%* or less solids. Unlike *Options 2* and *3*, *Option 1* is not restricted to any specific treatment technology; however, this option has certain limitations since it is not completely appropriate for treatments where the incoming sludge has been partially pre-stabilised (for example, sewage sludge from secondary treatments operated at medium/large *SRT*). In these situations, *Option 2* should be used instead. Since *Options 1* and *2* are valid for aerobic *thermophilic* digestion, a combination of both has been adopted to formulate *StI*:

$$StI(\%) = \frac{\sum_{i=1}^{N}\left[k_{st}^{(i)} \cdot V_{out}^{(i)} \cdot VS_{out}^{(i)}\right]}{\sum_{i=1}^{N}\left[V_{out}^{(i)} \cdot VS_{out}^{(i)}\right]} \cdot 100 \text{, where } k_{st}^{(i)} = \begin{cases} 1 & if \begin{Bmatrix} Option1^{(i)} \\ or \\ Option2^{(i)} \end{Bmatrix} is\ met \\ 0 & \text{Otherwise} \end{cases} \quad (3)$$

$Option1^{(i)}$ and $Option2^{(i)}$ are Boolean variables whose values result from the evaluation, at the end of the i-th batch, of the respective *Option 1* and *Option2* statements mentioned above; VS_{out} represents the volatile solids of the effluent sludge. Finally, complementary to *PQI* and *StQI*, three additional indicators have been included: (1) the total withdrawal volume (TWV_{out} - m^3); (2) the thermal energy in the treated sludge (ThE_{out} - cal/d); and (3) the biodegradability of the treated sludge ($bCOD_{out}$ - kg O_2/d):

$$TWV_{out} = \sum_{i=1}^{N} V_{out}^{(i)} \qquad ThE_{out} = \frac{\sum_{i=1}^{N}\left[C_{p,H_2O} \cdot \rho_{H_2O} \cdot V_{out}^{(i)} \cdot T_{out}^{(i)}\right]}{N \cdot T_{cycle}} \qquad bCOD_{out} = \frac{\sum_{i=1}^{N}\left[V_{out}^{(i)} \cdot bCOD_{out}^{(i)}\right]}{N \cdot T_{cycle}} \quad (4)$$

- **Sensors and actuators**

Control strategies are constrained to the use of the following online measurements: water level and temperature in the pre-holding tank; inlet sludge flow-rate, outlet sludge flow-rate, water level, air flow-rate and temperature in the *ATAD*. Furthermore, the selected plant-layout (see **Figure 1**) leads to the three following manipulated variables: (1) the feeding volume per cycle (Q_{in} - m^3/cycle); (2) the outlet sludge volume per cycle (Q_{out} - m^3/cycle); and (3) the external air flow-rate (Q_{air} - m^3/h). In the current version, all the sensors and actuators have been modelled as "ideal" (instantaneous dynamic response and no-noise).

- **Simulation procedure**

Like in the *BSM2*, every control strategy is assessed according to a pre-defined 2-year simulation procedure in which four events are distinguished (**Figure 2**). At $T_{sim}=0$, the simulation procedure starts with the process operating under constant conditions in order to reach a steady-state regime; at $T_{sim}=100$ *days*, the sludge defined in the influent file "sludge_lt.txt" starts to be fed into the pre-holding tank; at $T_{sim}= 182$ *days*, the control strategy to be assessed must be activated; finally, from $T_{sim} = 364$ *days* to *728 days*, the results of the performance indices defined above are computed.

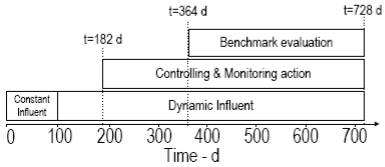

Figure 2. Simulation procedure

Table 2. OL strategy: performance results

PQI	%	100
StQI	%	100
TWV_{out}	m^3	63852
ThE_{out}	*Mcal/d*	12324
$bCOD_{out}$	*kg* O_2*/d*	623
OCI	*kWh/d*	5351
AE	*kWh/d*	2457

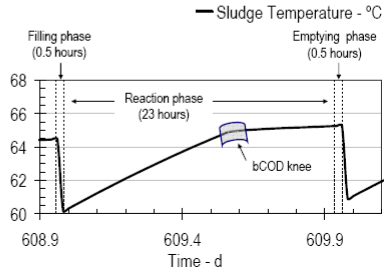

Figure 3. Temperature profile for an over-aerated batch

Additionally, the *AT_BSM* includes an <u>*open-loop strategy*</u> (*OL*) aimed at providing a reference basis for the comparison of control strategies. The operating parameters for the *ATAD* reactor in the *OL* strategy are: the feeding volume = 200 m^3/cycle; T_{cycle} = 24 hours (feeding phase = 30 min.; reaction phase = 23.5 hours; decant phase = 30 min.); Q_{air} = 65000 m^3/d. **Table 2** summarises the values of the performance indices that result from the application of the simulation procedure to *OL*.

AUTOMATIC CONTROL OF EXTERNAL AERATION

Two different approaches for the automatic regulation of the external air flow-rate (Q_{air}) have been designed and then evaluated with the *AT_BSM*. The proposed control strategies rely on the same basic idea: to automatically detect the depletion of biodegradable organic substrate fed into the digester. **Figure 3** shows an example of the typical temperature trajectories that take place in the *ATAD* when it is operated at both influent under-loads and constant air-flow rates. It is observed that during the reaction phase the temperature trajectory has a bend-point, which is related to the depletion of biodegradable substrate in the reactor ("*bCOD* knee"). After the occurrence of the "*bCOD knee*", the lack of substrate to be oxidised makes external aeration unnecessary; therefore, it can be stopped until the next cycle in order to save energy costs.

- **Strategy 1 (ST1): Automatic switching-off of external aeration**

The design of control strategies based on the automatic detection of bend-points in signal trajectories is not new. In fact, many works on the control of sequencing batch reactors for nutrient removal make use of these techniques to optimise the length of anoxic and aerated phases (Puig *et al.*, 2005). In a similar way, the *ST1* has been designed to perform the following actions (**Figure 4**):

- At the beginning of every reaction phase, the external aeration is switched on and fixed to a constant air flow-rate of Q_{air} = 65000 m^3/d (the same value as in the *OL* strategy)
- During the reaction phase, a real-time signal processing algorithm collects the temperature trajectory with the objective of detecting the occurrence of the "*bCOD* knee" bend-point
- If the "*bCOD* knee" happens, the air supply is automatically switched off until the next cycle

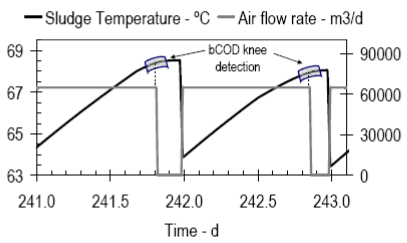

Figure 4. ST1 performance: switching-off of aeration after detection of the "*bCOD* knee"

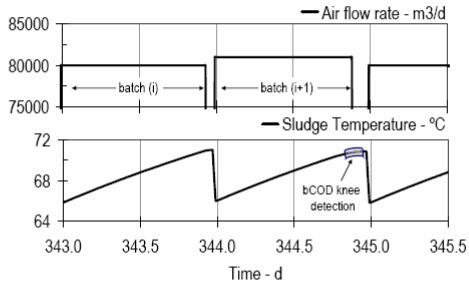

Figure 5. Air flow-rate action in Strategy 2 when a knee is detected during the batch

- **Strategy 2 (ST2): *ST1* combined with air flow-rate regulation from cycle to cycle**

The *ST1* strategy is effective at managing those cycles in which the air flow-rate is greater than that needed to oxidise the sludge fed into the reactor. However, it has no provision for any corrective action on the air flow-rate in situations where "*bCOD* knees" repeatedly remain undetected. *ST2* takes into account the above and adapts the air flow-rate set-point from batch to batch depending on whether the "*bCOD* knee" is observed or not. As in *ST1*, the air flow-rate set-point is constant throughout the reaction phase, but now this value increases or decreases from batch to batch according to the following algorithm (**Figure 5**):

$$Q_{air}^{(i+1)} = Q_{air}^{(i)} + k_a^{(i)} \cdot \Delta Q \; ; \qquad k_a^{(i)} = \begin{cases} 1 & \text{IF "knee}^{(i)}\text{" NOT detected} \\ -1 & \text{otherwise} \end{cases} \tag{5}$$

Although *ST2* involves higher aeration costs compared to *ST1*, a maximum degree of stabilisation in the final sludge is achieved. Accordingly, with the *ST2* algorithm, maximum heat is generated biologically which, a priori, pushes the ATAD to reach higher temperatures. If not appropriately controlled, these temperatures can exceed safe limits for *thermophilic* micro-organisms. Therefore, it is recommended that *ST2* implements an upper limit value for the *ATAD* temperature (e.g., 65 ºC) so that increments in Q_{air} are allowed only if the temperature in the *ATAD* is below the upper limit.

PERFORMANCE ANALYSIS OF THE CONTROL STRATEGIES

The *AT_BSM* and the proposed control strategies have been implemented and simulated using the Matlab/Simulink platform. **Table 3** shows the values corresponding to the performance indices for *ST1* and *ST2*. As expected, *ST1* produces the same sludge quality as the *OL* strategy (*PQI* and *StQI* values). In contrast, savings in aeration are achieved with *ST1* (≈ -1.5%), but less than anticipated. An analysis of the 141 "*bCOD* knees" observed in *ST1* revealed that most of them occurred close to the end of the reaction phase, which explains the reason for such a small reduction in aeration costs. Nevertheless, higher energy savings would have been obtained with *ST1,* if air flow-rates greater than 65000 m^3/d had been employed in the *OL* strategy. For example, by setting the air flow-rate to 80000 m^3/d, aeration savings increase up to 4%.

In comparison to *OL* and *ST1*, *ST2* leads to a significantly smaller value of $bCOD_{out}$ (≈ -17%), which means a more stabilised treated sludge. The increase in the number of "*bCOD* knees" (141 in

ST1 versus 312 in *ST2*) confirms the above conclusion. Nevertheless, these results are achieved at the expense of higher aeration (21%). Therefore, when aeration costs need to be prioritised, *ST1* is more effective than *ST2*. In contrast, if the degree of stabilisation in the treated sludge is a major requirement, then the *ST2* strategy becomes superior.

Table 3. Results of the performance indices for *ST1* and *ST2*

Strategy	PQI	Knees	StQI	TWV_{out}	ThE_{out}	$bCOD_{out}$	OCI	AE
	%		*%*	*m^3*	*Mcal/d*	*Kg O2/d*	*kWh/d*	*kWh/d*
OL	100		100	63852	12324	623	5351	2457
ST1	100 (-)	141	100 (-)	63885 (0.05%)	12360 (0.3%)	627 (0.6%)	5314 (-0.7%)	2420 (-1.5%)
ST2	100 (-)	312	100 (-)	63370 (-0.8%)	12008 (-2.6%)	518 (-16.9%)	5857 (9.5%)	2963 (20.6%)

In brackets, performance results expressed as percentage with respect to that of the OL strategy

The analysis of the ThE_{out} index for the three strategies concludes/reveals that *ST2* leads to lower temperatures in the *ATAD* than that reached with *ST1* or *OL*. This result contradicts the initial assumption that the application of the *ST2* strategy would cause higher temperatures in the *ATAD*. However, the reason for this behaviour lies in the heat losses associated with the air flow-rate and, in particular, with water evaporation. As *ST2* strategy increases the air flow-rate from batch to batch in order to find a maximum stabilisation of the sludge, the evaporative heat losses also increase. Within the normal operating range (i.e., air flow-rates of 65000 m^3/d), changes in the air flow-rate do not have significant effects on the evaporative losses. Conversely, at very high air flow-rates, the cooling effect due to evaporation prevails over the heat generated biologically, and causes a decrease in the sludge temperature. In this respect, simulation results agree with experimental observations of this cooling effect under over-aerated conditions (Cheng and Zhu, 2008).

In order for the above limitations of the *ST2* strategy to be overcome, the two following approaches should be dealt with in future works/studies: (1), an approach that involves a combined aeration using air-based and oxygen-based supply systems; and (2), an approach that involves an automatic control of the feeding volume per cycle. With respect to air-based injection systems, an advantage of using pure-oxygen systems is that the evaporative heat losses are considerably reduced. Therefore, since air is not effective at very high flow-rates, *ST2* combined with an automatic on/off of the oxygen supply might be an effective solution to ensure maximum sludge stabilisation with no risk of cooling effects due to water evaporation. The air system would work alone (i.e., with the oxygen supply switched off), unless the air flow-rate set-point reached a preset upper limit value. In that case, *ST2* would set the air flow-rate to its upper limit and, simultaneously, the pure oxygen supply would be switched on. Under these circumstances, any perturbation in the oxygen demand would be handled by regulating the oxygen flow-rate. *ST2* combined with automatic regulation of the feeding volume from batch to batch would also allow greater control on the evaporative losses. *ST2* would work as the original approach (see the previous section), unless the air flow-rate set-point reached a preset upper limit value. In that case, *ST2* would set the air flow-rate to its upper limit and, simultaneously, the feeding volume would be reduced. Under these circumstances, any perturbation in the oxygen demand would be handled by regulating the feeding volume.

CONCLUSIONS

Considering the crucial role that the *BSM1* protocol has played in the realisation of automatic control in the secondary treatment of *WWTPs*, the implementation of similar protocols for the sludge treatment would also seem necessary to meet the challenges of integrating enhanced control into sludge technologies. Aligned with this, the proposed *AT_BSM* protocol has proved its usefulness at analysing the performance of control strategies for *ATAD* technology. Moreover, with

slight modifications, the *AT_BSM* can be easily adapted to enable control studies in other *ATAD* configurations, such as for example, *ATAD*s designed to operate as pre-treatment unit for anaerobic digesters. The simulation study has concluded that with appropriate control strategies for aeration, either energy savings or enhanced sludge qualities in the effluent can be obtained. Further steps should address the experimental verification of simulation results. When requirements for maximum sludge stabilisation are prioritised, the proposed control strategy for aeration (*ST2*) promotes cooling effects due to high evaporation losses. In this respect, the design of more complex control strategies based on additional manipulated variables (such as a pure-oxygen supply or the feeding volume per cycle) should be investigated in the future.

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C. Participation in research projects

1. GAITEK Project: "Design and development of advanced automatic control strategies for Autothermal Thermophilic Aerobic Digestion processes (ATAD technology)".
   Principal researcher: Dr. Ion Irizar Picón
   Financed by: Basque Government.
   Period: January 2008 – December 2009.
2. Project: "Design of automatic control strategies in treatment plant of solid residues from a methodology based on the mathematical model and systems simulation" (DPI2006-15522-C02-02).
   Principal researcher: Dr. Antonio Salterain Ezquerra
   Financed by: Spanish Ministry of Science and Innovation.
   Period: July 2006 – July 2009

(Dr. Ion Irizar Picón)	(Dr. Antonio Salterain Ezquerra)

D. Most recent publications related to the work

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Mathematical modelling of autothermal thermophilic aerobic digesters

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ABSTRACT

This paper presents a new mathematical model for Autothermal Thermophilic Aerobic Digesters. The reactor has been modelled as two completely mixed volumes to separately predict the behaviour of the liquid and gaseous phases as well as the interrelation between them. The model includes biochemical transformations based on the standard Activated Sludge Models of IWA, as well as physico-chemical transformations associated with the chemical equilibria and the mass transfer between the liquid and the gaseous phases similar to those proposed in the ADM1 of IWA. An energy balance has also been included in the model in order to predict the temperature of the system. This thermal balance takes into account all those biochemical and physico-chemical transformations that entail the most relevant heat interchanges. Reactor performance has been explored by simulation in two different scenarios: in the first where it acts as the initial stage in a Dual system, and in the second where it acts as a single-stage treatment. Each scenario enabled the identification of the relevance of the different parameters.

1. Introduction

Autothermal thermophilic aerobic digestion (ATAD) is an advanced sewage sludge treatment that provides hygienization. The process has a self-heating ability and removes pathogens when working at a temperature of at least 55 °C for 20 h as a batch (European Commission, 2000). This operation pattern complies with the regulations for obtaining Class A biosolids (US EPA, 2003).

Many experimental studies have been carried out to analyse the ATAD technology in particular cases (Messenger et al., 1993; Kelly and Warren, 1995; Gomez et al., 2007) leading to very helpful process guidelines. However, experiments have limitations concerning time and feasibility. These limitations, combined with the very long time-response of many of the involved process transformations, suggest the usefulness of a simulator based on a dynamic model of ATAD. Using the model certain significant aspects of this technology could thus be explored in a fast and easy way and the performance of ATAD reactors in different scenarios could be predicted. In view of the fact that the temperature in the process is one of the most important aspects to take into account, a thermal model could help develop a more realistic overall model. Previous works on modelling the heat balance can be found in Vismara (1985), Messenger et al. (1990, 1993) and Gillot and Vanrolleghem (2003).

This paper presents a mathematical model to predict the main biodegradation transformations, physico-chemical reactions and energy interchange occurring in an ATAD reactor. The overall model is made up of two sub-models, namely, that of mass balance and that of energy balance. The biochemical transformations are presented in matrix

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notation. They take part in both sub-models, since the degradation of the organic matter is one of the main factors of energy generation.

The construction of the model has been based on a complete characterization for all the model components in their elemental mass fractions, guaranteeing mass continuity for all the process transformations (de Gracia et al., 2006) and facilitating the connection between different unit-process models to simulate a whole wastewater treatment plant. The resulting model has been implemented in the simulation platform WEST®.

The model has been conceived as a tool for ATAD design and operation. It also enables the identification of the relevance of the different parameters in each case that is explored by simulation. The exploration of two different scenarios by simulation has been carried out in order to check the model's capacity to qualitatively predict the process performance. The general parameter estimation is currently under way. A proper calibration of the parameters that are specific for each case will be needed to make it possible to use the model for particular ATAD reactors.

2. Mathematical modelling of ATAD

In order to dynamically simulate ATAD performance, a model of a reactor with two completely stirred volumes (liquid and gaseous phases) has been developed. This model describes the changes in the model components as a result of biochemical and physico-chemical transformations, inlet and outlet mass flow transport and the interactions between the liquid and gaseous phases (stripping–dissolution and condensation–evaporation). Although the biochemical transformations occur only in the liquid phase, it is necessary to consider both phases simultaneously, as the partial pressure of some gaseous components play an important role in the kinetics of the physico-chemical transformations that affect the pH value. The model also includes the energy balance in both volumes to account for the great importance of temperature variation in the reactor.

2.1. *Model of mass balance*

The basic equation for the mass balance of a generic model component X_i within the boundaries of both volumes described above is

$$\frac{d(V_k X_i)}{dt} = \sum_j \nu_{ij}\rho_j V_k + \mathrm{Input}(X_i) - \mathrm{Output}(X_i), \quad (1)$$

where X_i is the concentration of the generic model component, ν_{ij} is the stoichiometric coefficient for the component X_i in the transformation j, ρ_j is the kinetic expression of the transformation j and V_k is the volume of the phase in which the component is defined. The volumes of the liquid phase (V_{liq}) and the gaseous phase (V_{gas}) are also simulated by the model. Input(X_i) and Output(X_i) are the transport terms of the component in mass flow.

The biochemical model includes hydrolysis of the particulate substrate and growth and lysis of the heterotrophic biomass. These transformations are based on the IWA ASM1 (Henze et al., 2000). The mechanisms to describe the acid-base reactions and liquid-gas transfers are also included, the former being based on the IWA ADM1 (Batstone et al., 2002) and the latter on the traditional stationary liquid-film theory (Whitman, 1923) adopted by Siegrist et al. (2002).

Components, transformations, kinetics and stoichiometric coefficients of the biochemical model are expressed in the matrix format proposed by Petersen (1965). The stoichiometry of the transformations is expressed on a mass transfer basis, as the traditionally used unit of concentration cannot be used to define the liquid-gas mass transfer processes. Altogether, there are 20 components involved in the 11 processes considered.

The model components are characterized in terms of their CHONP content and charge density. From this elemental composition the relationship between mass and COD and the conversion parameters $\beta_{E,i}$ (e.g. $\beta_{C,i}$ mass of carbon per stoichiometric unit of the component i) can be directly calculated for each of the components. This approach characterizes organic waste in terms of measurable parameters in accordance with the COD, C and N mass balances approach and leads to a complete and rigorous model definition (de Gracia et al., 2006).

Closed mass balances require some model components to act as source-sink (or mass "compensation terms") for every element. This role has been associated with the model components carbon dioxide (S_{CO_2}), water (S_{H_2O}), dissolved oxygen (S_{O_2}), ammonia ($S_{NH_4^+}$), dehydrogen phosphate ($S_{H_2PO_4^-}$) and hydrogen cations (S_{H^+}) for C, H, O, N, P and charge (Ch) continuity, respectively. Their stoichiometric coefficients are computed by conservation equations as shown in Table 1, where $\beta_{E,i}$ is the conversion factor from the units of the component i to the units of the element or charge (E) to which conservation is to be applied.

Table 2 shows the complete stoichiometric matrix of the ATAD model. The kinetic expressions (ρ_j) for each transformation (j) are recorded in the leftmost column of the matrix.

Additional comments must be made for some of the kinetics as follows. The hydrolysis rate of particulate substrate (ρ_1) includes both aerobic and anaerobic hydrolysis depending on the dissolved oxygen concentration, since the oxygen limited conditions of the ATAD reactor implies the process is neither absolutely aerobic nor anaerobic. In this model only one hydrolysis transformation has been included, which considers a single generic kind of sludge as a mixture of primary and secondary sludge.

Transformations 4–7 deal with the mass transfer between the liquid and the gaseous phases. The expression of the kinetics of water evaporation is proportional to the difference between the actual water vapour concentration in the gas and the concentration of saturated water vapour $G_{H_2O}^{SAT}$ in the boundary layer interface, with the proportionality factor $k_M a$ (dependent on the area between phases and the effect of the reactor systems). The mass of the (dissolved) gas i continuously transferred from the liquid to the gaseous phase equals the gas transfer coefficient ($k_L a_i$) multiplied by the difference between the dissolved gas saturation concentration in the bulk liquid (S_i) and the dissolved gas concentration at the boundary layer interface, the latter being obtained as the product of the Henry constant for this component (H_i) and

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Table 1 – Stoichiometric coefficients of the compensation terms to complete the mass and charge balance of the model

Conserved element	C	H	O	N	P	Charge
Compensation term	S_{CO_2} (i = 6)	S_{H_2O} (i = 1)	S_{O_2} (i = 4)	$S_{NH_4^+}$ (i = 8)	$S_{H_2PO_4^-}$ (i = 12)	S_{H^+} (i = 10)
Stoichiometric coefficient in transformation j	$\upsilon_{6j} = -(\sum_i \beta_{C,i}\upsilon_{ij})/\beta_{C,6}$	$\upsilon_{1j} = -(\sum_i \beta_{H,i}\upsilon_{ij})/\beta_{H,1}$	$\upsilon_{4j} = -(\sum_i \beta_{O,i}\upsilon_{ij})/\beta_{O,4}$	$\upsilon_{8j} = -(\sum_i \beta_{N,i}\upsilon_{ij})/\beta_{N,8}$	$\upsilon_{12j} = -(\sum_i \beta_{P,i}\upsilon_{ij})/\beta_{P,12}$	$\upsilon_{10j} = -(\sum_i \beta_{Ch,i}\upsilon_{ij})/\beta_{Ch,10}$

the partial pressure of this component in the gas (Pp_i) calculated from the concentration of its component G_i. The parameters ($k_L a_i$) depend on the area between phases, the air feed rate and the effect of the reactor systems, and are related to each other through the diffusivity of each of the gases.

The total volume gas flow q_{tot} (m^3/day) is calculated with a pressure control loop that keeps the sum of the partial pressures ΣPp_i (atm) in the gaseous volume as close as possible to the pressure of work P_{work} (atm) in the reactor headspace.

$$q_{tot} = k_{gas}\left(\sum Pp_i - P_{work}\right) \quad [m^3/day], \tag{2}$$

where K_{gas} is the proportional control constant (m^3/day atm).

2.2. *Model of energy*

In order to model the evolution of the temperature in the ATAD reactor, an energy balance has been applied to the system. One of the factors that affects net enthalpy within the boundaries of the liquid phase is the heat associated with the model transformations. To be rigorous, all the model transformations imply an enthalpy variation per stoichiometric unit, ΔH_j (kJ/kg). However, two of the terms of these enthalpy variations are significantly higher than the others: one is associated with the growth of the heterotrophic bacteria and the other wth water evaporation–condensation. Consequently, only these two terms are considered in this model.

To predict the temperature variations in both liquid and gaseous phases, an equation for energy balance has been applied to each phase. Two more state variables have been defined for that purpose, namely the net enthalpy in the liquid phase $H_{net\ liq}$ (kJ) and the net enthalpy in the gas phase $H_{net\ gas}$ (kJ). Fig. 1 presents the terms considered for the proposed energy flow balance, which are described in Eqs. (3)–(13).

The basic equation for the energy balance within the boundaries of the liquid phase is

$$\frac{dH_{net\ liq}}{dt} = \text{Input}(H_{liq}) - \text{Output}(H_{liq}) + Pw_{bio} - Pw_{ve} + Pw_{mi} - Pw_{we\ liq} - Pw_{phase}^{conduction} \quad [kJ/day], \tag{3}$$

where

- Input(H_{liq}) and Output(H_{liq}) are the values of power associated with the liquid influent at a temperature T_{inf} and the liquid effluent at the temperature of the liquid phase T_{liq}.

$$\text{Input}(H_{liq}) = m_{inf} C_p T_{inf} \quad [kJ/day], \tag{4}$$

$$\text{Output}(H_{liq}) = m_{out} C_p T_{liq} \quad [kJ/day], \tag{5}$$

where m_{inf} and m_{out} are, respectively, the mass inflow (kg/day) and the mass outflow (kg/day) and and C_p is the specific heat of the sludge, considered equal to the specific heat of the water component S_{H_2O} (kJ/kg °C).

- Pw_{bio} is the power produced in the growth of heterotrophic bacteria. This power is proportional to the oxygen consumption rate by the 'biological specific heat yield' Y_{heat} (kJ/kg °C). Within the thermophilic temperature range there is no nitrification and the only oxygen consumption is due to the aerobic degradation of organic matter.

$$Pw_{bio} = Y_{heat}\left(\frac{(1 - Y_H)}{Y_H}\right)\rho_2 V_{liq} \quad [kJ/day]. \tag{6}$$

- Pw_{ve} is the power lost in the liquid phase due to water evaporation. This term is evaluated as the mass flow rate of water evaporated multiplied by the vaporization heat (ΔH_{vap}).

$$Pw_{ve} = (\Delta H_{vap}\rho_4)V_{gas} \quad [kJ/day]. \tag{7}$$

- Pw_{mi} is the power added by the mixing and aeration systems. It is calculated as the power consumed by the equipments ($Pw_{consumed}$) multiplied by the coefficient of heat utilization (η_m).

$$Pw_{mi} = Pw_{consumed}\eta_m \quad [kJ/day]. \tag{8}$$

- $Pw_{we\ liq}$ is the power lost through the reactor walls by conduction.

$$Pw_{we\ liq} = K_{we\ liq}(T_{liq} - T_{ambient})\text{Area}_{liq} \quad [kJ/day], \tag{9}$$

where $K_{we\ liq}$ is the coefficient of heat conduction through the walls (kJ/°C m^2 day), $T_{ambient}$ is the ambient temperature and Area_{liq} is the area of the reactor that surrounds the liquid phase (m^2).

- $Pw_{phases}^{conduction}$ is the rate of heat conduction through the surfaces of the liquid and gaseous volumes.

$$Pw_{phases}^{conduction} = k_Q a(T_{liq} - T_{gas}) \quad [kJ/day], \tag{10}$$

where $k_Q a$ is the heat transfer coefficient (kJ/°C day) which depends on the area between phases and the effect of the reactor systems.

The basic equation for the heat balance within the boundaries of the gaseous phase is

$$\frac{dH_{net\ gas}}{dt} = \text{Input}(H_{gas}) - \text{Output}(H_{gas}) - Pw_{we\ gas} + Pw_{phases}^{conduction} \quad [kJ/day], \tag{11}$$

Table 2 – Stoichiometric matrix for the ATAD model

↓Transformation kinetics (ρ_j)	Components $X_i \rightarrow$	1 S_{H_2O}	2 S_s	3 S_I	4 S_{O_2}	5 S_{HCO_3}	6 S_{CO_2}	7 S_{N_2}	8 $S_{NH_4^+}$	9 S_{NH_3}	10 S_{H^+}
1. *Hidrolysis of particulate substrate*											
$K_H\left(\frac{S_O}{S_O+K_{OXI}}+\eta\frac{K_{OXI}}{S_O+K_{OXI}}\right)\frac{X_S/X_{BH}}{K_X+X_S/X_{BH}}X_{BH}$		$\upsilon_{1,1}$	1		$\upsilon_{4,1}$		$\upsilon_{6,1}$		$\upsilon_{8,1}$		$\upsilon_{10,1}$
2. *Growth of heterotrophs*											
$\mu_H\cdot\frac{S_S}{K_S+S_S}\cdot\frac{S_O}{S_O+K_{OXI}}X_{BH}$		$\upsilon_{1,2}$	$-1/Y_H$		$\upsilon_{4,2}$		$\upsilon_{6,2}$		$\upsilon_{8,2}$		$\upsilon_{10,2}$
3. *Biomass lysis*											
b_HX_{BH}		$\upsilon_{1,3}$			$\upsilon_{4,3}$		$\upsilon_{6,3}$		$\upsilon_{8,3}$		$\upsilon_{10,3}$
4. *Water evaporation*											
$k_Ma\left(C^{SAT}_{H_2O}(T)-C_{H_2O}\right)$		$\upsilon_{1,4}$			$\upsilon_{4,4}$		$\upsilon_{6,4}$		$\upsilon_{8,4}$		$\upsilon_{10,4}$
5. *Stripping-dissolution of CO_2*											
$k_La_{CO_2}\left(S_{CO_2}-MW_CH_{CO_2}(T)\cdot Pp_{CO_2}\right)$		$\upsilon_{1,5}$			$\upsilon_{4,5}$		$\upsilon_{6,5}$		$\upsilon_{8,5}$		$\upsilon_{10,5}$
6. *Stripping-dissolution of O_2*											
$k_La_{O_2}\left(S_{O_2}-MW_{O_2}H_{O_2}(T)Pp_{O_2}\right)$		$\upsilon_{1,6}$			$\upsilon_{4,6}$		$\upsilon_{6,6}$		$\upsilon_{8,6}$		$\upsilon_{10,6}$
7. *Stripping-dissolution of N_2*											
$k_La_{N_2}\left(S_{N_2}-MW_{N_2}H_{N_2}(T)Pp_{N_2}\right)$		$\upsilon_{1,7}$			$\upsilon_{4,7}$		$\upsilon_{6,7}$	-1	$\upsilon_{8,7}$		$\upsilon_{10,7}$
8. *Inorganic C acid-base*											
$K^{CO_2}_{dissociation}\left(S_{HCO_3}\cdot S_H-K_{eq(CO_2)}\cdot S_{CO_2}\right)$		$\upsilon_{1,8}$			$\upsilon_{4,8}$	-1	$\upsilon_{6,8}$		$\upsilon_{8,8}$		$\upsilon_{10,8}$
9. *Inorganic N acid-base*											
$K^{NH_4}_{dissociation}\left(S_{NH_3}S_H-K_{eq(NH_4)}S_{NH_4}\right)$		$\upsilon_{1,9}$			$\upsilon_{4,9}$		$\upsilon_{6,9}$		$\upsilon_{8,9}$	-1	$\upsilon_{10,9}$
10. *Inorganic P acid-base*											
$K^{H_2PO_4}_{dissociation}\left(S_{HPO_4}S_H-K_{eq(H_2PO_4)}S_{H_2PO_4}\right)$		$\upsilon_{1,10}$			$\upsilon_{4,10}$		$\upsilon_{6,10}$		$\upsilon_{8,10}$		$\upsilon_{10,10}$
11. *Equilibrium water*											
$K^{H_2O}_{dissociation}\left(S_{OH}S_H-K_{eq(H_2O)}\right)$		$\upsilon_{1,11}$			$\upsilon_{4,11}$		$\upsilon_{6,11}$		$\upsilon_{8,11}$		$\upsilon_{10,11}$
		Water (kg H_2O)	Readily biod. subst. (kg COD)	Soluble inert org. matter (kg COD)	Dissolved O_2 (kg O)	Bicarbonate HCO_3^- (kg C)	Carbon-dioxide (kg C)	Dissolved N gas (kg N)	Ammonium (kg N)	Ammonia (kg N)	Ion hydrogen (kg H)

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↓Transformation kinetics (ρ_j)	Components $X_i \rightarrow$	11 S_{OH^-}	12 $S_{H_2PO_4^-}$	13 $S_{HPO_4^-}$	14 X_S	15 X_{BH}	16 X_I	17 G_{H_2O}	18 G_{CO_2}	19 G_{O_2}	20 G_{N_2}
1. *Hidrolysis of particulate substrate*											
$K_H\left(\frac{S_O}{S_O+K_{OXI}}+\eta\frac{K_{OXI}}{S_O+K_{OXI}}\right)\frac{X_S/X_{BH}}{K_X+X_S/X_{BH}}X_{BH}$			$\upsilon_{12,1}$		-1						
2. *Growth of heterotrophs*											
$\mu_H\cdot\frac{S_S}{K_S+S_S}\cdot\frac{S_O}{S_O+K_{OXI}}X_{BH}$			$\upsilon_{12,2}$			1					
3. *Biomass lysis*											
$b_H X_{BH}$			$\upsilon_{12,1}$		$1-f_{xi}$	-1	f_{xi}				
4. *Water evaporation*											
$k_M a\left(G^{SAT}_{H_2O}(T)-G_{H_2O}\right)$			$\upsilon_{12,4}$					1			
5. *Stripping-dissolution of CO_2*											
$k_L a_{CO_2}\left(S_{CO_2}-MW_C H_{CO_2}(T)\cdot Pp_{CO_2}\right)$			$\upsilon_{12,5}$						1		
6. *Stripping-dissolution of O_2*											
$k_L a_{O_2}\left(S_{O_2}-MW_{O_2}H_{O_2}(T)Pp_{O_2}\right)$			$\upsilon_{12,6}$							1	
7. *Stripping-dissolution of N_2*											
$k_L a_{N_2}\left(S_{N_2}-MW_{N_2}H_{N_2}(T)Pp_{N_2}\right)$			$\upsilon_{12,7}$								1
8. *Inorganic C acid-base*											
$K^{CO_2}_{dissociation}\left(S_{HCO_3}\cdot S_H-K_{eq(CO_2)}\cdot S_{CO_2}\right)$			$\upsilon_{12,8}$								
9. *Inorganic N acid-base*											
$K^{NH_4}_{dissociation}\left(S_{NH_3}S_H-K_{eq(NH_4)}S_{NH_4}\right)$			$\upsilon_{12,9}$								
10. *Inorganic P acid-base*											
$K^{H_2PO_4}_{dissociation}\left(S_{HPO_4}S_H-K_{eq(H_2PO_4)}S_{H_2PO_4}\right)$			$\upsilon_{12,10}$	-1							
11. *Equilibrium water*											
$K^{H_2O}_{dissociation}\left(S_{OH}S_H-K_{eq(H_2O)}\right)$		-1	$\upsilon_{12,11}$								
		Ion hydroxil (kg H)	Hydrogen-phosphate (kg P)	Dehydrogen-phosphate (kg P)	Slowly biod.subst. (kg COD)	Active hetero. biomass (kg COD)	Part. inert org. matter (kg COD)	Water vapour (kg H_2O)	Carbon-dioxide gas (kg C)	O_2 gas (kg O)	N gas (kg N)

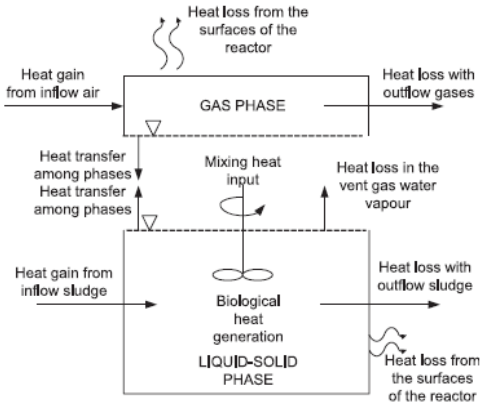

Fig. 1 – Terms of the heat balance in the liquid and gaseous phase.

where

- Input(H_{gas}) and Output(H_{gas}) are the values of power associated with the inflow gas at a temperature $T_{inf(g)}$ and the outflow gas at the temperature of the gaseous phase T_{gas}.

$$\text{Input}(H_{gas}) = m_{\text{inf gas}} C_{p\ \text{inf gas}} T_{\text{inf}(g)} \quad [\text{kJ/day}],$$
$$\text{Output}(H_{gas}) = m_{\text{out gas}} C_{p\ \text{out gas}} T_{gas} \quad [\text{kJ/day}], \tag{12}$$

  where $m_{\text{inf gas}}$ and $m_{\text{out gas}}$ are the mass inflow and the mass outflow (kg/day). $C_{p\ \text{inf gas}}$ and $C_{p\ \text{out gas}}$ are the specific heat (kJ/kg °C) of the inflow and outflow gases calculated from the specific heat and the partial pressure of each gas component.
- $Pw_{\text{we gas}}$ is the power lost through walls by conduction.

$$Pw_{\text{we gas}} = K_{\text{we gas}}(T_{gas} - T_{\text{ambient}})\text{Area}_{gas}, \tag{13}$$

  where $K_{\text{we gas}}$ is the heat conduction coefficient through the walls of the reactor (kJ/°C m^2 day) and Area_{gas} is the surface area of the reactor that surrounds the gaseous phase (m^2).
- $Pw^{\text{conduction}}_{\text{phases}}$. This term has the same value as the corresponding term in the liquid phase (Eq. (10)) but with the opposite positive or negative sign.

The temperature T_{liq} and T_{gas} in time t are calculated as

$$T_{liq}(t) = \frac{H_{\text{net liq}}(t)}{M_{liq}(t) C_{p\ \text{liq}}(t)} \quad [°\text{C}], \tag{14}$$

$$T_{gas}(t) = \frac{H_{\text{net gas}}(t)}{M_{gas}(t) C_{p\ \text{gas}}(t)} \quad [°\text{C}], \tag{15}$$

where $M_{liq}(t)$ and $C_{p\ \text{liq}}(t)$ are the mass and specific heat of the water state S_{H_2O}.

$$M_{gas}(t) = \sum_{i=\text{gas}} M_i(t)$$

is the total mass of gas in the reactor and

$$C_{p\ \text{gas}}(t) = \sum_{i=\text{gas}} \frac{C_{p_i} M_i(t)}{M_{gas}(t)}.$$

2.3. General parameters proposed by default

Table 3 summarizes the default values proposed for some stoichiometric, kinetic, physico-chemical and thermal general parameters of the model (Table 2).

Stoichiometric parameters (Y_H, f_{xi}): Assuming a similar biomass elemental composition at mesophilic and thermophilic temperatures, similar values of theoretical yield have been measured at both ranges by Vogelaar et al. (2003).

Kinetic parameters (μ_H, b_H): There is some controversy regarding these kinetic parameters under thermophilic conditions. The values given in the literature depend on the method of measurement. Substrate utilization rates reported in thermophilic ranges are 3–10 times larger than in mesophilic processes. The values chosen for this work are those given in Vogelaar et al. (2003).

Thermal parameters (Y_{heat}, $\Delta H_{vap}(T)$): The exothermic process of aerobic biological oxidation of the organic matter produces 13.9 MJ/kg $O_{2\ \text{consumed}}$ (Gomez et al., 2007).

The enthalpy associated with the water evaporation processes ($\Delta H_{vap}(T)$) has been computed as

$$\Delta H_{vap}(T) = \Delta H^0_{vap} - C_P(H_2O(l))(T - 100\,°\text{C}) - C_P(H_2O(g))(100\,°\text{C} - T) \quad [\text{MJ/kg H}_2\text{O}], \tag{16}$$

where T is the temperature of the liquid phase (°C) and $C_p(H_2O(l))$ and $C_p(H_2O(g))$ are the specific heat of water and water vapour (MJ/kg H_2O °C), respectively.

3. Practical application

3.1. Characterization of inflow sludge and gas in terms of the model components

A key-factor for the successful application of a mathematical model of this type is good characterization of the organic waste (Huete et al., 2006). In the particular case of the ATAD model, it is also necessary to characterize the incoming gas.

The organic waste (sewage sludge) must be characterized by the analytical measurements periodically performed. The viable biomass content in the inflow sludge is considered to be zero under thermophilic conditions. Biodegradability tests indicate the amount of biodegradable (X_S and S_S) and inert components (X_I and S_I) in the filtered and particulate COD. The component S_{HCO_3} can be obtained from the experimental measurement of bicarbonate alkalinity, whereas S_{CO_2} is calculated from the acid/base equilibrium reaction at the influent pH. The experimentally measured N–NH_4 concentration represents the total inorganic nitrogen, which is the sum of the S_{NH_3} and S_{NH_4} components of the model in equilibrium at the influent pH.

The value of the concentration of the gaseous components present in the incoming gases can be obtained from the dry composition, temperature, pressure and humidity of the air.

3.2. Specific parameters for a particular case

Certain parameters have to be evaluated for each particular reactor. This calibration makes the model application for any particular digester possible.

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Table 3 – Values proposed by default for the general parameters of the ATAD model

	Parameter	Value	Source
Stochiometric			
Y_H	Yield for heterotrophic biomass (g COD/g COD)	0.63	ASM2 (Henze et al., 2000)
f_{XI}	Fraction of inert COD generated by lysis (g COD/g COD)	0.1	ASM2 (Henze et al., 2000)
Kinetic			
K_H	Maximum specific hydrolysis rate (day^{-1})	4	By default. Need of calibration
K_{OXI}	Saturation/Inhibition coefficient for oxygen (kg O_2/m^3)	0.0002	ASM1 (Henze et al., 2000)
η	Anaerobic hydrolysis reduction factor (dimensionless)	0.4	ASM2d (Henze et al., 2000)
K_X	Half-saturation coefficient for hydrolysis (g COD/g COD)	0.03	ASM1 (Henze et al., 2000)
K_S	Half-saturation coefficient for growth (kg COD/m^3)	0.02	ASM1 (Henze et al., 2000)
μ_H	Maximum specific growth rate (day^{-1})	17.05	Vogelaar et al. (2003)
b_H	Lysis rate coefficient (day^{-1})	0.94	Vogelaar et al. (2003)
Physico-chemical			
$H_{CO_2}(T)$	Henry constant of CO_2 (kmol/m^3 atm)	$0.0345e^{-2400((1/298.15)-(1/T(^\circ K)))}$	Lide and Frederikse (1995)
$H_{O_2}(T)$	Henry constant of O_2 (kmol/m^3 atm)	$0.00128e^{-1500((1/298.15)-(1/T(^\circ K)))}$	Lide and Frederikse (1995)
$H_{N_2}(T)$	Henry constant of N_2 (kmol/m^3 atm)	$0.000602e^{-1300((1/298.15)-(1/T(^\circ K)))}$	Kavanaugh and Trussell (1980)
$K^{CO_2}_{dissociation}$	Eq. rate constant CO_2/HCO_3	10^7	Big enough for immediate reaction
$K_{eq(CO_2)}$	Acidity const. CO_2/HCO_3 (kg H/m^3)	$10^{-6.35}e^{(7646/8.314)((1/298.15)-(1/T(^\circ K)))}$	Lide and Frederikse (1995)
$K^{NH_4}_{dissociation}$	Eq. rate constant NH_4/NH_3 (m^3/kg H d)	10^7	Big enough for immediate reaction
$K_{eq(NH_4)}$	Acidity const. NH_4/NH_3 (kg H/m^3)	$10^{-9.25}e^{(51965/8.314)((1/298.15)-(1/T(^\circ K)))}$	Lide and Frederikse (1995)
$K^{H_2PO_4}_{dissociation}$	Eq. rate const. H_2PO_4/HPO_4 (m^3/kg H d)	10^7	Big enough for immediate reaction
$K_{eq(H_2PO_4)}$	Acidity const. H_2PO_4/HPO_4 (kg H/m^3)	$10^{((-219.4/T(^\circ K))-6.46)}$	Lide and Frederikse (1995)
$K^{H_2O}_{dissociation}$	Eq. rate const. H^+/OH^- (m^3/kg H d)	10^9	Big enough for immediate reaction
$K_{eq(H_2O)}$	Acidity const. H^+/OH^- (kg H^2/m^6)	$10^{-14}e^{(55\,900/8.314)((1/298.15)-(1/T(^\circ K)))}$	Lide and Frederikse (1995)
$G^{SAT}_{H_2O}$	Water vapour sat. (kg H_2O/m^3)	$0.0313e^{(5290((1/298.15)-(1/T_{gas})))(18/(T_{gas}R))}$	Lide and Frederikse (1995)
Thermal			
Y_{heat}	Specific heat yield (MJ/kg $O_{2\,consumed}$)	13.9	Gomez et al. (2007)
ΔH^0_{vap}	Vaporization enthalpy at 100 °C (MJ/kg H_2O)	2.26	

The value of parameters $k_M a$, k_{gas} and $k_Q a$ can be obtained based on the experimental measurement of the humidity and pressure in the gaseous phase, and the temperature gap between both phases. The coefficients for heat conduction through the walls (K_{we}) and the power supply by the reactor equipments depend on the particular reactor. The value of $k_L a_{O_2}$, depends on the aeration systems, air feed rate, temperature, reactor design and specific diffusion characteristics. The values of $k_L a_{CO_2}$ and $k_L a_{N_2}$ are related with the value of $k_L a_{O_2}$ through the diffusivity of the gases (D_{gas}).

$$k_L a_{gas} = k_L a_{O_2}\sqrt{\frac{D_{gas}}{D_{O_2}}} \quad [\mathrm{kJ/day}], \tag{17}$$

where $D_{N_2} = 1.64 \times 10^{-4}$, $D_{O_2} = 2.16 \times 10^{-4}$ and $D_{CO_2} = 1.69 \times 10^{-4}\,m^2/day$ (Lide and Frederikse, 1995; Perry and Chilton, 1973).

4. Exploration of two different scenarios by simulation

At present, a description of the process behaviour under different operational conditions can be easily explored with this model by simulation and compared with existing results in the literature (Gomez et al., 2007; Garcia et al., 2007). Thus, an initial qualitative assessment of the model's predictive capacity has been made. The model structure and component interactions within the model have been also reflected. In other words, a preliminary model verification has been carried out (Olsson and Newell, 1999).

The ATAD reactor performance has been explored by simulation in two different scenarios to investigate the behaviour of the model in different situations: acting as a first stage in a Dual system (ATAD+Mesophilic Anaerobic Digestion in series, Scenario 1) and acting as a single stage sludge treatment (Scenario 2). The operational conditions in

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Table 4 – Values for specific parameters of the ATAD model used in the exploration

	Parameter	Value	Source
$K_M a$	Water evaporation rate coefficient (day^{-1})	100	By default. Need of calibration
K_{gas}	Gas flow rate coefficient (m^3/day atm)	80 000	Big enough for immediate reaction
$K_Q a$	Heat transfer kinetic coeff. (kJ/h m^2 °C)	160 000	By default. Need of calibration
$K_{we\ liq}$	Heat conduction coefficient through walls (kJ/°C m^2 day)	25	Gomez et al. (2007)
$K_{we\ gas}$	Heat conduction coefficient through walls (kJ/°C m^2 day)	25	Gomez et al. (2007)
$k_L a_{O_2}$	O_2 gas transfer coefficient (day^{-1})	1000	By default. Need of calibration
$Pw_{consumed}$	Power consumed by the equipment (kW)	30	Gomez et al. (2007)
η_m	Coefficient of heat utilization (dimensionless)	1	Gomez et al. (2007)

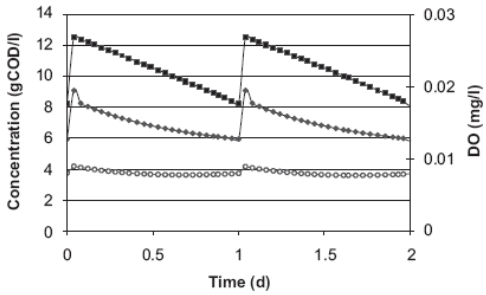

Fig. 2 – Evolution of X_S (■), S_S (○) and dissolved oxygen (◆) in scenario 1 with Q_{air} = 10 000 m^3/h.

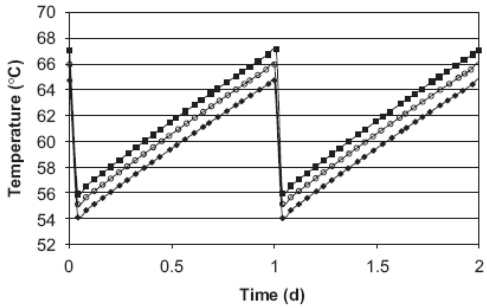

Fig. 3 – Evolution of temperature with Q_{air} values of 11 000 (■), 10 000 (○) and 9000 (◆) (m^3/h) in scenario 1.

both scenarios are quite different but in both cases the reactor operates in a 23-h semi-batch pattern with an effective sludge volume of 300 m^3. The values for the specific parameters chosen for these examples are given in Table 4.

The $k_L a$ values have been considered independent from the temperature fluctuations. The non-biodegradable fraction of the raw sewage sludge particulate and soluble COD has a value of 40% for these simulations.

4.1. *Scenario 1: ATAD with a 5-day hydraulic retention time as the first stage in a dual system*

In this first scenario the ATAD aims to maintain the sludge for a long enough period of time at the hygienization temperature with minimal organic matter removal. This implies that the reactor is oxygen limited (dissolved oxygen nearly zero) and has an excess of readily biodegradable substrate (S_S) (Fig. 2). Under such conditions the kinetics of the heterotrophic biomass growth is modulated by the dissolved oxygen concentration, and the temperature ascent rate is directly controlled by the oxygen feed rate (Fig. 3) up to the oxygen uptake rate capacity (Messenger et al., 1990). If the air feed rate is maintained constant, the temperature ascent rate will also be constant throughout the batch period provided there is enough concentration of S_S. The scenario was simulated by treating sewage sludge with 60 g total COD/l and 10 g filtered COD/l.

A qualitative sensitivity analysis of most parameters has been carried out (graphs not included) leading to the conclusion that in this first scenario the system shows no sensitivity to the μ_H, b_H and K_{OXI} values. The value of the hydrolysis rate coefficient (K_H) significantly affects the soluble and particulate biodegradable effluent COD (S_S and X_S). The sensitivity analysis also showed that, as expected, the organic matter removal rate is directly controlled by the oxygen supply rate, i.e., by the aeration flow rate and the $k_L a_{O_2}$ value. Whilst the former is an operational parameter, the latter depends on the aeration system and must be assessed for each particular case.

4.2. *Scenario 2: ATAD as a single sludge treatment (10-day hydraulic retention time)*

In this case, the reactor aims both to maintain the hygienization requirements and to achieve the maximum removal of organic matter (stabilization). Hence, the kinetics of heterotrophic biomass growth is also limited by the lack of readily biodegradable organic matter (S_S). The results of the simulations presented here have been obtained with an inflow concentration of sludge of 30 g total COD/l and 5 g filtered COD/l.

In this scenario the S_S concentration in the effluent is zero and the organic matter removal is the maximum. Fig. 4 shows that the temperature ascent rate is variable throughout the cycle. The first slope (0–A) is controlled by the oxygen supply. When the reactor runs out of S_S the hydrolysis becomes the limiting step (A–B); therefore this second slope is controlled by the hydrolysis coefficient (K_H). In this stage the dissolved oxygen level increases (Fig. 5) due to the oxygen consumption rate drop. In the last stage (B–C) the lysis rate coefficient (b_H)

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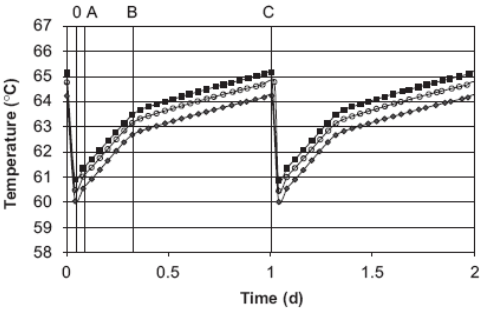

Fig. 4 – Evolution of temperature with Q_{air} values of 11 000 (■), 10 000 (○) and 9000 (◆) (m^3/h) in scenario 2.

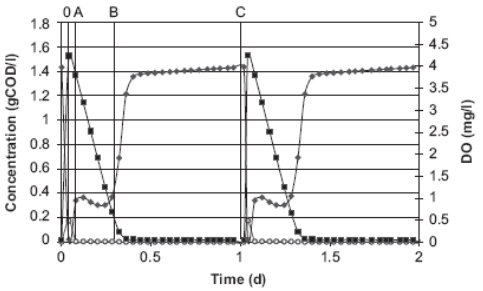

Fig. 5 – Evolution of X_S (■), S_S (○) and dissolved oxygen (◆) in scenario 2 with Q_{air} = 10 000 m^3/h.

becomes the limiting rate since the particulate substrate is exhausted. In this scenario, the biological kinetic parameters K_H and b_H are the key values needed to predict the oxygen consumption rate and the biological heating in the reactor. Hence, a future calibration of these parameters is recommended in order to obtain accurate model predictions for these operational conditions.

The value obtained for the respiration coefficient (mol CO_2 generated per mol O_2 consumed) in the simulations was 0.89. This value depends on the substrate elemental mass composition and is below 1 due to the fact that in the lysis of biomass part of the carbon returns as particulate substrate instead of finishing up as CO_2.

Currently, several points in the model are being studied in order to improve its predictive capacity. The addition of another hydrolysis transformation will enable different kinetics for the hydrolysis of primary and secondary sludge to be incorporated. In addition, a thermal hydrolysis will be included to simulate the effect of temperature shock over the sewage sludge. It would be also useful to consider the influence of temperature in certain kinetic parameters to predict the effect of changes in the operational temperature. The addition of anaerobic degradation processes and partial oxidations seems necessary for a good prediction of the respiration coefficient (mol $CO_{2\ generated}$ per mol $O_{2\ consumed}$) at low levels of dissolved oxygen in the digester.

5. Conclusions

A new mathematical model for the dynamic description of Autothermal Thermophilic Aerobic Digesters has been developed. The proposed model is able to simulate the main biochemical transformations of the process, guaranteeing a rigorous continuity for mass and energy. The flexible modelling structure facilitates both the modification of the set of transformations and the easy connection with other unit process models so as to simulate more complex plant configurations.

In order to facilitate the practical model implementation, some default values and general rules for the selection of the most relevant model parameters have been suggested.

Two scenarios have been explored as an example by simulation: (1) where the ATAD acts as a first stage in a Dual system and (2) where it acts as a single stage sludge treatment. The exploration of both scenarios has been carried out to investigate the behaviour of the model in different situations and to prove the model's capacity to qualitatively predict the process performance. The relevant parameters, which are not the same for each scenario, have been identified enabling the comprehension of the process behaviour working under two different operational conditions.

There are some model parameters that are specific for each reactor; therefore, a proper calibration is recommended to make the model application for any particular scenario possible. Currently, a calibration and validation of the proposed model is being carried out by using the experimental results obtained in a full scale ATAD reactor.

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Appendix A. Supplementary materials

Supplementary data associated with this article can be found in the online version at doi:10.1016/j.watres.2006.11.042.

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Benchmark Simulation Model No 2 in Matlab-Simulink: towards plant-wide WWTP control strategy evaluation

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Abstract In this paper, implementation of the Benchmark Simulation Model No 2 (BSM2) within Matlab-Simulink is presented. The BSM2 is developed for plant-wide WWTP control strategy evaluation on a long-term basis. It consists of a pre-treatment process, an activated sludge process and sludge treatment processes. Extended evaluation criteria are proposed for plant-wide control strategy assessment. Default open-loop and closed-loop strategies are also proposed to be used as references with which to compare other control strategies. Simulations indicate that the BM2 is an appropriate tool for plant-wide control strategy evaluation.

Keywords Benchmarking; BSM2; control; evaluation criteria; modelling; simulation; wastewater treatment

Introduction

The COST/IWA Benchmark Simulation Model No 1 (BSM1) (Copp, 2002) is widely used all over the world within the research community for testing and evaluating various control strategies for biological nitrogen removal wastewater treatment plants. Although the BSM1 is a valuable tool, it does not allow for evaluation of control strategies on a longer time scale and on a plant-wide basis. Therefore, extensions of the benchmark system for long-term control and monitoring system performance evaluation (Rosen *et al.*, 2004) and for plant-wide control (Jeppsson *et al.*, 2006) have been proposed recently. A model for generating dynamic influent disturbance profiles, which is an essential module for long-term control system evaluation, has also been suggested (Gernaey *et al.*, 2006). The benchmark system is extended with the state-of-the-art anaerobic digestion model No 1 (ADM1) (Batstone *et al.*, 2002). To achieve reasonable simulation times with dynamic influent and with active measurement noise the ADM1 had to be modified with algebraic solvers for the pH and hydrogen (S_{h2}) states (Rosen *et al.*, 2006). Based on these proposals the Benchmark Simulation Model No 2 (BSM2) has been developed within the Matlab-Simulink software. In this paper, a description of the BSM2 implementation is provided, and some simulation results using a proposed default control strategy are presented and evaluated based on a set of extended evaluation criteria.

The paper is organised as follows. In the following section, the Matlab-Simulink implementation of BSM2 is described. Next, simulation results for the open-loop and closed-loop control strategies are shown and are evaluated by the extended BSM2 evaluation criteria. At the end some conclusions are drawn.

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Matlab-Simulink implementation of BSM2

The layout of the BSM2 is shown in Figure 1. All the models used in Simulink are written as C MEX file S-Functions. Such an implementation is elegant and allows for considerably higher simulation speed compared to m-file S-Functions. The BSM2 contains an influent wastewater, the pre-treatment process with a primary clarifier, the activated sludge (AS) process with a secondary clarifier (settler) and the sludge treatment processes with thickener unit, anaerobic digester (AD) and dewatering unit. All these components are connected in Simulink by combiner and splitter model blocks. There are also two extra blocks, the controller block that contains selected controllers and the monitoring block, which is mainly used for representing control results. These two blocks are elegantly connected to the process by using Simulink tags for sending and receiving signals.

State vector

The state vector is extended to 21 components: 13 ASM1 states, total suspended solids (TSS), flow rate, temperature and five dummy states (three soluble and two particulate). The dummy states are added to allow for easy future expansion of the system with additional states, for example to model inhibition of biological processes.

Activated sludge process

The model of the activated sludge (AS) process that is used in BSM2 is the same as the one defined for BSM1 (Copp, 2002). The only difference is that the kinetic parameters of the Activated Sludge Model No 1 (ASM1; Henze *et al.*, 1987) are now temperature dependent.

Secondary clarifier (settler)

The model of the secondary clarifier is the same as in BSM1. It consists of a 10-layer one-dimensional non-reactive settler model using the Takács double-exponential settling function (Takács *et al.*, 1991). The settling properties of the sludge in the secondary clarifier are not temperature dependent.

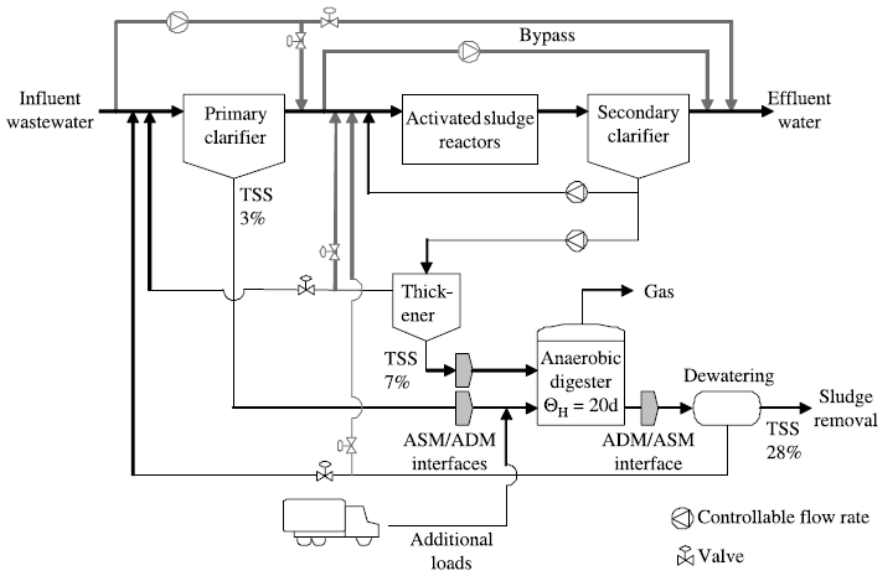

Figure 1 Layout of the BSM2

Primary clarifier

The input to the primary clarifier combines the influent to the treatment plant, the reject water stream resulting from the sludge dewatering and the recycle from the thickener overflow. The underflow rate of the primary clarifier is controlled proportionally to the incoming flow. The primary clarifier is implemented using the model proposed by Otterpohl and Freund (1992). The volume of the primary clarifier and the parameters of the clarifier are set so as to obtain a TSS concentration in the sludge stream of about 30,000 g/m^3 and a TSS removal efficiency of about 50% (Jeppsson *et al.*, 2006).

Thickener and dewatering unit

Ideal thickener and dewatering units are used, where 98% of all particulate matter entering the units concentrates in the sludge stream. The TSS concentration in the sludge stream of the thickener unit is set to 70,000 g SS/m^3, and for the dewatering unit to 280,000 g SS/m^3. In order to achieve the defined TSS concentration in the sludge stream the underflow flow rate is changing instantaneously.

Anaerobic digester

The anaerobic digester model No. 1 (ADM1) (Batstone *et al.*, 2002) is used for modelling the anaerobic digester process. The model has been implemented within Matlab-Simulink aimed at obtaining fast simulation speed (see next subsection). The input of the anaerobic digester combines the sludge output from the thickener unit, the sludge output from the primary clarifier and additional loads with concentrated waste reaching the plant sporadically.

ADM1 model implementation

Applicability of the BSM2 largely depends on simulation speed. To implement the ADM1 model to obtain reasonable simulation times of the BSM2 even with dynamic influent and measurement noise, the stiffness of the ADM1 model has to be reduced. This is done by approximating the differential equations of the pH and liquid hydrogen S_{h2} states of the ADM1 by algebraic equations, and solving them with an iterative numerical method (Rosen *et al.*, 2006). The differences that are introduced by applying these solvers can be neglected.

ASM1–ADM1 interfaces

Since the state variables of the ADM1 are not the same as in the ASM1, interfaces are needed to connect these two processes. The interfaces proposed by Copp *et al.* (2003) are applied. As the sludge generated in the primary clarifier (primary sludge) and the thickener unit (secondary sludge) differs in composition and characteristics, two interfaces having somewhat different parameters are used.

Temperature dependence

Only temperature dependence of the AS kinetic model parameters is considered. The temperature dependence of these parameters is modelled using an Arrhenius function. At 15 °C BSM2 AS parameters have exactly the same values as in BSM1 for reasons of compatibility. Temperature dynamics in the processes are described by simple temperature mass balance equations.

Hydraulic delays

At the input of the primary clarifier and AS process, hydraulic delays have been added to prevent algebraic loops (a result of wastewater recycling). The hydraulic delays are

modelled in the same way as in BSM1 (Copp, 2002), simply by filtering the model states with a fast first-order filter.

Combiners and splitters

The combiners and splitters that are applied are the same as those used in BSM1. Combiners calculate the output concentrations by summing up the influx of each input stream for each pollutant, and subsequently dividing the resulting total flux by the total flow rate to obtain the output concentrations. On the other hand the splitters divide the input stream into two streams with identical concentrations but different flow rates.

Influent

The influent data are obtained from the influent model proposed by Gernaey *et al.* (2006), although the nitrogen load is reduced by 15% in order to obtain a plant, which is not significantly overloaded in nitrogen during the winter. The data represent a dynamic influent including diurnal, weekend, seasonal and holiday effects as well as rainfall, sludge settling in the sewer system, first-flush events etc. The temperature trajectory in the influent is composed of a sinusoidal function with a period of one year to model seasonal temperature variations, with a second sinusoidal function superimposed on it (period of one day) to model small variations of the temperature between day and night.

Additional loads

Additional loads from outside sources are also considered in the BSM2. These loads are considered to be added directly into the AD. However, they are not used in this study.

Simulation results

Simulation procedure

The BSM2 is first simulated with a constant influent for 200 days to reach its steady-state. The values of the influent soluble components and the flow rate are the same as those used in BSM1, whereas the values of the influent particulate components are doubled. The temperature, which is also included in the constant influent, is 15 °C. The steady-state values obtained in this first simulation are subsequently used as initial values for simulation with the dynamic influent. The BSM2 is then simulated with the dynamic influent for 63 days (9 weeks, from $t = 0$ d to $t = 63$ d) so that a quasi steady state is reached. These 63 days are followed by 182 days of dynamic simulation (26 weeks, from $t = 63$ d to $t = 245$ d) in order to get dynamic data, which are intended to be used for training monitoring and control strategies. The start date for this period of 26 weeks corresponds to January 1st. Finally, the BSM2 is simulated for an additional 364 days (52 weeks, starting at $t = 245$ d, ending at $t = 609$ d) and the data obtained during this period are used for plant performance and monitoring strategy performance evaluation. The dynamic influent data for plant performance evaluation start on July 1st, and finish on June 30th the next year.

Default open-loop case

In the proposed open-loop case, constant control variables are used to obtain a reasonable behaviour of the plant for the whole year. The oxygen transfer rates for the last three tanks of the AS (K_La_3, K_La_4, K_La_5) are set to 240 d^{-1}, values of the internal (Q_{intr}) and external (Q_r) recycle flow rates of the AS are the same as defined for BSM1, the waste sludge flow rate (Q_w) is set to 300 m^3/d and the external carbon flow rate (Q_{carb}) is set to 2 m^3/d (800 kg COD/d). The possibility to bypass the influent wastewater was not used. Similarly, a recycle stream to the AS from the thickener and dewatering unit has not

been used. Selected daily averaged values of open-loop results for the evaluation period of one year are shown in Figure 2.

The ammonia removal efficiency in the AS process is highly correlated with the influent temperature. During the winter period, when the wastewater temperature values are at their lowest, the nitrification process is not as efficient compared to the summer period, and as a consequence higher effluent ammonia values are obtained during the winter period. Because of this, effluent total nitrogen values increase during the winter as well. The primary clarifier output concentration of the TSS represents around 50% of the TSS concentration at the influent (400 g SS/m^3). This means that approximately 50% of the TSS is removed in the primary clarifier, as was defined in the design phase. The whole plant, including the secondary clarifier, efficiently removes the TSS. The digester effluent ammonia concentration is high. Since the ammonia is recycled back to the primary clarifier it contributes significantly to the nitrogen load of the AS plant, even though the supernatant flow rate from the dewatering unit is small in comparison with the influent flow rate. This implies that special treatment of the supernatant from the dewatering may be a valid operational strategy. Such an approach is evaluated in Volcke *et al.* (2006).

Default closed-loop case

In the proposed default closed-loop case, simple proportional-integral (PI) and proportional (P) controllers are applied. Oxygen in the last aerobic tank (S_{O5}) is controlled with a PI controller by manipulating the K_La_5, whereas K_La_3 and K_La_4 are kept constant as in BSM1. Nitrate in the second anoxic tank (S_{NO2}) is controlled with a PI controller that uses the internal recycle flow (Q_{intr}) as the manipulated variable. The daily averaged total suspended solids concentration in the last aerobic tank (TSS_5) is controlled by a P controller that uses waste sludge flow rate (Q_w) as the manipulated variable. Parameters for the PI and P controllers were tuned from step response experiments using the internal model control (IMC) tuning rules (Olsson and Newell, 1999). Controller parameter values are given in Table 1. The set-point for the total suspended solids in the last aerobic tank ($TSS_{5\text{set}}$) is changing according to the weekly averaged wastewater temperature (T). The TSS set-point is set to 3,500 g SS/m^3 for temperatures above 15 °C and to 4,500 g SS/m^3 below 15 °C. Averaged values of TSS and T are approximately obtained by using

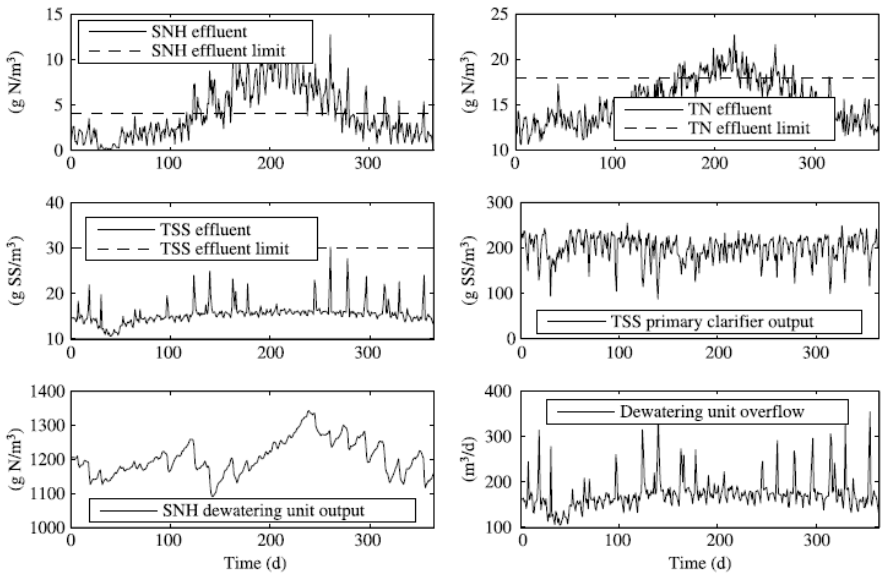

Figure 2 Daily averaged values obtained in open-loop simulation for the evaluation period

Table 1 Parameters used for controllers

	S_{O5}	S_{NO2}	TSS_5
K_P	$100\,d^{-1}\,(g(O_2)m^{-3})^{-1}$	$10{,}000\,m^3\,d^{-1}\,(gNm^{-3})^{-1}$	$-1\,m^3\,d^{-1}\,(gSSm^{-3})^{-1}$
T_i	0.01 d	0.04 d	–
T_t	0.01 d	0.04 d	–
Set-point	$2\,g\,O_2\,m^{-3}$	$1\,g\,N\,m^{-3}$	Variable

first-order filters with a time constant of one and seven days, respectively. Both external carbon flow (Q_{carb}) and external recycle flow (Q_r) are controlled proportional to the influent flow rate. Proportional constants were tuned manually to obtain a reasonable behaviour of the AS system (for Q_{carb} 0.00008 and for Q_r 1.5). The waste sludge flow rate (Q_w) is kept constant as in the open-loop case. Bypassing of the influent and recycling of the reject water to the AS is not used. In Figure 3, daily averaged values of effluent ammonia and total nitrogen, controlled and manipulated variables obtained in closed-loop simulation for the evaluation period are shown.

Effluent limit violations of ammonia and total nitrogen are much lower than in the open-loop case. Oxygen and nitrate PI controllers follow defined set-points so tightly that the difference between the set-points and true values cannot be distinguished in the graphs. This is not the case for the P control of *TSS*, where a noticeable but approximately constant offset from the set-point is obtained. It can be also noticed that the average external carbon flow rate is lower than the constant external carbon flow rate used in the open-loop case.

Evaluation criteria

Evaluation criteria are proposed for simple evaluation and comparison of control strategies for the BSM2. The evaluation criteria represent an extension of the criteria for BSM1 and consider the features of the new processes that are included in BSM2. The evaluation period has been extended from one week to one year. The effluent quality (*EQ*) is calculated the same as in the BSM1. However, for the operational cost index

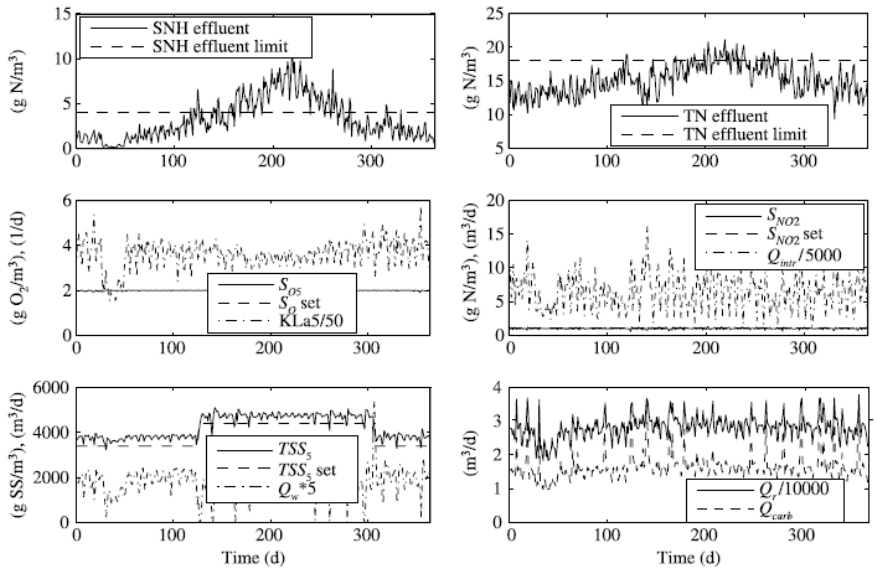

Figure 3 Daily averaged values obtained in closed-loop simulation for the evaluation period

Table 2 Values of evaluation criteria for default open-loop and closed-loop cases

Evaluation criteria	*EQ* (kg/d)	*OCI* (1/d)	*AE* (kWh/d)	*PE* (kWh/d)	*SP* (kgSS/d)	*EC* (kgCOD/d)	*ME* (kWh/d)	*MP* ($kgCH_4/d$)
Open-loop	8,847	18,967	8,548	2,978	3,216	800	648	876
Closed-loop	8,727	17,268	7,771	2,453	3,205	661	648	867

Evaluation criteria	HE^{net} (kWh/d)	TN_{e95} (g/m^3)	$S_{NH,e95}$ (g/m^3)	TSS_{e95} (g/m^3)	$T_{viol}TN_e$ (%)	$T_{viol}S_{NH,e}$ (%)	$T_{viol}TSS_e$ (%)
Open-loop	0	21.4	11.9	20.7	23.9	40.9	0.7
Closed-loop	0	20.5	10.1	21.7	19.5	33	0.7

(*OCI*), a modified version of the weighted sum of different costs is proposed:

$$OCI = AE + PE + 3 \cdot SP + 3 \cdot EC + ME - 6 \cdot MP + HE^{net}, \quad (1)$$

where *AE* is aeration energy, *PE* is pumping energy, *SP* is sludge production for disposal, *EC* is external carbon addition, *ME* is mixing energy, *MP* represents methane production in the AD and HE^{net} is a net heating energy for the AD. Aeration energy (*AE*) is calculated in a similar way as in the BSM1. External carbon addition (*EC*) is calculated the same as in BSM1. For the pumping energy calculation (*PE*) also the primary clarifier underflow, thickener underflow and dewatering overflow are included. Sludge production for disposal (SP) is calculated based on the amount of solids that are accumulated in the plant and from the solids that are removed from the plant as dewatered sludge. The weight for the *SP* has been reduced from the value 5, which was used in BSM1 (no sludge treatment explicitly included in BSM1 but the cost was still included in the *OCI*), to the value 3. Mixing energy (*ME*) combines energy used for mixing the AS anoxic tanks and energy used for mixing the AD. Methane production (*ME*) represents an economic benefit, and can be included in the cost index as a negative cost. The weight for the *MP* is set to 6 and it defines that around 50% of the energy content of the methane is converted into electricity using a gas motor. Net heating energy (HE^{net}) represents energy that is needed for heating the sludge of the AD in case the assumed heat exchange system together with the heat provided by the gas motor is not sufficient. Other criteria, such as time of effluent limit violations (T_{viol}) and 95 percentile of effluent variables, are calculated as in the BSM1. Values for the proposed evaluation criteria for the open-loop and closed-loop cases are given in Table 2.

In both cases approximately the same effluent quality (*EQ*) is obtained. However, in the closed-loop case the operating cost index (*OCI*) is about 10% lower. Moreover, also effluent limit violations of ammonia and total nitrogen are much lower. The net heating energy (HE^{net}) is in both cases zero, which means that the energy from the heat exchange system together with the heat generated by the gas motor is sufficient to heat the sludge of the AD.

Conclusions

A first version of the BSM2 has been implemented and tested within the Matlab-Simulink software. It consists of a pre-treatment process, an activated sludge process and the sludge treatment processes. For objective evaluation and comparison of plant-wide control strategies on a long-term basis, a set of extended evaluation criteria are proposed for BSM2. Assessment of simulation results in terms of evaluation criteria has shown that in the closed-loop example considerably lower operational costs and lower effluent nitrogen violations are achieved in comparison with the open-loop case. Simulations indicate that the BSM2 can be successfully used for plant-wide control strategy evaluation. It should

be noted that the BSM2 is still in its development phase by the IWA Task Group on Benchmarking of Control Strategies for WWTPs. Consequently, some details regarding the ADM1 model implementation, ASM1–ADM1 interfaces and evaluation criteria presented in this paper are likely to change in the future. The finalised version of the BSM2 will be presented in 2008.

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Effect of Atuothermal Thermophilic Aerobic Digestion operation on reactor temperatures

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Abstract: **A pilot scale facilities of Autothermal Thermophilic Aerobic Digestion (ATAD) was designed for stabilization of sewage sludge. The operating parameters that influence reactor temperature change were studied such as volatile solids (VS) concentration of feed sludge, hydraulic retention time (HRT), aeration rate, and mixing mechanical energy. The effect of reactor temperature on the destruction pathogens was also determined. This study found that: 1) the optimal initial VS concentration and HRT of the sludge were 30~40g/L and 10~15d respectively; 2) aeration rate has relevance to VS concentration and has important effect on reactor temperature rise and VS reduction. An excessive air flowrate resulted in a drop in temperature; 3) reactor can keep normal temperature in condition of circular flowrates at 12~14 m^3/h and mixing power density at 280~330W/m^3; 4) ORP values demonstrated that the ATAD process was an anaerobic environment first, and then microaerophilic. This study demonstrates that the ATAD process was not completely autothermal, but required a heat component from mechanical mixing energy to maintain reactor temperature. The temperature in ATAD tank was likely, the most important factor influencing the rate of pathogen inactivation.**

Key words: Autothermal thermophilic aerobic digestion; Temperature; Solids concentration of feed sludge; Hydraulic retention time; Aeration rate; Mixing mechanical energy; Pathogen inactivation

A large amount of sludge solids is generated from municipal sewage plants. Handling and use of these sludge solids can cause environmental and health problems, and need to be sanitized and stabilized. Unlike conventional aerobic digestion (CAD), autothermal thermophilic aerobic digestion (ATAD) can utilize heat released by an exothermic microbial oxidation process without an external heat input to achieve a thermophilic temperature (45~65^{0}C), and operate with relatively short detention time (6~12days) (*Kelly et al., 2003*). More recently the process has been promoted for its ability to produce Class A biosolids, therefore, it has been popularly constructed in Europe in considerable number since U.S. EPA enforced restrictions on pathogen destruction (*U. S. EPA, 1990*). As such the product has unrestricted reuse and thus wider marketing opportunities than unpasteurized biosolids.

It is common that two or more cylinder reactors in series form a plug flow for continuous or semi-continuous treatment. However, valves on pipes between multiple reactors and other equipment valves frequent opening and closing make reactor operations complex (*Kelly et al., 2003*). Use of one-stage instead of two-stage reactor not only simplifies operation, but saves space and reduces investment and running costs compared to two-stage reactors (*Skjelhaugen, 1999*). In addition, it is convenient to use existing facilities as an ATAD reactor. This paper described a self-designed, one-stage rectangle ATAD digestion reactor with recirculation pumps to sufficiently mix sludge. In present, there are less public reports about one-stage ATAD reactor applying in sludge treatment.

The ATAD digestion system depends on a thermophilic environment created by microbial aerobic metabolism. Some reports indicated that temperature was the most significant factor controlling pathogen inactivation during aerobic digestion of wastewater sludge (*Carrington et al., 1991; Scheuerman et al., 1991; Surampalli et al., 1993*). Therefore, the digestion temperature of sludge probably plays an important role in the stabilization. Using a specially designed, closed, and insulated facility, this paper investigated that operating parameters influenced ATAD temperature variations and VS destruction, and reactor temperature influenced on degradation of sludge solids and elimination of pathogens in return.

I. METHODS AND MATERIALS

A. *Experimental facilities and operation*

As shown in Fig. 1, the ATAD is a compact rectangle reactor, fully covered with 120mm thick insulation material. The tank net size is 1.5×1.0×2.0m with 2.25 m^3 active volume. Installed aeration devices consist of a diffuser tube at the bottom of the reactor and electric air compressors to disperse atmospheric air into the biomass. Sludge inside the reactor is sufficiently mixed in a continuous circulating manner in submerged pipe powered by 1.5KW power recirculation pumps, which create a slow, upward stream in large parts of the tank contents. Sludge circulation pipes are installed with a whirlpool flow sensor to monitor circulation flowrates (6.5~14.0m^3/h) controlled by valves on the pipe. Because monitoring of dissolved oxygen in ATAD system has proven to be quite difficult, the ATAD system used an on-line oxidation-reduction potential (ORP) as an indicator of the aerobic character. Normally foam is produced during aerobic

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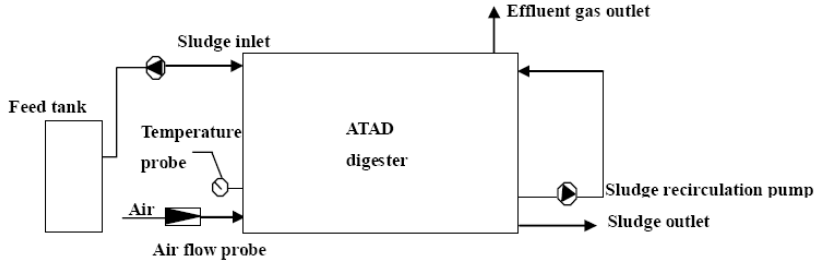

Figure 1. Schematic diagram of ATAD process system

treatment. A foam cutter positioned on a liquid surface can convert some of the foam bubbles to liquid. On top of the reactor, an odor controller was used to collect effluent gas. Other equipments include temperature probes in the middle of the reactor and level monitors. Sludge was first pre-thickened in a 2m deep feed tank, and then fed in batches until completely filled and drained over each period of hydraulic retention time (HRT) of 15~20d.

B. *Experimental sludge*

Feed sludge from Shanghai Municipal Sewage Plant, consisting of a mixture of primary sludge and waste biological sludge at a ratio of 4:6~3:7, pre-thickened to 5~8% total solid in a feed tank was used. It is necessary that concentrated VS inputs provide the required energy-rich substrate to support the autothermal mode of an ATAD operation. The municipal sludge contained 4.2~8.2% (42~82g/L) total solids (TS), 25~49g/L VS, 55~72% VS/TS and a pH of 6.5~7.0.

C. *Experimental design*

In a batch operation, the effects of sludge VS levels on reactor temperature were studied. The operating parameters of different VS concentration of feed sludge are shown in Table Ⅰ.

To investigate effect of HRT on reactor temperature, the HRT of feed sludge at 34.3gVS/L was extended to 20 days. Operational parameters of aeration rate and circulation mixing flowrates were set at 1.16$m^3/m^3 \cdot h$ and 10~12 m^3/hr respectively.

To determine the effect of aeration on temperature, a feed sludge with a TS concentration of 4.9~5.2% (VS/TS at 60~65%) was evaluated in a batch operation for 15d under a mixing flowrates of 10~14m^3/h and three different aeration rates ($m^3/m^3 \cdot hr$) of 0.8~0.9, 1.16~1.18 and 1.7~2.0. The ORP of feed sludge VS at 24.7g/L and 49.2g/L were measured in condition of airflow at 0.6~0.8 $m^3/m^3 \cdot h$ and 0.9~1.1 $m^3/m^3 \cdot h$ respectively.

To research the effect of circulation flowrates on reactor auto-rise temperature and VS removal rate, a feed sludge with TS levels of 6~7% and VS concentration(g/L) of 30~40 g/L was operated for 15d under aeration rates($m^3/m^3 \cdot h$) of 1.16~1.56. The effect of temperature on pathogen reduction was studied using sludge with VS concentrations of 34.3g/L and 24.7g/L reduction during digestion. *Fecal coliforms*, *Fecal streptococci* and *Salmonella* were selected as pathogen indicators to be determined at different digestion time.

D. *Analytical methods*

Total solids (TS), volatile solids (VS) and pH were analyzed daily from duplicate samples according to Standard Methods (APHA 1976). *Fecal coliforms*, *Fecal streptococci* and *Salmonella* were determined on dry mass basis by methods described in Standard Methods for Examination of Water and Wastewater (1995).

Operational data such as power level, temperature and aeration rates were recorded daily. The Oxidation-reduction potential (ORP) was measured by an on-line instrument twice every day.

II. RESULTS AND DISCUSSION

A. *Effect of reactor temperature*

1) *Feed sludge concentrations*

Heat released by the destruction of volatile solids (VS) of influent sludge during the digestion process resulted in thermophilic temperatures (45~60°C) in the ATAD. Apparently, influent sludge concentrations have significant effects on temperature auto-rise in the ATAD reactor. The changes in reactor temperature with detention time are shown in Fig.2. The corresponding changes in VS are also shown (Fig.3).

As shown in Fig.2 and Fig.3, under fixed circulation flowrates (10~12m^3/h), heat produced by biodegradable volatile solids decomposition was related to volatile solids (VS) concentrations of feed sludge. Different VS levels of feed sludge caused different reactor temperatures in the ATAD. Within a digestion detention time of 15 days, the temperature was up for the first few days, and then leveled off. The reactor temperature of VS concentration at 34.3g/L (54°C) was the highest among the three VS concentrations of feed sludge, closely followed by VS concentration at 49.2g/L (53°C) and VS concentration at 24.7 g/L (46°C). Based on the working principle of ATAD, thermophilic temperatures are mainly obtained through heat released by the destruction of VS. The higher the VS reduction rate, the more heat released and the higher the temperature.

Among the three VS concentrations of feed sludge, the VS reduction rate for the VS value of 34.3g/L was highest (48.7%) on the 11th day in the 15d retention time. The reduction rate for VS of 49.2g/L was 43.1% while that for VS of 24.7g/L was the lowest (30.8%) since it had the lowest reactor temperature.

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TABLE I. PROCESS PARAMETERS OF THE BATCH PROCESS IN THE ATAD REACTOR

Index Name	Range		
Feed VS /(g/L)	49.2	34.3	24.7
Feed TS /(g/L)	81.6	62.2	45.3
Feed VS/TSS	60.2	55.0	61.2
Detention time /d	15	15	15
Organic loading /(KgVS/m^3• d)	3.51	2.46	1.76
Circulation flowrates /(m^3/ h)		10-12	
Airflow /(m^3/m^3• h)	0.9-1.1	0.8-0.9	0.6-0.8
Feed sludge temperature /°C	28	29	25

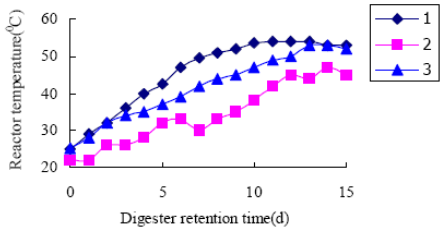

Figure 2. Change of temperature with retention time

1. Concentration of VS of feed sludge at 34.3g/L
2. Concentration of VS of feed sludge at 24.7g/L
3. Concentration of VS of feed sludge at 49.2g/L

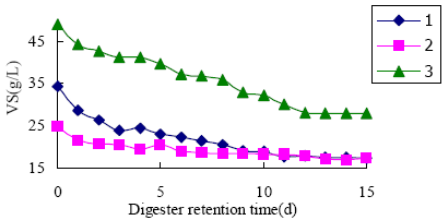

Figure 3. Change of Volatile solids with retention time

1. Concentration of VS of feed sludge at 34.3g/L
2. Concentration of VS of feed sludge at 24.7g/L
3. Concentration of VS of feed sludge at 49.2g/L

Compared with VS concentration at 24.7g/L, sludge of VS at 34.3g/L has more organics that release greater energy to maintain the thermophilic temperature, which increase biodegradation rates of organics and removal rate of VS. However, the stabilization effect of sludge VS at 49.2g/L did not exceed that of VS at 34.3g/L. The reason was that sludge at 49.2g/L, which has the high (8.2%) total solids level, caused a high viscosity which tended to make mixing difficult and inhibit oxygen transfer. Table Ⅱ presents the viscosities of high total solids contents under different temperature condition. Compared with low concentrations, it was possible that the high sludge solids concentrations affected oxygen transfer, the oxygen gradient and degradation rate.

The ORP of feed sludge VS at 49.2g/L measured tended to rise progressively from the lowest value at -425~-356 mv the first day to -330~-303 mv on the 15th day. Different from VS at 24.7g/L (the results were shown in "Aeration rate and

TABLE II. VISCOSITIES OF DIFFERENT SOLIDS CONTENT UNDER DIFFERENT TEMPERATURE CONDITION

Sludge solids content /(%)	Air temperatures /(°C)	Sludge viscosities /(Pa·s)
7.27	12	0.821
7.26	15	0.717
7.37	33	0.263
7.0	32	0.157
6.86	32	0.151
4.38	32	0.077

Oxidation Reduction Potential (ORP)" section), sludge VS at 49.2g/L is at anaerobic situation.

Insufficient levels of oxygen would directly result in poor aerobic digestion and incomplete oxidation of organic matter thereby decreasing the rate of reduction in VS. Furthermore, reactor temperature would be low and unable to achieve a thermophilic range. A minimum VS of 3% to 4% and a TS of less than 7% were typical requirements for ATAD (*Kelly et al, 2003*). In this study concentration of feed sludge below 30g/L was unable to achieve stabilization within 15d detention time. Consequently, VS concentrations of initial sludge at 30~40g/L with total sludge solids at 6~7% are suitable.

2) Hydraulic retention time (HRT)

The hydraulic retention time of ATAD reactor was shorter than that of conventional biodegrading digesters (Roš and Zupančič, 2002). Therefore, the operating temperature is the main parameter in the ATAD process.

As shown in Fig. 4, the reactor temperature leveled off at 54~53°C following an initial rise to the highest value of 54°C on the 11th day, then gradually declined from the 16th day. The curve demonstrated the difficulty of maintaining the thermophilic temperature during 20 days HRT. In the same way, the rate of removal of VS decreased progressively with the extension of HRT, then almost leveled off between the 16th and 20th day. The rate of reduction in VS was highest (48.7% on the 12th day, 40.2% on the 9th day) and therefore met the 38% VS reduction specified by U.S. EPA (1990). Based on the removal rate of VS, the results were apparently satisfactory if HRT is in 10d~15d. However, the continuous removal of VS with a little extent from the 15th to the 20th day meant some biodegradable volatile sludge remained in the reactor. It was unnecessary to extend HRT beyond 15days due to increasing running costs.

The heat produced by the destruction of organics gradually decreased with the decrease in VS according to Fig.4, and simultaneously, the reactor wall heat loss due to climate. A balance between heat input and output created the desired thermophilic temperature, while a decrease in heat produced resulted in a continuous decline in reactor temperature.

3) Aeration rate and Oxidation Reduction Potential (ORP)

The study observed that aeration rates had direct and apparent correlation with temperature. It was considered that an increase in aeration rate should elevate dissolved oxygen levels in the biomass and enhance the VS reduction rate. However, increasing aeration rate resulted in a drop in reactor temperature (Table Ⅲ).

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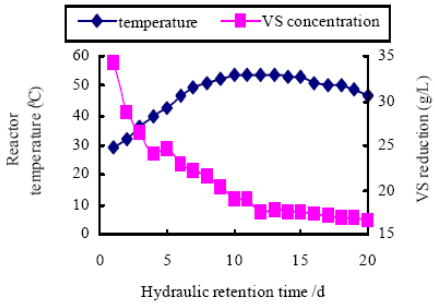

Fig.4 Chang of hydraulic retention time with temperature

An appropriate adjustment of aeration rate is advantageous to temperature rise and VS reduction. At the aeration rate of 0.8~0.9 $m^3/m^3 \cdot h$, the highest temperature and VS reduction rate were 54 ^{0}C and 44.8% respectively during the 15d HRT. The corresponding temperature of sludge at the aeration rate of 1.7~2.0 $m^3/m^3 \cdot h$ was only 47.2 ^{0}C, with the lowest daily increase of 1.8 ^{0}C /d (Table Ⅲ). It was noted that the effluent gas contained vapour heat, and when off gas quality increased with increasing aeration rate, a large amount of heat within the reactor was lost.

Compared with CAD process, ATAD process is characteristic of high sludge concentration and thermophilic temperature, therefore, dissolved oxygen should not normally be measurable by DO sensor above 40 ^{0}C temperature. The oxidation reduction potential (ORP) was used to monitor oxidative conditions and biological activity. The ORP of feed sludge VS at 24.7g/L measured during 18d HRT tended to rise progressively from the lowest value of -357~-300mv in the previous 3 days to -122~-57mv on 8th~14th d, and the highest values ranging between –87mv and +32mv was observed from the 14.8th to 18th day. When raw sludge was fed to the reactor, the large amount of oxygen needed for biodegradation of organics caused a fall in oxygen levels and ORP. Oxygen demand for aerobic bacteria declined when sludge tended to be stable and, as a result, the ORP gradually increased.

ORP values of feed sludge VS at 24.7g/L showed that the ATAD process was not a completely aerobic digestion system. The low-level ORP suggested a considerable portion of anaerobic and fermentative bacteria in ATAD, as contained in previous reports (*Staton et al.,2001*) since sludge concentration in ATAD digester was 30~40 times higher than in conventional aerobic digestion (CAD) tank. In fact, it is difficult that CAD process presents a completely aerobic digestion. Most ATAD operators believe that ambient oxygen levels of approximately 0.5mg/L, which is often classified as microaerobic, are acceptable (*Staton et al., 2001; Mavinic et al., 2001*).

From Table Ⅲ, an increase in aeration rate couldnot lead to an increase in VS reduction and temperature. Excessively high airflow could cause cooling due to evaporative latent heat loss, and fell in VS reduction rate. Nevertheless, a too low aeration rate would lead to anaerobic conditions and was a potential for odors due to inadequate oxygen supply.

Proper aeration rate was a significant operational variable for determining the performance of the ATAD. It was observed that the exact amount of aerobic bacteria needed for optimum performance was not constant; hence, a single oxygen supply rate during the entire digestion process was not feasible. Oxygen supply must vary according to the stage of digestion to avoid insufficiency in the early digestion stage and excessive supply in latter stages.

4) Circulation flowrates and Mechanical energy

The ATAD reactor was equipped with a recirculation sludge pump to facilitate mixing of the reactor contents instead of using a mechanical or aerator-mixer. Sludge driven by the recirculation sludge pump was continuously circulated in sludge pipes and reactor.

In condition of sufficient circulation flowrate, the sludge particle size was reduced and sludge can maintain in suspension. Its surface area was increased to promote contact between air and bacteria. As shown in Table Ⅳ, circulation flowrates increased with energy density raising and had benefit to reactor auto-rise temperature and VS removal. The results of removal rate of VS above 40% and reactor temperature above 50 ^{0}C were satisfied while circulation flowrates at 12~14 m^3/h and energy density at 280~330W/m^3. Instead, the low removal rate of VS indicated insufficient sludge mixing and unequal oxygen diffuse while circulation flowrates less than 8.4m^3/h. However, reactor temperature arrived at above 50 ^{0}C, it showed that mixing energy of circular pump could keep system thermophilic temperature, because high sludge viscosities leads to a part of mechanical energy transforming heat. Therefore, the mixing energy provides supplemental heat through energy dissipation, and could offset the heat loss due to ambient temperature particularly in winter and was capable of sustaining thermophilic temperature when a balance was achieved between heat generation and heat loss. Kelly et al (*2003*) reported that the mixing energy supplied less than 30 percent of the heat. It is demonstrated that the process is not completely autothermal and requires a heat component from mechanical mixing energy. Apparently, heat is produced in two ways: through energy release due to biodegradation volatile solids (*BVS*) and through heat input from mechanical mixing energy.

In conclusion，System can keep normal in condition of circular flowrates at 12~14 m^3/h and mixing power density at 280~330W/m^3.

B. Effect of Reactor temperature on inactivation of pathogens

Sludge generated from sewage plants contains a large number of pathogenic organisms including bacteria, virus, and parasites separated in the water treatment processes. For safe land application of sludge, it is necessary to reduce the levels of pathogens below detectable limits. The ATAD process is capable of producing a pasteurized biosolid that meets the EPA 503 regulations for Processes to Further Reduce Pathogens (PRFP). *Fecal coliforms*, *Fecal streptococci* and *Salmonella* were determined for sludge with a VS concentration of 34.3g/L , and *Fecal coliforms* for VS of 24.7g/L (Table Ⅴ).

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TABLE III. EFFECT OF AERATION RATE ON REACTOR TEMPERATURE AND VS REDUCTION

Test order	Temperature of feed sludge /(^{0}C)	Maximum temperature in 15d HRT /(^{0}C)	Aeration rate /(v/v·hr)	Change in temperature /(^{0}C /d)	Removal rate of VS /%
1	28.0	54	0.81－0.9	2.9	44.8
2	22.0	46	1.16-1.18	1.8	36.2
3	25.4	47.2	1.7-2.0	1.8	21.5

TABLE IV. EFFECTS OF CIRCULATION FLOWRATES ON REMOVAL RATE OF VS AND REACTOR TEMPERATURE

Test order	Average air temperature /(^{0}C)	Reactor highest temperature /(^{0}C)	Circulation flowrates /(m^3/h)	Power density /(W/m^3)	Rise temperature rate /(^{0}C /d)	Removal rate of VS /%
1	25	54	12-14	280-330	2.6	48.7
2	25	54.6	12-14	280-330	2.8	43.1
3	32	54	8.4-10	200-238	2.9	37.5
4	24	50	6.5-7.2	155-172	2.7	31.4
5	10.2	40.5	12-14	280-330	2.2	26.8
6	22.0	46	8-9	200-215	1.8	36.2

TABLE V. RESULTS OF PATHOGEN DESTRUCTION

	VS concentration of 34.3g/L					VS concentration of 24.7g/L		
HRT d	Temperature /(^{0}C)	pH	Fecal coliform /(cfu/gTS)	Salmonella /(MPN/4gTS)	Fecal streptococci /(cfu/gTS)	Temperature /(^{0}C)	pH	Fecal coliforms /(cfu/gTS)
0	25	7.00	6.67×10^7	7.33×10^2	9.15×10^5	22	7.12	1.01×10^6
2	31	7.06	-	-	6.10×10^3	-	-	-
8	46	7.68	-	-	3.88×10^2	-	-	-
11	54	7.92	No detection	3.4	No detection	42	7.24	3.04×10^2

At HRT of 11 days, the number of pathogens in sludge with VS concentration of 34.3g/L gradually decreased below detection level for *Fecal streptococci*, *Fecal coliforms* and *Salmonella*, which met Class A requirements. For feed sludge of VS at 34.3g/L, thermophilic digestion successfully inactivated enteric pathogenic bacteria at a temperature of 54 ℃ on the 11th day. However, sludge with VS of 24.7g/L still contained *Fecal coliforms* up to 3.04×10^2 cfu/gTS due to the relatively low reactor temperature of 42°C even on the 11th day. Temperature was likely the most important factor influencing the rate of pathogen inactivation. Long exposure to temperatures above 45°C was lethal for pathogens in sludge. A previous study also showed that at a temperature of 45°C *Salmonella* sp. were destroyed below detectable limits within 24 hours, while at 60°C, within a few hours (*Randolph and William, 1982*). It was possible cell lysis could have occurred. As sludge flowed into the reactor, visible constituent organism could not adapt to the thermophilic environment, and must have been killed and replaced with thermophilic bacteria. The cell walls and membranes of death bacteria were decomposed by exoenzymes of theromphiles, resulting in the release of intracellular components from the ruptured cells, which could be converted by thermophiles into biomass and metabolic energy, and ultimately resulting in destruction of VS. The processes of cell lysis simultaneously led to inactivation of pathogenic organisms. It was evident that the achievement of thermophilic temperatures can be used to effectively control most, if not all, pathogens in sludge.

Nevertheless, other factors in the reactor such as pH, HRT, dissolved oxygen, concentration of volatile solid, etc, also contributed to pathogen inactivation (*Ugwuanyi et al., 1999*). High pH or high concentration of ammonium nitrogen could promote pathogen reduction (*Zábranská et al., 2003*). Other studies have shown the concept of heat and pH synergism for virus inactivation (*Randolph and William, 1982; Ward et al., 1977*). In some ATAD tanks, pH levels between 8.0~9.0 have been measured with prolonged retention time due to the transformation of organic nitrogen to ammonium nitrogen, and inhibition of nitrification and denitrification at thermophilic temperatures (*U. S. EPA, 1990*). As shown in Table Ⅴ, the pH of sludge with a VS concentration of 34.3g/L and 24.7g/L increased with HRT.

III. CONCLUSIONS

In the ATAD digestion system, temperature rise in the reactor was a factor of VS concentration. High initial sludge solids concentration provided enough organic substrate for adequate heat production. On the other hand, the solids must also be at an optimum concentration (VS of 30~40g/L with TS of 6~7%) to allow adequate mixing and oxygen movement. Optimal results were obtained with an initial

In batch operation experiments, the drop in reactor temperature would occur after 15d HRT. A slight increase in removal of VS with prolongation of HRT from 15 to 20 days suggested it was unnecessary to prolong the HRT beyond 15days due to increasing running costs.

There was a correlation between aeration rate and temperature. High aeration rate was necessary to maintain the desired aerobic environment within an ATAD reactor; however,

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excessively high aeration rates could lead to exhaustion of gaseous heat release and drop in temperature. The ORP could describe the oxidizing state in an ATAD digester, but there seemed to be a combination of anaerobic, fermentative and aerobic biological activities in the ATAD as well.

Recirculation sludge pumps performed not only mixing of sludge particles with air, but also provided supplemental heat through mixing energy dissipation. It was shown that the ATAD process was not completely autothermal, but simultaneously requires a heat component from mechanical mixing energy to maintain reactor temperature.

Temperature in ATAD tank was likely the most important factor influencing the rate of pathogen inactivation. However, several other factors (e.g. pH, HRT or SRT, dissolved oxygen, and concentration of volatile solid) also contributed.

ACKNOWLEDGMENT

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