INTEGRATED CLOSED LOOP CONTROL AND PROCESS DESIGN

Gloria Gutiérrez[1], Pastora Vega[2], César de Prada [1]

[1]*Dpt. Systems Engineering and Automatic Control, University of Valladolid, Spain*
[2]*Dpt. Computing and Automatic Control, University of Salamanca, Spain*

Abstract: The joint design of processes and control is an integration technique that incorporates control requirements in the process design stage. This paper presents an integrated design methodology and illustrate it with an alternative design for a real plant of activated sludge processing. The process design obtained is not only optimum from the economic point of view, but exhibits certain previously fixed dynamic characteristics. The point of operation and the size of the processing units are obtained through mathematical programming, together with the parameters of the controller.

Keywords: integrated process and control design, process optimization, process synthesis, process modelling.

1. INTRODUCTION

Traditional plant design is centred on determining the structure, size and other variables associated to the processing units, in a chosen stationary state as required for the production or processing of specific materials or products. In general, in the design stages, although optimization is used, the dynamic behaviour of the process is not taken into account and the cost function to be optimized includes only terms associated to building up or operating costs.

Once the plant has been designed, the control engineers concentrate on established patterns of operation and can start the design of automatic control systems that ensure the correct operation of these plants, even in the presence of changes and disturbances. This task is often difficult or even impossible as the original design did not take control requirements into account. The adequate functioning of the process, together with its automatic control system, does not depend exclusively on the kind of controller and its parameters, since the control of a process is conditioned by its own design (Luyben and Floudas, 1994).

Integrated process and control design is a plant design technique in which the control characteristics are considered in the design stages, thus permitting dynamic specifications to be fixed for the system to ensure flexibility and ease of operation. By applying this design methodology, the physical parameters of the plant can be obtained, that is, the optimum plant size to minimize investment and operating costs while guaranteeing the dynamic behaviour. (Morari, 1983) have made significant contributions to control analysis and dynamic adaptability of systems, introducing and analysing control magnitudes for the interaction of the variables and the rejection of disturbances.

Integrated control and process design can be considered in open or closed loop. The first one aims at obtaining plants that are easier to operate. The solution obtained when solving the problem of design and control in open loop, are the physical parameters of the plant associated to the lowest values of a cost function that, at the same time, provide a desired dynamical behaviour of the system and comply with the design requirements of the process. In this case, in the design stage, the process is not considered to be governed by a controller, but the optimum design obtained should verify the desired dynamical conditions for the system, facilitating the task of control implementation.

On the other hand, closed loop design aims at obtaining process plus control systems that, in closed loop, give better performance. The solution to the problem of design and control in closed loop, besides providing the optimum physical parameters of the plant, also provides the parameters of the controller. That is, the controller synthesis problem is also introduced into the problem of the plant design. Along with the size of the processing units, the parameters of the controller are also obtained, which should guarantee the correct closed loop functioning of the process, satisfying the control requirements, even if the process is not operable in open loop.

The most commonly used control magnitudes in open loop were introduced by (Skogestad and Wolf, 1992; Morari, 1992): disturbance condition number and the combination of the Bristol relative gain matrix dependent on the frequency and disturbance gain of the system in closed loop, adding other magnitudes known as the system's condition number, the singular values, etc. In the cases where closed loop design is the choice, the use of PID controllers or predictive controllers based on plant models is frequent.

This paper presents a design methodology applied to an industrial process of activated sludge, considering its dynamic behaviour in closed loop.

The process, an urban wastewater treatment plant located in Manresa, Spain, is described in section 2, and has been chosen as a reference for this work. The integrated design methodology is explained in section 3, which includes the simultaneous use of the linear and non-linear models of the process in a framework of optimization. Next, in section 4, the proposed methodology is applied to the design of the waste water treatment plant, while section 5 provides results comparing the design obtained without considering control requirements and the one obtained using integrated design techniques in closed loop. Finally, some conclusions and references are provided.

2. DESCRIPTION AND MODEL OF THE PLANT

A diagram of the plant used in this work is shown in Figure 1. An inflow q_i with organic load s_i (measured as DQO) and residual biomass concentration x_i, after an initial treatment, is sent to the bioreactor containing micro-organisms (biomass), where the biological treatment takes place. The aim of the process is to eliminate the substrate (organic load) from the water, which is done by means of the biomass feeding on the substrate, thus removing them from the water. As the process is aerobic, there are some aerators which, when working, force the oxygen from the air to become mixed with the water, thus activating the biological reactions.

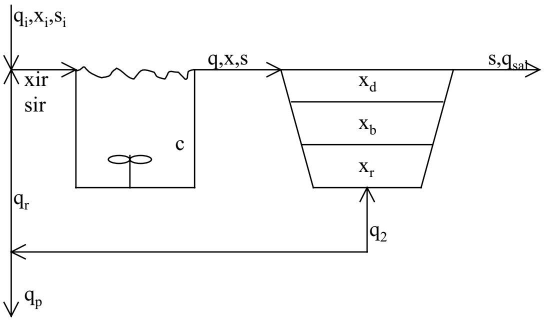

Fig. 1. Activated Sludge Process

The process is continuous and the treated product flows through a channel into a secondary clarifier. Here the micro-organisms are separated from the waste water while foam and other detritus are eliminated from the surface. The micro-organisms accumulate as sludge in the bottom by sedimentation and the clean water is taken off at the surface. In order to recuperate the micro-organisms dragged along with the sludge and to maintain the population of the bioreactors, the sludge is partly recycled to the bioreactors, the rest being eliminated by draining. Both, the recycled flow of sludge and the drained sludge, can be measured by means of the corresponding flow meters. A three-way-valve allows the drainage fraction to be changed.

The treatment process can also eliminate nitrates, phosphorous, potassium, sulphites, etc. contained in the waste water. For the sake of simplicity, such treatment has not been considered in the paper, but the design methodology shown here can be easily extended to this kind of process.

2.1 Process requirements

Due to its biological nature, and because it is a process with a great environmental impact, certain specific conditions should be met for its correct functioning:

The mass load must be within certain limits depending on the average flow to be treated. The time the sludge remains in the clarifier should be between 3 and 10 days. The residence time of the water in the bioreactors should be between 2 and 5 hours. The concentration of micro-organisms in the bioreactor should be between 2,000 and 5,000 mg/l.

Other operating conditions are regulated by environmental regulations: the percentage of the recycling flow with respect to the total flow, the hydraulic load capacity of the clarifiers, the concentration of the substrate in the clarified water that should be below a maximum allowed value, etc.

In addition to complain with the above constraints, an optimum design of a waste water plant will aim at minimising costs associated to both initial investments and operation of the plant and would normally be formulated as follows:

$$\min_{V,A,fk} \alpha_1 V + \alpha_2 A + \alpha_3 fk \quad (1)$$

Where the index to be minimized (1) includes the investment costs determined by the volume of the reactor and the surface of the tank (*V* and *A*) and the operating cost, expressed as a function of the aeration factor in the tank (*fk*) where $\alpha_1, \alpha_2, \alpha_3$ are the associated unit costs.

Subject to a set of constraints:

a) Fulfilment of the static nonlinear model equations in the optimal design point:

$$0 = \mu y \frac{sx}{(k_s + s)} - k_d \frac{x^2}{s} - k_c x + \frac{q}{V}(x_{ir} - x) \quad (2)$$

$$0 = -\mu \frac{sx}{k_s + s} - f_{kd} * k_d \frac{x^2}{s} - f_{kd} * k_c x + \frac{q}{V}(s_{ir} - s) \quad (3)$$

$$0 = k_{la} fk (c_s - c) - OUR - \frac{q}{V} c \quad (4)$$

$$0 = (q_{sal} x_b - q_{sal} x_d - Avs(x_d)) / Al_d \quad (5)$$

$$0 = (qx - q_{sal} x_b - q_2 x_b + Avs(x_d) - Avs(x_b)) / Al_b \quad (6)$$

$$0 = (q_2 x_b - q_2 x_r + Avs(x_b)) / Al_r \quad (7)$$

$$xir = \frac{(x_i q_i + x_r q_r)}{q} \quad (8)$$

$$sir = \frac{(s_i q_i + s_r q_r)}{q} \quad (9)$$

$$OUR = k_{01} \mu x \frac{x}{(k_s + s)} \quad (10)$$

$$vs(x) = nnrx \exp(aarx) \quad (11)$$

$$q_{sal} = q - q_2 \quad (12)$$

$$q_r = q_2 - q_p \quad (13)$$

$$q = q_i + q_r \quad (14)$$

Here, equations 2-4 represent the mass balances of the biomass, the substrate and the oxygen in the reactor respectively. Equations 5, 6 and 7 describe the mass balances of the biomass in three different layers of the clarifier. Equations 8 and 9 express the input of biomass and substrate respectively at the entrance to the bioreactor. Equation 10 is associated to the oxygen consumed in the reactor. The remaining equations establish the balances of the flows.

b) Process constraints:

Residence time in the bioreactor:

$$2.5 \le \frac{V}{q} \le 5 \quad (15)$$

Mass load in the bioreactor:

$$0.03 \le \frac{qs + q_r s}{Vx} \le 0.06 \quad (16)$$

Sludge age in the clarifier:

$$3 \le \frac{Vx + Al_r x_r}{q_p x_r 24} \le 10 \quad (17)$$

Constraints on the Recycled flow rate:

$$0.5 \le \frac{q_2}{q_i} \le 0.7 \quad (18)$$

Constraints on the Purge flow rate:

$$0.03 \le \frac{q_p}{q_2} \le 0.06 \quad (19)$$

Hydraulic capacity in the clarifier:

$$\frac{q}{A} \le 1.5 \quad (20)$$

Plus upper and lower limits of the design variables.

The nominal operating conditions for which the plant was designed are: inflow to the plant q_i = 1300 m^3/h, inflow substrate content s_i = 366 mg/l and inflow biomass content x_i = 80 mg/l. The kinetic coefficients of the equations were calculated with registers of data from a real process, minimizing the modelling errors, (Moreno, 1994): μ=0.1824, y = 0.5948, k_d = 5.0e-5, k_c =1.3333e-4, k_s=300 y f_{kd}= 0.2.

2.2 *Control problems*

Once the plant has been designed and built it needs to operate within a certain range in spite of periodic disturbances that this kind of plant is prone to and which were not taken into account at the design stage.

The variables to be controlled are the substrate *s* at the process outlet and the biomass concentration *x* in the reactor. The recycling flow q_2 and the purge flow q_p can be used as the main manipulated variables. Notice that, as the liquid overflows the reactor and clarifier, no level controller is required. Also, the oxygen concentration in the reactor is only required to be above a certain level, so, it won't be included in the integrated design.

The disturbances are due mainly to the different loads generated during the day and night, or at weekends compared to the rest of the week. Another important source of disturbances is associated with rain days. Figures 2 and 3 show real data corresponding to the variations in inflow and substrate concentration for one week of operation of treatment plant used as the reference. The inflow appears in m^3/h while the substrate is given in mg/l and the time scale in hours. In addition, the settling time respect to changes in the input variables is usually very large.

As it is impossible to predict exactly what the inflow will be or how much contamination it will bring, and these values move away from the design operating point over considerable periods of time, the control task is a difficult one. Not to mention the biological nature of the process.

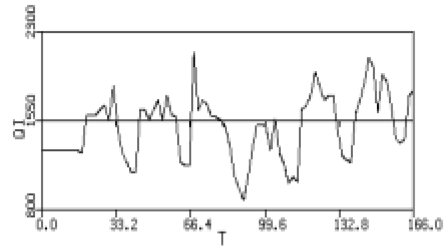

Fig. 2. Inflow to the plant

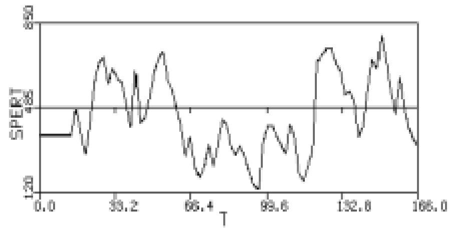

Fig. 3. Inflow substrate content

3. INTEGRATED DESIGN TAKING CONTROL REQUIREMENTS INTO ACCOUNT

As already mentioned, the dynamic and control characteristics are not normally taken into account explicitly in process design problems. Traditional design methods ignore the idea that changes in the process can make the system more controllable (Luyben, 1993). A more adequate design procedure is to consider not only economic criteria, but to include the controllability of the process at the same time.

In this paper, controllability is understood in the sense of operability as "the facility of the process to function at a point of operation and to reject disturbances while, at the same time, taking into consideration the necessary control effort".

Several works have been published in recent years that present and describe control magnitudes in open loop (Luyben & Floudas 1994). They can also be extended to systems in closed loop and, in fact, what is required is a closed loop design able to fulfil the performance and costs requirements, even if the open loop system has complex dynamics. Integrated process and control design in closed loop tries to find the sizing for the process in such a way that the investment and operating costs are kept to a minimum while, at the same time, the constraints and the dynamic requirements are satisfied by the control systems which is designed simultaneously.

In order to be able to include the dynamic characteristics, there are several alternatives. One of them is to add a term to the cost function so that the problem could be formulated as follows:

$$\min \alpha_1 V + \alpha_2 A + \alpha_3 fk + \beta \int e^2 dt \qquad (21)$$

Where *e* is the controller error. Notice that, in order to evaluate the error, the dynamic version of the model (2)-(14), has to be included by the other constraints, together with the controller equations. Also, the decision variables must contain, not only the ones of the process sizing, but also the parameters of the controllers.

Assignment of a value to the weight β it is not easy due to the non-economic nature of the error term. One solution is to derive an economic value for the error. Alternatively, the error term can be added as a constraint that fix a limit to the maximum error that can be tolerated as: $\int e^2 dt \le \beta$.

This formulation leads to a non-linear dynamic optimization problem. It can be solved using a sequential approach: that is, computing the cost function by integration of the system equations.

Alternatively, the closed loop design can be formulated as an expansion of the problem presented in section two, in which the decision variables and the constraints are extended to take into account the operability of the process in closed loop, but keeping it as one of stationary optimization, and avoiding the use dynamic optimization:

The key point in this approach is how to express the operability of the system in closed loop in a compact way. For this, the dynamic behaviour indexes must be expressed as a function of the physical parameters, since they depend on these latter parameters.

The integrated process and control design methodology used in this work has the following steps:

i. Propose a structure for the process.
ii. Formulate the mathematical (non-linear) model of the proposed structure.
iii. Linearize the model at a *generic* operating point so that the linear model matrices are functions of the parameters of the plant.
iv. Propose a control structure and calculate the corresponding closed loop transfer matrices.
v. Formulate the operability in terms of the transfer matrices in closed loop as inequality equations.

Then the optimization of the cost function (1) is performed considering **simultaneously** the constraints given by the non-linear model and the operability conditions calculated from the generic linear model. The solution provides the stationary operating point and the plant sizing (volume and area), as well as the tuning parameters of the controllers. Once the designs have been obtained, simulated experiments can be carried out in closed loop in order to check the optimal design.

In the example presented in this paper, the control objectives are considered to be: i) to reject disturbances at the most significant frequencies, ii) to decrease the settling time and iii) to stabilize the system in closed loop.

As for rejecting disturbances, we attenuate the dominant frequency components present in them. A measure of the rejection of perturbations in closed loop will be the matrix gain in closed loop at these dominant frequencies. In order to characterize the system stability in closed loop, as well as the settling time, the eigenvalues of the system state matrix in closed loop are used, which should to be placed in an appropriate region of the negative part of the *s*-plane.

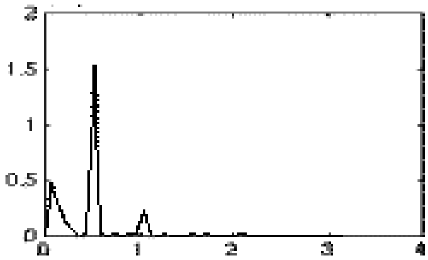

Fig. 4. The inflow signal spectrum

To determine the dominant frequencies of the main disturbances, the spectrums of real signals taken from the plant like the ones of fig. 2 and 3, were computed which are shown in figures 4 and 5. It can be seen that the large peaks are to be found in the frequencies 0.5 and 0.1.

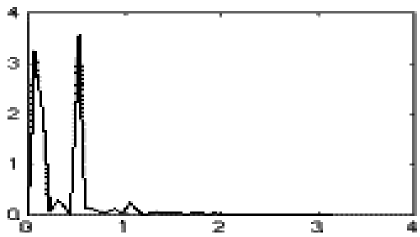

Fig. 5. The substrate signal spectrum

4. FORMULATION OF THE INTEGRATED DESIGN PROBLEM APPLIED TO THE ACTIVATED SLUDGE PROCESS

Steps i and ii of the proposed methodology have already been presented and correspond to the structure shown in Figure 1 and the mathematical model expressed in equations 2-14.

The next task is the model linearization, which is obtained through symbolic derivation with respect to the control variables and the disturbances considered, using the mathematical package MAPLE.

A key point in the methodology presented is the selection of the control structure. The problem can, of course, be generalized by considering a control superstructure and extending the optimization problem, which would become a MINLP one. Nevertheless, in our case, for the sake of simplicity, a simple control structure, as seen in Figure 6, has been proposed. It consists of a PID controller where the controlled variable is the substrate at the outlet of the plant *s*, and the manipulated variable is the recycled flow from the clarified, q_2. The state space model of the controller is shown below. Besides the generic linearized plant model, it can be used to compute the closed loop linear model.

$$A_r = [0]\ ;\ B_r = [1]\ ;\ C_r = \left[\frac{k_p}{t_i}\right];\ D_r = \lfloor k_p \rfloor \quad (22)$$

Here k_p is the controller gain and t_i the integral time.

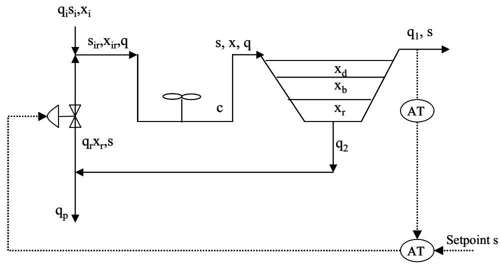

Fig. 6. Process control scheme

Then, the following constraints related to the dynamics in closed loop can be considered:

a) Attenuation of the disturbances substrate and inflow to the plant, at the system's dominant frequencies of 0.5 and 0.1. This constraints aims to achieve a lower range of variation of the substrate at the outlet in closed loop with respect to the changes in open loop. The magnitude of rejection is fixed through $L_{sup.}$

$$\frac{|\psi_{cl}|}{|\psi_{al}|} = \frac{\left|C_{cl}\left(j0.5I_{cl} - A_{cl}\right)^{-1}B_{cl}\right|}{\left|C_{ml}\left(j0.5I_{ml} - A_{ml}\right)^{-1}B_{ml}\right|} < L_{sup} \quad (23)$$

$$\frac{|\psi_{cl}|}{|\psi_{al}|} = \frac{\left|C_{cl}\left(j0.5I_{cl} - A_{cl}\right)^{-1}B_{cl}\right|}{\left|C_{ml}\left(j0.5I_{ml} - A_{ml}\right)^{-1}B_{ml}\right|} < L_{sup} \quad (24)$$

$$\frac{|\zeta_{cl}|}{|\zeta_{al}|} = \frac{\left|C_{cl}\left(j0.1I_{cl} - A_{cl}\right)^{-1}B_{cl}\right|}{\left|C_{ml}\left(j0.1I_{l} - A_{ml}\right)^{-1}B_{ml}\right|} < L_{sup} \quad (25)$$

$$\frac{|\zeta_{cl}|}{|\zeta_{al}|} = \frac{\left|C_{rl}\left(j0.1I_{rl} - A_{rl}\right)^{-1}B_{rl}\right|}{\left|C_{ml}\left(j0.1I_{l} - A_{ml}\right)^{-1}B_{ml}\right|} < L_{sup} \quad (26)$$

Where ψ_{al} y ζ_{al} are the transfer matrices between the substrate and the disturbances in open loop at the frequencies 0.5 and 0.1 radians/hours respectively, and ψ_{cl} y ζ_{cl} are the transfer matrices between the substrate and the disturbances in closed loop at the frequencies 0.5 and 0.1 radians/hours respectively. The matrices A_{ml}, B_{ml}, C_{ml}, A_{cl}, B_{cl} y C_{cl} correspond to the matrices in open and closed loop respectively, of the state space model. Notice that all of them are formulated for a generic linearization point, hence, they depend on the plant sizing and controller parameters..

b) Stability and dynamics of the system in closed loop. Fixing the position of the state matrix eigenvalues of the system in closed loop in the plane *s* within a region M in the left hand side, see Fig.7. The type of desired response and the settling time are also fixed with these constraints:

$$\sigma \in M \quad (27)$$

Where σ represents the eigenvalues of the state matrix in closed loop and *M* is the region where they should be placed.

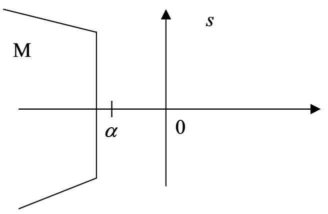

Fig.7. Allowed region for the closed system eigenvalues

The integrated design and control problem, thus formulated, is non-linear programming one that was solved with a sequential quadratic programming algorithm implemented on Matlab.

5. SIMULATION RESULTS

Simulation experiments in closed loop with the PID parameters and the sizing of the plant obtained from the optimal design were carried out in the presence of the same disturbances measured in the reference plant and given in Fig.2 and 3. Also, for comparison, a PID was applied to the original reference plant, and the results obtained were as in Fig.8 and 9.

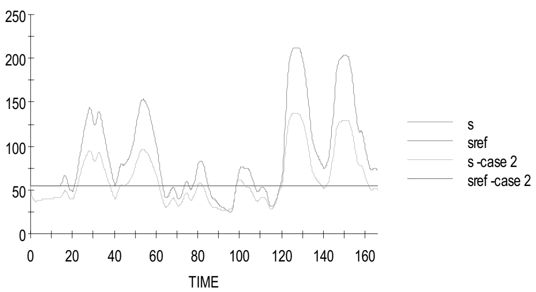

Fig.8. Substrate (mg/l) evolution in hours. Original plant (red) and the integrated design case (green)

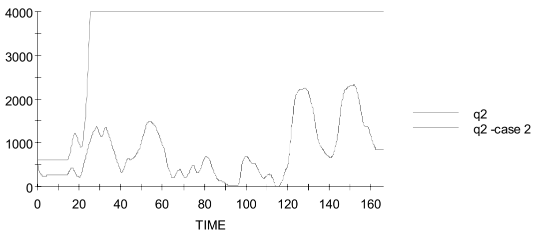

Fig.9. Recirculation flow (m^3/h). Original plant (red) and the integrated design case (brown).

As we can see, there is a significant attenuation, as given by constraints (23) to (26) of the disturbances on the output substrate over the original design. In fact, this controller saturates the control signal so that no effective regulation takes place.

Table I provides values of the plant sizing and controller tuning corresponding to the original plant and the one obtained by integrated design.

Table I. Plant sizing and controller parameters

Variable	Original plant	Integrated Design
Reactor volume (m^3)	7268	11958
Decanter surface (m^2)	2770	2917.3
Controller gain	-4.88	-30
Integral Time	0.2	100

6. CONCLUSIONS

The problems of integrated design, incorporating control characteristics in both open and closed loop, are complex. This is mainly due to the size of the problem itself and the non-linearity of the equations it consists of. Mathematical programming techniques do not always guarantee a global solution to the problem and the calculation time can be large. nevertheless, it is worth to make a couple of remarks: One, that the design is done off line, so the calculation time is not an important factor when applying this design methodology. Another point is that evolutionary algorithms for optimization are bringing about satisfactory results in the search for global optimums to complex and decidedly non-linear problems. In fact, in (Moles et. al 2002), a comparative study of various global optimization algorithms are discussed which uses stochastic and hybrid techniques with deterministic optimization, where, among other examples, the problem formulated here is solved with satisfactory results.

REFERENCES

Luyben, M. L., (1993). Analyzing the Interaction Between Design and Control In Chemical Reactor Systems. *Ph.D. Thesis*, Princeton University.

Luyben, M. L., C. A. Floudas, (1994). Analyzing the interaction of design and control –1. A multiobjective framework and application to binary distillation synthesis. *Comp. Chem. Eng.*, **18**, Nº. 10, 933-969.

Luyben, M. L., C. A. Floudas, (1994). Analyzing the interaction of design and control –2. Reactor-Separator-Recycle System. *Comp. Chem. Eng.*, **18**, Nº 10, 971-994

Moles, C. G., G. Gutiérrez, A. A. Alonso and J. R. Banga,(2003). Integrated process design and control via global optimization: a wastewater treatment plant case study. *Chem. Eng. Res. & Des.,* **81**, 507-517.

Morari, M. (1983). Design of resilient processing plants-III. A general framework for the assesment of dynamic resilience. *Chem. Eng. Sci.*, **38**, 1881-1891.

Morari, M. (1992). Effect of Design on the Controllability of Chemical Plants. *IFAC Workshop on Interactions Between Process Design and Process Control*, J. D. Perkins, De., Pergamon Press, 3-16.

Moreno R. O. (1994) Estimación de estados y Control Predictivo del Proceso de Fangos Activados. *Ph. D. Thesis*. Universidad Autónoma de Barcelona.