Modelling and Dynamic Compensator Control of the Anaerobic Digestion of Organic Wastes

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The paper deals with the modelling and control of anaerobic fermentation processes (anaerobic digestion). Laboratory experiments have been carried out on an automated laboratory-scale biogas unit. For this process 2nd and 5th order non-linear models have been considered. A simple methodology for parameters estimation, based on non-linear optimisation method, has been developed. The control is reduced to the regulation of biogas production rate or the concentration of the outlet polluted organic matter. For design purposes the non-linear model has been transformed into a linear one with interval parameters. In both cases (the regulation of biogas production rate or the concentration of the polluting organics) compensators have been designed according to the internal model principle. The effectiveness of the algorithms designed has been illustrated by simulation experiments. An important feature of the proposed algorithms is their robustness over the parameter uncertainties in the process models.

Keywords:
Anaerobic digestion; organic wastes; laboratory experiments, modelling; identification; compensator control; simulation.

Introduction

Anaerobic digestion (methane fermentation) is a biological process in which organic matter is transformed by microorganisms into methane and carbon dioxide (biogas) in the absence of oxygen (*Price* and *Cheremisinoff*,[7] 1981; *Dubrovskis* and *Viesturs*,[5] 1988). Recently, methane fermentation has been used widely in life processes and confirmed as a promising method of solving some energy and ecological problems in agriculture and industry. Generally, this process is carried out as a continuous biotechnological process (BTP).

Many mathematical models of the process in stirred tank bioreactors (STBR) are known (*Angelidaki*[1] et al., 1999; *Batstone*[2] et al., 2001; *Hill* and *Barth*,[6] 1977; *Bastin* and *Dochaian*,[3] 1991; *Simeonov*,[9] 1994; *Simeonov*[10] et al., 1996). However, most of them are very complicated and not appropriate for control design purposes.

Due to the very restrictive on-line information, the control of the process is often reduced to the regulation of biogas production rate or the concentration of the outlet organic matter at a desired value in the presence of some perturbations (*Bastin* and *Dochain*,[3] 1991; *Simeonov*,[9] 1994). Investigations show that classical linear controllers have no good performances in this case (*Simeonov*,[9] 1994). More sophisticated adaptive linearizing controllers heve been studied (*Bastin* and *Dochain*,[3] 1991, *Van Impe*[15] et al., 1998; *Dochain*,[4] 2001), but due to some implementation difficulties they are not so popular in practice.

A theoretical idea for robust compensator control of a continuous BTP is presented in (*Stoyanov* and *Simeonov*,[13] 1996). The proposed controller has a good robustness and simple realisation in the case of availability of reliable sensors for on-line measurement.

The aim of the paper is to choose an appropriate model of anaerobic digestion on the basis of laboratory-scale experiments and to design and investigate a linear controller with output dynamic compensator for this process.

Mathematical modelling of the process

Process description

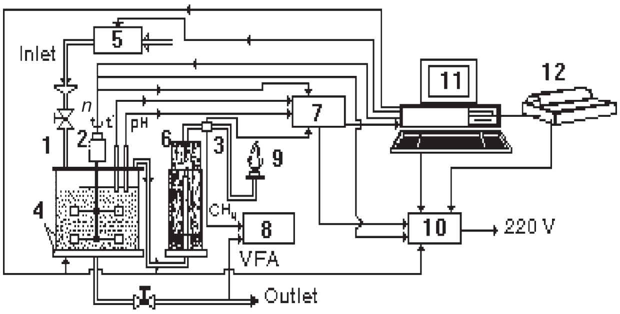
Anaerobic digestion of highly concentrated organic pollutants from animal farms is used. This process is a very complicated one, involving hundreds of possible intermediate compounds and reactions, each of which catalysed by specific enzymes or catalysts. Many of the transformations can be accomplished by one of several alternative metabolic pathways, and biochemists and microbiologists continue with their attempts to define and describe more precisely the various mechanisms. The overall biochemical reaction can be illustrated by the following scheme (*Price* and *Cheremisinoff*,[7] 1981):

$$\text{Organic matter} \xrightarrow[\text{microorganisms}]{\text{Anaerobic}} CH_4+CO_2+H_2+N_2+H_2S \quad (1)$$

For the studies of microbiology, biochemistry and technology, the anaerobic digestion is generally considered to progress in three stages (*Hill* and *Barth*,[6] 1977; *Price* and *Cheremisinoff*,[7] 1981): hydrolysis and liquefaction of the large, insoluble organic molecules by extracellular enzymes; acid production stage; a methane production stage. The low-molecular-mass acids produced in the acid production stage are further degraded to methane and carbon dioxide (biogas) by a highly specialised group of bacteria, commonly referred to as methane producing bacteria.

Experimental studies

The experiments presented in the paper have been carried out on an automated laboratory -scale biogas unit, developed and adapted to fulfil the requirements for anaerobic digestion. The experimental set-up is shown on Fig.1. It consists of a biotechnical part and a control unit (*Tzonkov S.*, *I. Simeonov*[14] et al.).



F i g . 1 – *Experimental set-up*

The biotechnical device includes an anaerobic fermentor with a maximum working volume of 2.5 dm^3 (1), a DC motor with an agitation system (2), a heating system (4), a system for inlet of the substrate with a peristaltic pump (5), and a biogas holder (6). The fermentor is mechanically stirred and maintained at a constant temperature.

The control unit has been designed on the basis of a conventional personal computer and consists of sensors for: temperature (*T*), agitation speed (*n*), pH and biogas flow rate (3); converters unit (7); a PC (11) and a printer (12). The experimental set-up includes also a watt-hour meter for its electrical energy consumption (10), a gas flame (9) and a gas chromatograph (8) for biochemical analysis. Experimental studies has been carried out in the research group "Mathematical Modelling and Computer Sciences" of the Institute of Microbiology of the Bulgarian Academy of Sciences with the aim of modelling and designing different algorithms for the control of methane fermentation. The continuous anaerobic digestion has been carried out under the following conditions: temperature has been regulated at a constant value (34 ± 0.5) °C with a PID controller; pH has been corrected in some cases of deviations from the allowed values (6,3 – 7,8); a given volume of the fermented biomass has been taken out of the fermentor once a day (F) and the same quantity of organic waste added into it. After starting the process, responses of Q have been taken down under step changes of dilution rate D ($D = F/V$, where V is the volume of the fermented medium) and of influent organic concentration γ_{S0i}. During the process, quantitative and qualitative analyses of the content of fatty acids (acetate and propionate) in the substrate have been performed on the gas chromatograph.

Process models

Many mathematical models of the process in STBR are known. They are usually presented as sets of ordinary non-linear differential equations. In this case, the following two models are considered:

– a model based on a one -stage reaction scheme (*Bastin* and *Dochain*, 1990):

$$\frac{d\gamma_X}{dt} = \mu\gamma_X - D\gamma_X \quad (2)$$

$$\frac{d\gamma_S}{dt} = -\mu\gamma_X + D(\gamma_{S0i} - \gamma_S) \quad (3)$$

$$Q = k_2\mu\gamma_X \quad (4)$$

with:
$$\mu = \frac{\mu_m\gamma_S}{K_S + \gamma_S} \quad (5)$$

– a model based on a three-stage reaction scheme (*Hill* and *Barth*, 1977; *Simeonov* et al., 1996):

$$\frac{d\gamma_{S0}}{dt} = -D\gamma_{S0} - \beta\gamma_X\gamma_{S0} + DY_p\gamma_{S0i} \quad (6)$$

$$\frac{d\gamma_{X1}}{dt} = (\mu_1 - k_1 - D)\gamma_{X1} \quad (7)$$

$$\frac{d\gamma_{S1}}{dt} = -D\gamma_{S1} + \beta\gamma_{X1}\gamma_{S0} - \frac{(\mu_1\gamma_{X1})}{Y_1} \quad (8)$$

$$\frac{d\gamma_{X2}}{dt} = (\mu_2 - k_2 - D)\gamma_{X2} \tag{9}$$

$$\frac{d\gamma_{S2}}{dt} = -D\gamma_{S2} + Y_b\mu_1\gamma_{X1} - \frac{(\mu_1\gamma_{X2})}{Y_2} \tag{10}$$

$$Q = Y_g\mu_2\gamma_{X2} \tag{11}$$

with: $$\mu_1 = \frac{\mu_1^{\max}\gamma_{S1}}{\gamma_{S1}+K_{S1}}, \quad \mu_2 = \frac{\mu_2^{\max}\gamma_{S2}}{\gamma_{S2}+K_{S2}} \tag{12}$$

$$\gamma_{S3} = c_0\gamma_{S0} + c_1\gamma_{S1} + c_2\gamma_{S2} \tag{13}$$

The mass concentration of the outlet polluted organic matter γ_{S3}, defined by equation (13) (c_0, c_1, c_2 are constants), may be associated with the parameter COD (Chemical Oxygen Demand), characterising the depollution effect.

Parameters identification

Parameters identification is a very important step in building a dynamical model which fairly represents the biological reactions in an anaerobic fermentor and designing the control system. Experimental data needed for the identification of the unknown parameters have been obtained by means of the above described lab-scale unit located at the Institute of Microbiology of the Bulgarian Academy of Sciences. Sets of experimental data for the input (dilution rate D) and the output (biogas production rate Q) for constant values of γ_{S0i}, are available. Experimental data for Q have been registrated for step changes of D in the range from 0,025 to 0,1.

The estimation of the coefficients in the non-linear models (2), (3), (4) and (5) or (6), . . ., (12) is a very hard problem, due to the rather restricted information – only concentration γ_{S0i} and biogas production rate Q can be measured. In our case, an optimisation software with the Hook and Jeevs method and a quadratic performance index (criterion J) depicted on Fig. 2 has been used to perform the identification. This method is robust when varying the initial conditions and manages to find the solution even in the case of a complicated non-linear function. Different initial conditions results in different estimates of the model parameters. The performance index has a relatively constant value. Since each parameter has its own physical, biological or chemical meaning, the appropriate parameters set must be chosen so as to satisfy some requirements determined by the process.

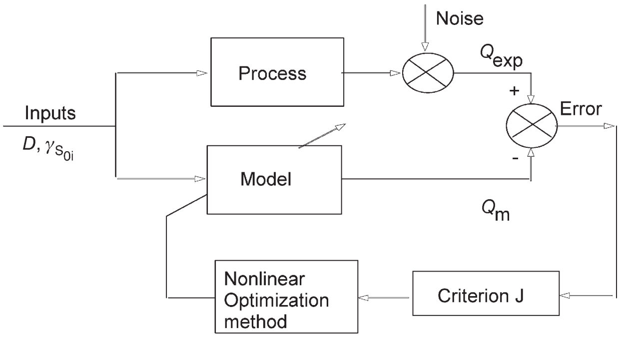


Fig. 2 – *Non-linear identification method*

On this basis a simple methodology was developed (Simeonov, 2000). It consists of the following step-by-step procedure:

1. Determination of the initial values of the unmeasurable state variables using an optimisation method.

2. Separation of the model coefficients into two (several) groups using sensitivity analysis and expert knowledge.

3. Estimation of the first group of coefficients (the most sensitive ones) with a known second part (by expert knowledge or from literature), using an optimisation method.

4. Estimation of the second part of coefficients with the above-determined values of the first group of coefficients.

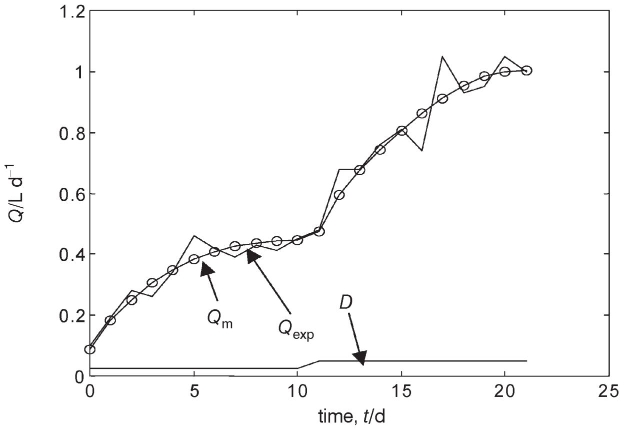
5. Validation of the obtained results with other sets of data.

This methodology has been applied for the model described by equations (2), . . . ,(5). μ_m and k_2 have been in the first group of coefficients, and in the second – k_s and k_1. The quantities obtained are: μ_m = 0.2; k_2 = 31; k_s = 0.8 and k_1 = 27. On Fig. 3a and 3b (for step-changes of D = 0.05 and 0.075 d^{-1}) and on Fig. 3c (for step-changes of γ_{S0i}) the good fitting between the experimental and simulated curves is evident.

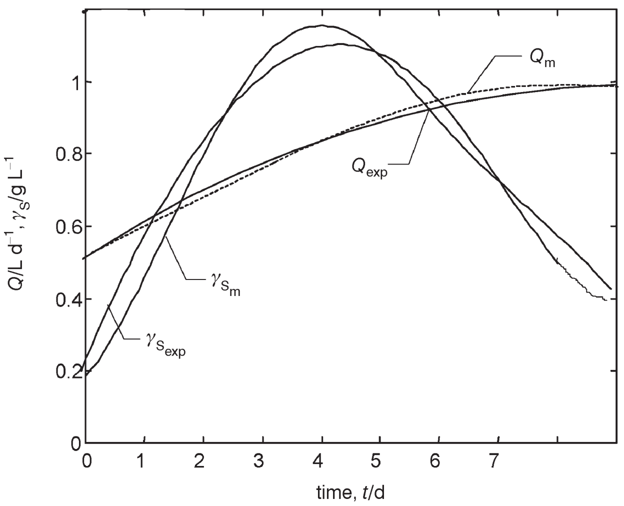
For the 5^{th}-order model described by equations (6), . . . , (12), the following coefficients have been obtained (*Simeonov* et al.,[10] 1996): μ_{1max} = 0.202; μ_{2max} = 0.5976; k_1 = 0.000667; k_2 = 0.00389; K_{s1} = 0.0089; K_{s2} = 0.594; Y_1 = 0.000575; Y_2 = 0.71; β = 2.465; Y_b = 49.74; Y_g = 109.5 and Y_p = 1.678.

Input-output static characteristics

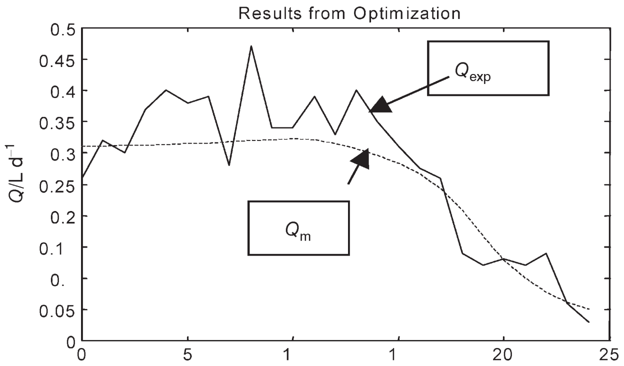
For the 2^{nd}-order model (2), (3), (4) and (5) a set of input-output static characteristics $Q = Q(D)$ and $\gamma_S = \gamma_S(D)$ have been obtained (γ_{S0i} = 2,. . . ,5) (Fig. 4a). For the 5^{th}-order model (8), …, (14), the similar input-output characteristics $Q = Q(D)$ and $\gamma_{S3} = \gamma_{S3}(D)$ are shown on Fig.4b (γ_{S0i} = 30,..,70). It is obvious that the characteristics $Q = Q(D)$ are strongly non-linear with an explicit maximum for the two models.



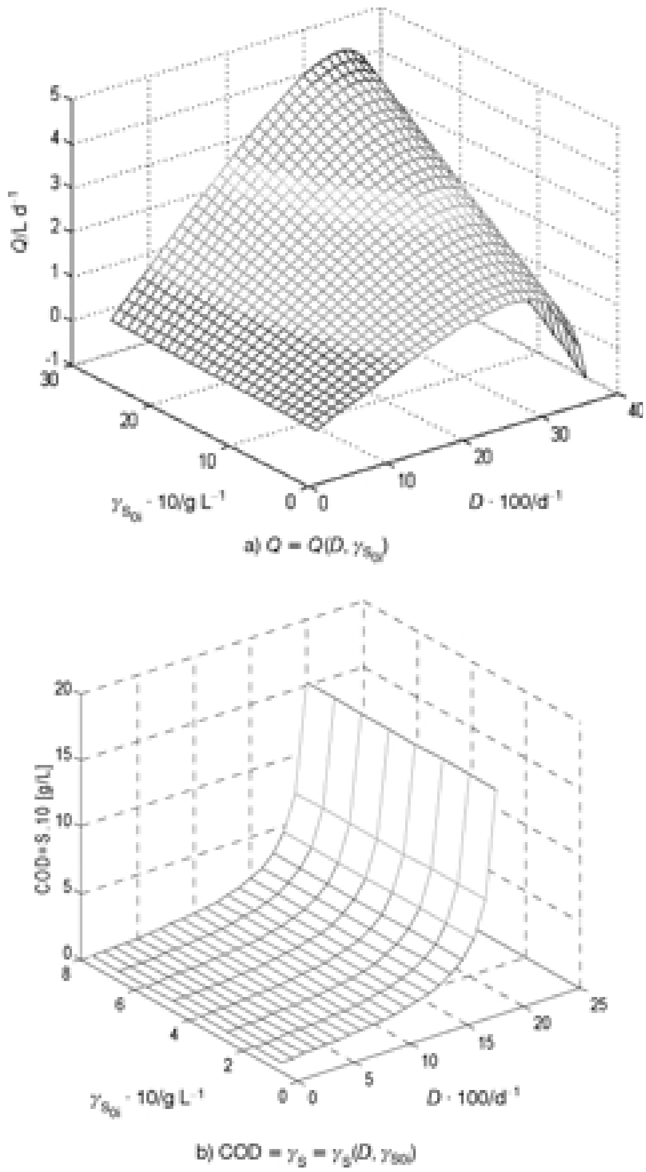
F i g . 3 a – *Transient response of biogas flow rate (Q) for step changes of the dilution rate (D) (Q_{exp} – experimental data, Q_m – simulation results for the 2^{nd} order model)*



F i g . 3 b – *Transient response of biogas flow rate (Q) and effluent organics concentration (S) for step changes of the dilution rate (D) (Q_{exp} and S_{exp} – experimental data, Q_m and S_m– simulation results for the 2^{nd} order model)*



F i g . 3 c – *Transient response of biogas flow rate (Q) for step change of influent organics concentration (S_{0i}) (Q_{exp} – experimental data, Q_m – simulation results for the 2^{nd} order model)*



F i g . 4 – *Input-output static characteristics*

Control of the process with dynamic compensator

The aim of the control is the regulation of biogas production rate Q (energy effect) or effluent organics concentration γ_S (depolution effect) at a desired value (set point) Q^* or $\gamma_S{}^*$, respectively, acting upon dilution rate D. Usually γ_{S0i} is an unmeasurable disturbance. One simple approach for the control of the methane fermentation process is the linearization of the non-linear model in an admissible range and linear control design (*Tzonkov* et al.,[14] 1992). Usually there is not only one working point but a whole interval of normal work. In this paper, a transformation of the non-linear model into a linear one with interval coefficients, is suggested. It is obtained as follows:

$$\frac{dx(t)}{dt} = [A_0 \pm \Delta A]\, x(t) + [b_0 \pm \Delta b]\, D(t),$$

$$y(t) = [c_0 \pm \Delta c]\, x(t) \tag{14}$$

$$A_0 = \begin{bmatrix} 0 & a_{12}^0 \\ a_{21}^0 & a_{22}^0 \end{bmatrix}, \quad \Delta A = \begin{bmatrix} 0 & \Delta a_{12} \\ \Delta a_{21} & \Delta a_{22} \end{bmatrix}$$

$$b_0 = \begin{bmatrix} b_1^0 \\ b_2^0 \end{bmatrix}, \ \Delta b = \begin{bmatrix} \Delta b_1 \\ \Delta b_2 \end{bmatrix}, \ c = [c_1^0 \ c_2^0], \ \Delta c = [\Delta c_1 \ \Delta c_2]$$

where $x^T = [\gamma_X \ \gamma_S]$ is state vector, D is control input, y (Q or γ_S) is output, γ_{S0i} is unmeasurable disturbance. The elements $a_{12}{}^0 = 0.0530$, $a_{21}{}^0 = -1.89$, $a_{22}{}^0 = -1.5$, $b_1{}^0 = -0.5$, $b_2{}^0 = 13.6$, $c_1{}^0 = 2.17$ and $c_2{}^0 = 1.64$ are the nominal coefficients of the model (15). $\Delta a_{12} = 0.43$, $\Delta a_{21} = 1.8$, $\Delta a_{22} = 1.1$, $\Delta b_1 = 0.05$, $\Delta b_2 = 1.5$, $\Delta c_1 = 2.1$ and $\Delta c_2 = 1.3$ are their uncertain parts. If $\Delta a_1 = \Delta a_2 = \Delta b_1 = \Delta b_2 = \Delta c_1 = \Delta c_2 = 0$, a nominal description is obtained.

A two-part dynamic output compensator is suggested (*Stoyanov*,[12] 1989; *Stoyanov* and *Simeonov*,[13] 1996). The first part is an internal model of step exogenic signals (the set points and disturbances) which includes an integrator. The internal model ensures the asymptotic tracking of stepwise-changed set-points and the rejection of unmeasurable step disturbances. Its description is as follows:

$$\frac{dq(t)}{dt} = e(t) = r(t) - y(t), \ q(t_0) = q_0, \tag{15}$$

$$u_p(t) = k_i q(t), \tag{16}$$

where q is integrator state, u_p – internal model output, r – set point, e – error and k_i – internal model coefficient.

The second part of the compensator (stabilisation part) is as follows:

$$\frac{d}{dt}\begin{bmatrix} v_1 \\ v_2 \end{bmatrix} = \begin{bmatrix} 1 & -r_1 \\ 0 & -r_2 \end{bmatrix}\begin{bmatrix} v_1 \\ v_2 \end{bmatrix} + \begin{bmatrix} p_1 \\ p_2 \end{bmatrix} v, \tag{17}$$

$$u_c = [0 \ 1]\begin{bmatrix} v_1 \\ v_2 \end{bmatrix} + wv, \tag{18}$$

where $v = [v_1, v_2]'$ is the state of the stabilising compensator, u_c – its output, r_1, r_2, p_1, p_2 and w are compensator coefficients.

Let the desired characteristic polynomial of the closed loop system for measurement of Q is:

$$H_d = s^5 + 17.55\, s^4 + 122.5\, s^3 + \\ + 428.75\, s^2 + 750.312\, s + 525.219. \tag{19}$$

Let the desired characteristic polynom of the closed loop system for measurement of S is:

$$H_d = s^5 + 5\, s^4 + 10\, s^3 + 10\, s^2 + 5\, s + 0.685. \tag{20}$$

The following linear matrix equation including the unknown compensator parameters has been obtained:

$$\Lambda \xi = \alpha - \beta, \tag{21}$$

where Λ and β include process parameters, α – coefficients of the desired characteristic polynom and ξ – dynamic compensator parameters.

When biogas production rate Q is the output for the 2nd order model (2), (3), (4) and (5), the compensator consists of a first-order internal model and a second order stabilising part. The following compensator parameters have been calculated using equation (21):

$r_1 = 63.89$, $r_2 = 15.84$, $p_1 = 175.71$, $p_2 = 24.01$, $w = -3.87$, $k_i = 1$. When the concentration of outlet organics γ_S is the output, the following compensator parameters have been calculated: $r_1 = 1.318$, $r_2 = 1.774$, $p_1 = 0.1624$, $p_2 = 0.0937$, $w = -0.9134$, $k_i = 1$. The compensators designed in the two cases ensure the desired poles of the nominal closed-loop systems being in the left complex half-plane.

According to the algorithm in (*Schmitendorf* and *Barmish*,[8] 1986): the compensator parameters have to be corrected in the second stage, so that the closed-loop system is stable for the whole range of parameters uncertainty. This correction is not needed in this case, because the closed loop system designed in the first stage is stable for the model with interval parameters. The closed-loop system is shown in Fig. 5.

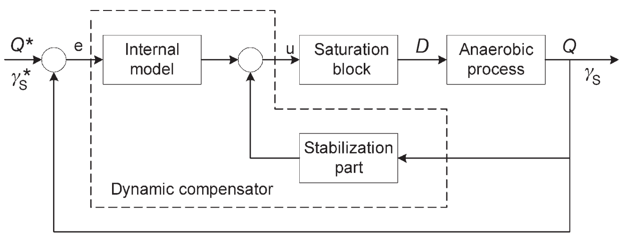


Fig. 5 – *Closed loop system for control of the anaerobic digestion with dynamic compensator*

The theoretical design of controllers with output dynamic compensators on the basis of the 5th-order model of anaerobic digestion is a very hard problem not useful from practical point of view. However, in this case it is possible to apply the obtained result for the 2nd (reduced) order model with some heuristical modifications.

A restriction on the control input is used from a practical point of view (*Bastin* and *Dochain*,[2] 1990):

$$0 \le D(t) \le D_{\max}, \tag{22}$$

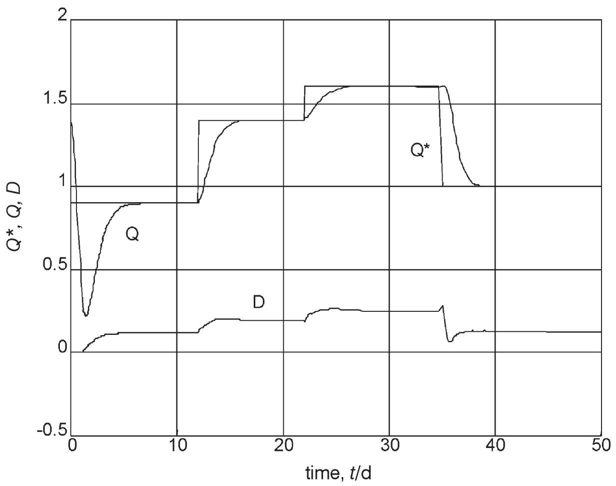
where $D_{\max}$ is a technological bound. On Fig. 5 the condition (22) is shown as a saturation block.

Simulation experiments

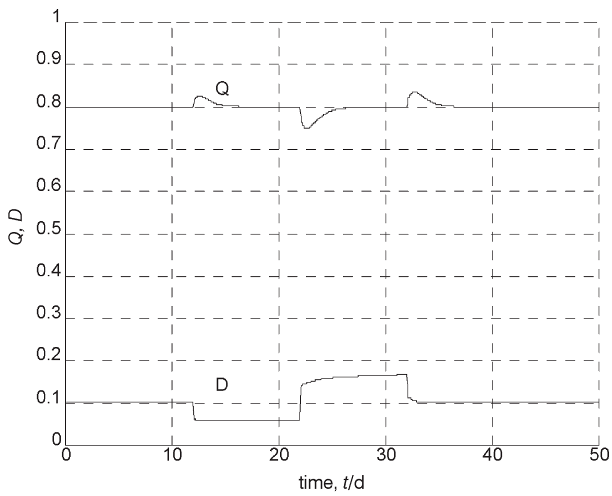
The performances of the designed controllers with a dynamic compensator for the model (2), (3), (4), (5) and output $Q(t)$ have been evaluated by simulation for step changes of:

– set points Q^* in the interval from 0.9 to 1.6 (Fig. 6)

– disturbance γ_{S0i} in the interval from 2 to 5 (Fig. 7).



F i g . 6 – *Simulation results for dynamic compensator control, designed for the 2nd order model, measurement of the biogas flow rate (Q) and step changes of the set point (Q*)*

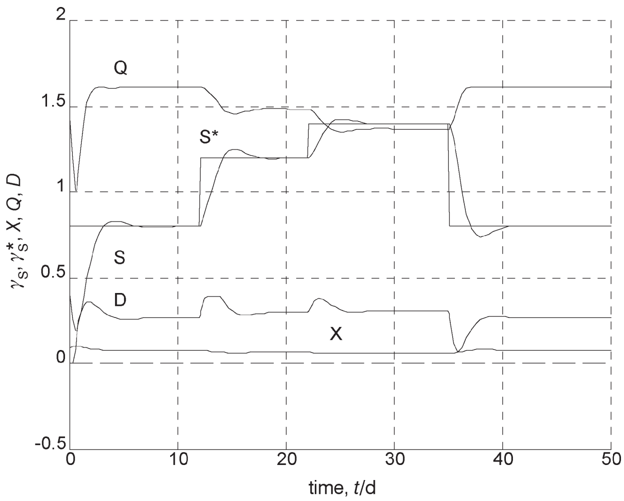


F i g . 7 – *Simulation results for dynamic compensator control, designed for the 2nd order model, measurement of the biogas flow rate (Q) and step changes of the influent organics concentration (γ_{S0i})*

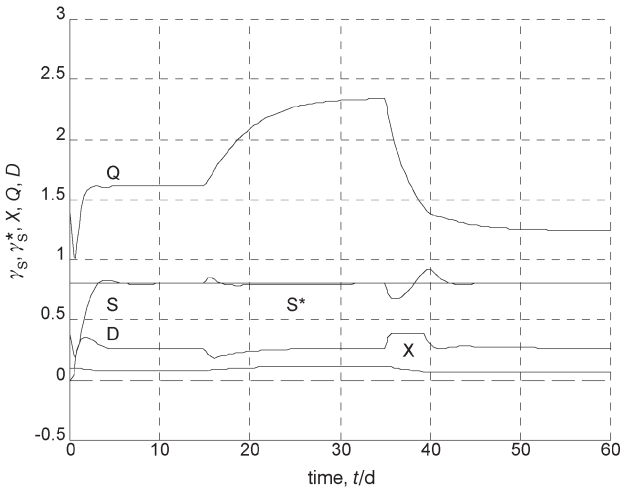
The compensator proves to be able to maintain biogas flow rate at desired set-points including at the maximum one, and to reject disturbances in a large range.

The same experiments have been made with the compensator designed for the model (2), (3), (4), (5) and S(t) as a measurable output (Fig. 8 and Fig. 9) . In this case the regulation interval has been augmented due to the lack of maxima in the characteristic $\gamma_S = \gamma_S(D)$.

Simulation experiments have been carried out for compensators designed for the 2nd-order model (2), (3), (4), (5) but controlling the 5th order model (6), …, (12) (outputs $Q(t)$ and $\gamma_S(t)$). In the case of $Q(t)$ being a measurable output, good performances



F i g . 8 – *Simulation results for dynamic compensator control, designed for the 2nd order model, measurement of the and effluent organics concentration (γ_S) and step changes of the set point (γ_S*)*



F i g . 9 – *Simulation results for dynamic compensator control, designed for the 2nd order model, measurement of the and effluent organics concentration (γ_S) and step changes of the influent organics concentration (γ_{S0i})*

have been obtained in the presence of bounds on the control input derivative:

$$-\delta D \leq d\,D(t)/\mathrm{d}t \leq \delta D, \qquad (23)$$

where δ D is a positive constant (*Bastin* and *Dochain*,[3] 1991).

In the case of $\gamma_{S3}(t)$ being a measurable output (*Bastin* and *Dochain*,[2] 1990), the compensator designed for the 2-nd order model is directly applicable to the 5th order model (Fig.10).

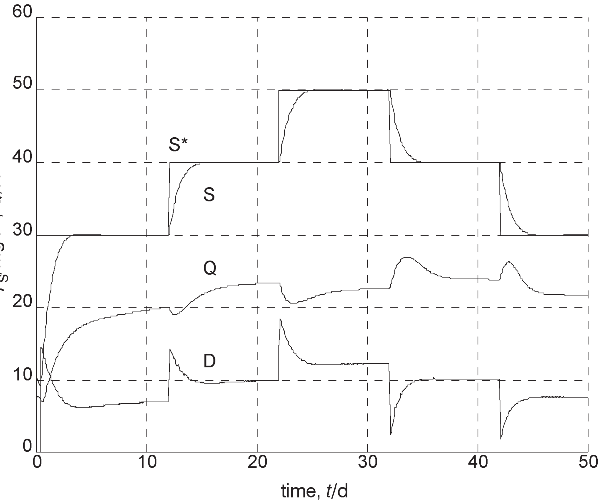


Fig. 10– *Simulation results for dynamic compensator control, designed for the 5th order model, measurement of the and effluent organics concentration (γ_S) and step changes of the set point (γ_S*)*

Conclusion

An appropriate non-linear model of methane fermentation for controller design purposes has been adopted and its coefficients estimated on the basis of a simple methodology for non-linear identification. A set of lab-scale data allow the parameters identification and model validation by comparing simulation results with experimental data obtained for different experimental conditions. For the design purposes, the non-linear 2nd-order models with inputs Q or γ_S have been transformed into linear ones with interval parameters. In both cases, controllers with dynamic output linear compensators have been designed according to the internal model principle. The effectiveness of those algorithms has been illustrated by simulation experiments with step changes of the set-points (Q^* and γ_S^*) and disturbances (γ_{S0i}) in both cases (regulation of biogas production rate Q or the concentration of the outlet polluted organic matter γ_S) for 2nd and 5th order non-linear models of the process. Important features of the proposed algorithms are their robustness and simple realisation.

ACKNOWLEDGEMENTS

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List of symbols

(in this paper we use units popular in the field of the biotechnology)

t – time, d
γ_{S0i} – influent organic mass concentration, g L^{-1}
Q – methane production rate, L d^{-1}
D – dilution rate, d^{-1}
F – flow rate, L d^{-1}
x – state vector
y – output
Δ – deviation from the nominal value

For the model based on a one-stage reaction scheme:

γ_S – the effluent organics mass concentration, g L^{-1}
γ_X – concentration of methanogenic bacteria, g L^{-1}
μ – specific growth rate, d^{-1}
k_1, k_2, k_s and μ_m are coefficients.

For the model based on a three-stage reaction scheme:

γ_{X1} – concentration of acidogenic bacteria, g L^{-1}
γ_{X2} – concentration of methanogenic bacteria, g L^{-1}
μ_1 – specific growth rate of acidogenic bacteria, d^{-1}
μ_{1max} – maximum specific growth rate of acidogenic bacteria, d^{-1}
μ_2 – specific growth rate of methanogenic bacteria, d^{-1}
μ_{2max} – maximum specific growth rate of methanogenic bacteria, d^{-1}
k_1 – decay coefficient for acidogenic bacteria, d^{-1}
k_2 – decay coefficient for methanogenic bacteria, d^{-1}
K_{S1}, K_{S2} – saturation constants, g L^{-1}
γ_{S0} – concentration of soluble organics, g L^{-1}
γ_{S1} – substrate concentration for acidogenic bacteria, g L^{-1}
γ_{S2} – substrate concentration for methanogenic bacteria, g L^{-1}
Y_1, Y_2 – yield coefficients, g L^{-1}
β – coefficient, dm^3 g^{-1} d^{-1}
Y_B, Y_P – coefficients;
Y_g – coefficient, g L^{-1}
γ_{S3} – concentration of the outlet polluted organic matter, g L^{-1}
COD – Chemical Oxygen Demand;

Subscripts

i,j – number indices

Superscripts

^{0} – nominal value
^{T} – transpose symbol

References

1. *Angelidaki, I., Ellegaard, L.* and *Ahring, B.*, Biotechnology and Bioengineering, **63** (1999) 363.
2. *Batstone, D. J.* et al., The IWA anaerobic digestion model No 1 (ADM1), (th Word Congress "Anaerobic digestion 2001", Antwerpen, Sept. 2–6, 2001.
3. *Bastin, G., Dochain, D.*, *On-line estimation and adaptive control of bioreactors*, Elsevier Science Publishers, Amsterdam and N. Y. 1991.
4. *Dochain, D.* (Editeur): Automatique des bioprocedes, HERMES Science Publications, Paris, 2001.
5. *Dubrovskis, V.*, *Viesturs, U.*, "Anaerobic digestion of agricultural waste". *Zinathe Publishing House, Riga, (in Russian),* 1988.
6. *Hill, D. T., Barth, C. L.*, "A dynamic model for simulation of animal waste digestion". *J. Wat. Pol. Centr.Fed.,* **10** (1977) 2129.
7. *Price, E. C., Cheremisinoff, R. N.*, "Biogas-production and utilization". *Ann Arbor Science Publ.*, 1981.
8. *Schmitendorf, W. E., Barmish, B. R.*, Automatica. **22** (1981) 355.
9. *Simeonov, I.,* "Modelling and control of Anaerobic Digestion of Organic Waste". Chem. Biochem. Eng., **8** (1994) 45.
10. *Simeonov, I., Momchev, V., Grancharov, D.,* Wat.Res., **30** (1996) 1087.
11. *Simeonov, I.,* Methodology for parameter estimation of non-linear models of anaerobic wastewater treatment processes in stirred tank bioreactors, *5th Int. Symp. Systems analysis and computing in water quality management-WATERMATEX 2000*, Gent, Belgium, September 2000 pp. 18–20, 8.40–8.47.
12. *Stoyanov S.*, "Synthesis of Multiloop Systems with Disturbances Rejection". Ph.D. Thesis, Sofia Technical Univ. 1989.
13. *Stoyanov, S., Simeonov, I.*, Bioprocess Eng., **15** (1996) 295.
14. *Tzonkov, S., Filev, D., Simeonov, I., Vaklev, L.*, Control of biotechnological processes, *Tehnika*, Sofia, (in Bulgarian), 1992.
15. *Van Impe, Jan F. M., Vanrolleghem P. A., Iserentant D. M.*, *Advanced Instrumentation, Date Interpretation and Control of Biotechnological Processes*, Kluwer Acad. Publ., 1998.