

MODELING, SIMULATION AND ANALYSIS OF ETHANOL FERMENTATION PROCESS WITH CONTROL STRUCTURE IN INDUSTRIAL SCALE

Marcus Vinícius Americano da Costa Filho, mvamericano@lepten.ufsc.br

Julieta Barbosa Monteiro, julieta@lepten.ufsc.br

Felipe Costa Magazoni, magazoni@lepten.ufsc.br

Sergio Colle, colle@emc.ufsc.br

Laboratory of Energy Conversion Processes and Energy Technology - LEPTEN
Department of Mechanical Engineering - Federal University of Santa Catarina
88040-900, Florianópolis, Brazil

Abstract. *In this paper a complete model of a fed-batch ethanol fermentation process as a function of the temperature is presented. Also, a control structure using a heat exchanger and PID controllers with anti-windup is proposed. The control methodologies are tested using a nonlinear fermentation model and a 200 order heat exchanger model which can be considered as the real system. Identification techniques of the System Identification Toolbox for MATLAB were used to design the controllers and approximate the real situation. The volume of the reactor is 1000m^3 and the controlled fluid flows can be up $1000\text{kg}\cdot\text{h}^{-1}$. The heat transfer coefficient and the length of the heat exchanger are $1\text{kW}\cdot\text{K}^{-1}\cdot\text{m}^{-2}$ and 2m , respectively. Thus, all of the parameters and dimensions are used in industrial scale to simulate the process.*

Keywords. *Fermentation, ethanol, heat exchanger, modeling, identification, PID control, simulation.*

1 INTRODUCTION

The commercial ethanol production in Brazil was given significant incentive with the creation of the governmental program known as *Proálcool*. Similar programmes in the USA and in Canada later followed the same strategy (Wheals et al. 1999). *Proálcool* initiated the technological developments which led to Brazil becoming a world reference. Nowadays, with the further increase in pretroleum prices and productivity improvements obtained by the plants, there is a great interest in the optimization of all stages of the ethanol production process (Costa et al. 2001).

Although there is an accentuated concern with the modernization of the stages of ethanol production, there are still some problems with respect to the control systems and a clear perspective of improvement in the performance of their processes.

Due to the importance of fed-batch processes in chemical and biotech process industries, several dynamic models which have been proposed by researchers are useful in testing modeling, monitoring, diagnosis and control techniques (Menezes et al. 1994; Çinar and Ündey 1999; Birol et al. 2002; Xiong and Zhang 2004).

Experiences in other plants show that the investment in automation and control quickly offset by the profits obtained in the production. In order to increase the ethanol productivity, a control structure able to maintain the fermentation process under optimal conditions is proposed. The choice of the control structure is an important step in the development of a control strategy.

In industrial ethanol fermentation the yeast cells are submitted to stresses inherent to the process, which are caused by environmental conditions and physicochemical factors such as high temperature, pH and high salt, ethanol and sugar concentrations. Among these factors, temperature is that which most affects the fermentation kinetics of the process and cell viability. At high temperatures, the specific growth rate of contaminant microorganisms is increased leading to a higher probability of contamination in the process. This contamination causes flocculation of the yeast cells, leading to problems in the centrifugation. On the other hand, at lower fermentation temperatures, the use of antibiotics is reduced due to the lower presence of contamination. The presence of high ethanol concentration combined with high temperatures submits the cells to stress conditions, where the yeast has a higher tendency to produce glycerol which is the main by-product of the alcoholic fermentation. Studies have shown that there is an ideal fermentation temperature for maintenance of cell viability, reducing the glycerol production and increasing the process efficiency (Atala et al. 2001; Aldiguiet et al. 2004). In fed-batch mode, another variable that must be considered to increase the performance of the process is the feed flow rate. How to feed the reactor to obtain greater quantity of desired product in the shortest possible time is a question that some researchers have addressed (Wang and Shyu 1997; Xiong and Zhang 2004; Santos et al. 2006). Therefore, it is also very important that the control structure allows manipulation of the feed flow rate.

The focus of this paper is the alcoholic fermentation stage. The models of heat exchanger and fermentation process in industrial scale are coupled in accordance with the proposed structure control. It is generated a new system which permits the analysis of the dynamic behaviour of the principal variables subject to level and temperature control.

This paper is organized as follows. In section 2, the models of the devices and processes that comprise the system are presented. In section 3, a design control procedure that can be used in industry is described. In section 4, the simulations are shown and the results discussed. The paper ends with the conclusions.

2 SYSTEM DESCRIPTION

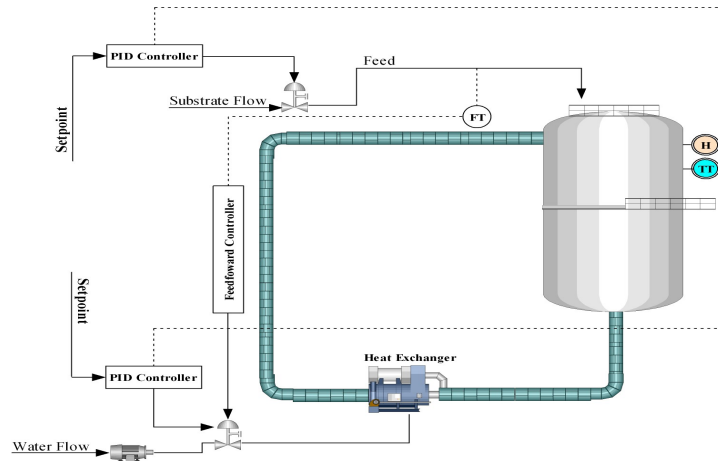


Figure 1: Control Structure

A control structure is proposed as shown in Fig. 1. In this section, the devices and processes that comprise the whole system are described as follow: heat exchanger, fermentation process and PID controllers.

2.1 Heat Exchanger

The heat exchanger is a thermal piece of equipment that allows the heat transmitted to enter two fluids with different temperatures in two adjacent chambers. Thus, to keep the process fermentation at the desired temperature, a heat exchanger is required.

The number of parameters and nonlinear dynamics make the modeling of a heat exchanger very difficult. Starting from a nonlinear and continuous model, a state space representation of the distributed parameter system is obtained using the "Direct Lumping of the Process" technique presented in Xia et al. (1991); Bonivento et al. (2001). Basically, this method subdivides the thermal exchange surface into Δx sections, that is, the state vector is defined by the temperature of the sections (see Fig. 2). Assuming that the velocity of the fluids across the heat exchanger is constant and neglecting the

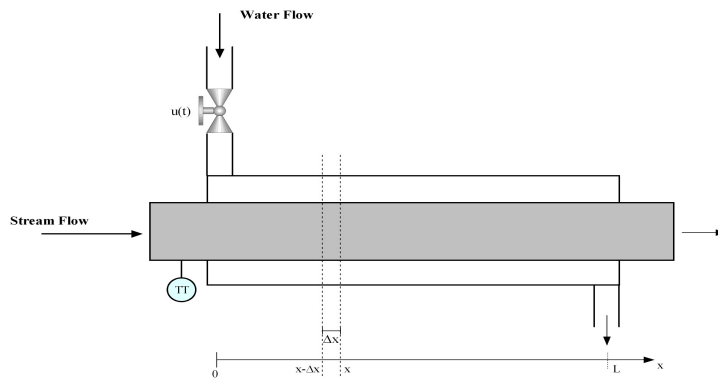


Figure 2: Heat exchanger

effects of the metal between the sections, the mass and energy balances can be calculated applying the Energy Conservation Principle to every lump (for time t , at point $x - \Delta x$), resulting in:

$$\frac{\partial}{\partial t}(Mc_p T) = \dot{m}c_p(T_{x-\Delta x} - T_x) - UA\Delta T \quad (1)$$

where $M = a\Delta x\rho$ the lump mass; $\dot{m} = av\rho$ the mass flow rate; a is the section of the chamber where the fluid flows; ρ is the fluid density; c_p is the specific heat capacity; T is the temperature of the fluid considered; t is the time; v is the fluid

velocity; ΔT is the temperature difference; Δx is the incremental distance; A is the surface of the space included in the lump considered and U is the overall heat transfer coefficient.

The heat exchanger is divided into N lumps so that the space model of the system refers to a state vector T representing the temperature of each lump. This vector can be subdivided into two parts, representing the temperature of each fluid:

$$T = [T_{r1}, T_{r2}, T_{r3}, \dots, T_{rN} \quad T_{f1}, T_{f2}, T_{f3}, \dots, T_{fN}]^T$$

where T_{rj} and T_{fj} are the must and water temperatures in the j th point room with $j = 0, 1, \dots, N$, respectively. T_{r0} and T_{f0} are considered as part of the inputs. Therefore, the following system with an order of $2N$ can be obtained:

$$\begin{cases} M_1 c_{p1} \frac{dT_{r,j}(t)}{dt} = \dot{m}_1(t) c_{p1} [T_{r,j-1}(t) - T_{r,j}(t)] - UA \Delta T(t) \\ M_2 c_{p2} \frac{dT_{f,j}(t)}{dt} = \dot{m}_2(t) c_{p2} [T_{f,j-1}(t) - T_{f,j}(t)] - UA \Delta T(t) \end{cases} \quad (2)$$

The parameters for the heat exchanger are:

$$\begin{array}{ll} N = 100 & a_1 = a_2 = 4.63 \times 10^{-2} m^2 \\ v_1 = 6 m/s & \rho_1 = \rho_2 = 1000 kg/m^3 \\ v_2 = 6u(t) m/s & c_{p1} = c_{p2} = 4.187 kJ/(K.kg) \\ U = 1 kW/(K.m^2) & T_{f0} = 22^\circ C \\ \Delta x = 0.02 m & A = 1.53 \times 10^{-2} m^2 \end{array}$$

The temperature of the water is constant and the length of the thermal exchanger is $L = 2m$. The must from the reactor enters in the heat exchanger with a constant mass flow rate given by $\dot{m}_1 = 10^6 kg/h$ (In this study, the time unit considered is h (hour) because in an industrial scale, the processes are normally very slow) and the water mass flow rate $\dot{m}_2 = 10^6 u(t)$ (kg/h), regulated by $u(t) \in [0, 1]$, is used as a manipulated variable to control the temperature of the must.

In a real situation, the static gain of the heat exchanger varies as a function of the input temperatures of the fluids. This important characteristic is therefore considered. In the test shown in Fig. 3, the temperature variation is approximately $3.7^\circ C$ for a must temperature of $30^\circ C$, while the temperature variation is approximately $8.3^\circ C$ for a must temperature of $40^\circ C$. In both cases, the input control signals are the same.

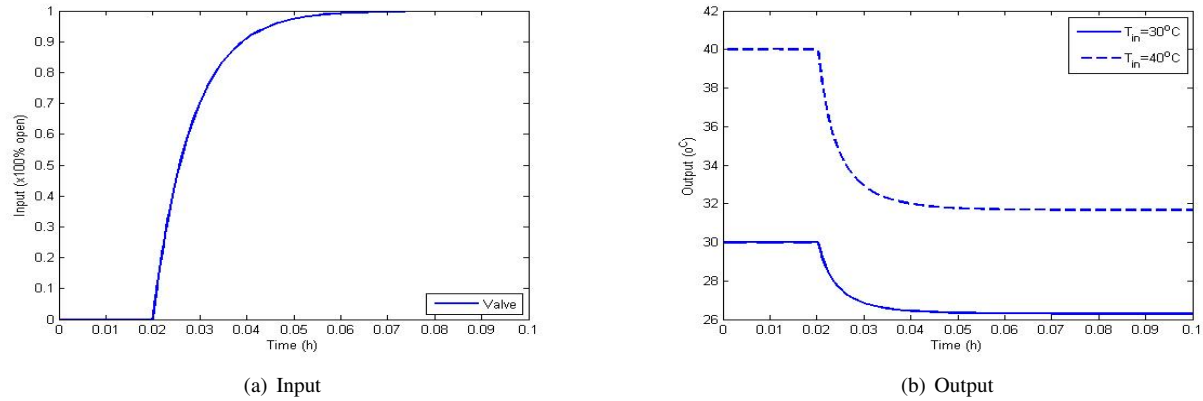


Figure 3: Temperature variations for different input must temperatures

2.2 Fermentation Process

The mathematical model for the ethanol fermentation is an unstructured model and the modeling was performed based on kinetic rate models coupled with mass balance equations for cell, substrate, ethanol and global energy balance, for an industrial fed-batch fermentation process. It is assumed that the feed is sterile, (i.e. the value of the inlet biomass concentration is zero).

The working volume variation during the fermentation process was performed by global mass balance and it is described as:

$$\frac{dV}{dt} = \dot{F} \quad (3)$$

where \dot{F} (m^3/h) is the substrate feed volume flow rate.

The cell growth rate is defined as follows:

$$\frac{dX}{dt} = \mu X - \frac{\dot{F}}{V}X - K_d X \quad (4)$$

where μ (h^{-1}) is the specific growth rate and K_d (h^{-1}) is the global coefficient of cell death. The factor \dot{F}/V (h^{-1}) is the dilution rate as the feed is added during the fermentation process.

The substrate consumption is modeled by the following equation:

$$\frac{dS}{dt} = \frac{\dot{F}}{V}(S_i - S) - \mu \frac{X}{Y_{X/S}} - m_X X \quad (5)$$

where $S_i = 200 \text{ kg/m}^3$ is the feed substrate concentration, $Y_{X/S}$ (kg/kg) is the yield factor of the biomass based on the substrate consumption and $m_X = 0.2 \text{ kg/(kg.h)}$ is the maintenance coefficient.

The ethanol formation is written as:

$$\frac{dP}{dt} = Y_{P/X} \mu X + m_P X - \frac{\dot{F}}{V}P \quad (6)$$

where $Y_{P/X}$ (kg/kg) represents the yield factor of the ethanol based on cell growth and $m_P = 0.1 \text{ kg/(kg.h)}$ is the ethanol production associated with cell growth.

The dead biomass concentration into the reactor is calculated as:

$$\frac{dX_d}{dt} = K_d X - \frac{\dot{F}}{V}X_d \quad (7)$$

where $K_d = K_{dT} \exp(K_{dP}P)$ (see Tab. 1).

Through the energy balance of the system, reactor and fermentation heat exchanger, the variation in the fermentation temperature during the process is described by:

$$\frac{dT_{r0}}{dt} = \frac{\dot{F}}{V}(T_i - T_{r0}) + \frac{\dot{m}_1}{\rho_1 V}(T_{rN} - T_{r0}) + \frac{\mu X}{Y_{X/S}} \frac{\Delta H_S}{\rho_1 c_{p1}} + \frac{m_X X \Delta H_S}{\rho_1 c_{p1}} \quad (8)$$

where ΔH_S is the heat released during the fermentation process the value of which is 678.3 kJ per kilogram of substrate consumed (Williams 1982) and $T_i = 28^\circ\text{C}$ is the feed temperature.

In the model shown, the specific growth rate μ is expressed as a function of the limiting substrate concentration S and of the inhibitory effects of the substrate, ethanol and cell concentrations, as represented by the following equation:

$$\mu = \mu_{max} \frac{S}{K_s + S} e^{-K_i S} \left(1 - \frac{P}{P_{max}}\right)^n \left(1 - \frac{X + X_d}{X_{max}}\right)^m \quad (9)$$

where $m = 1$ and $n = 1.5$ are constant values, μ_{max} (h^{-1}) is the maximum specific growth rate, $K_s = 4.1 \text{ kg/m}^3$ is the substrate saturation constant, K_i (m^3/kg) is the substrate inhibition coefficient, P_{max} (kg/m^3) is the ethanol concentration when cell growth ceases and X_{max} (kg/m^3) is the biomass concentration when cell growth ceases.

The parameters used in above equations have been described as functions of temperature by Atala et al. (2001) and their values are given in Tab. 1. These expressions were determined for temperatures ranging from 28 to 40°C, using the industrial yeast *Saccharomyces cerevisiae* and cane molasses as the substrate.

Table 1: Kinetic parameters as a function of fermentation temperature

Parameter	Expression
μ_{max}	$1.57 \exp(-41.47/T_{r0}) - 1.29 \times 10^4 \exp(-431.4/T_{r0})$
E_{max}	$-0.4421T_{r0}^2 + 26.41T_{r0} - 279.75$
X_{max}	$-0.3279T_{r0}^2 + 18.484T_{r0} - 191.06$
$Y_{X/S}$	$2.704 \exp(-0.1225T_{r0})$
$Y_{P/S}$	$0.6911 \exp(-0.0139T_{r0})$
$Y_{P/X}$	$0.2556 \exp(0.1086T_{r0})$
K_i	$1.393 \times 10^{-4} \exp(0.1004T_{r0})$
K_{dT}	$4 \times 10^{13} \exp\left(\frac{-41947}{1.987(T_{r0}+273.15)}\right)$
K_{dP}	$7.421 \times 10^{-3}T_{r0}^2 - 0.4654T_{r0} + 7.69$

2.3 PID Controller

The PID algorithm has been successfully applied in many applications. This success is a result of the many good features of the algorithm. The PID is used for single input-single output (SISO) systems, which have one controlled and one manipulated variable. Usually, many single-loop systems are implemented simultaneously on a process, and the performance of each control system can be affected by interaction with the other loops.

The control action $u(t)$ of an ideal standard PID controller can be computed as:

$$u(t) = K_c \left[e(t) + \frac{1}{T_i} \int_0^t e(\tau) d\tau + T_d \frac{de(t)}{dt} \right] \quad (10)$$

where $e(t)$ is the error between the reference and output signals. K_c is the proportional gain, T_i is the integral time and T_d the derivative time.

Applying the Laplace transforms in Eq. (10), the transfer function of the controller is:

$$C(s) = \frac{U(s)}{E(s)} = K_c \left(1 + \frac{1}{T_i s} + T_d s \right) \quad (11)$$

In practice, the ideal representation of the PID controller given in Eq. (11) cannot be implemented because the degree of the numerator is higher than the degree of the denominator, that is, $C(s)$ is improper. A low pass filter is normally used in cascade with the controller to make it proper, resulting in:

$$C(s) = \frac{K_c \left(1 + \frac{1}{T_i s} + T_d s \right)}{\alpha T_d s + 1} \quad (12)$$

with $\alpha \in [0,1]$. For industrial controllers, the value of α normally varies between 0.05 and 0.5. This parameter can be used to adjust the attenuation of noise and also the robustness of the closed loop.

3 CONTROL DESIGN

According to the above equations, the PID has a linear dynamic. Thus, a linear model of the process is required for the correct tuning of the PID. There are several ways to achieve this. However, one procedure that can be used in industry is described and analyzed.

3.1 Modeling of the System

The physical behavior of dynamic systems can be described by a set of differential equations. When this is not possible, modeling of the system modeling based on tests and experiments is used. This procedure is known as system identification and can be carried out, basically, in three steps (Ljung 1987): the data recording, the definition of the set of models and the determination of the best model in the set, based on the available data.

Using the *System Identification Toolbox for MATLAB*, the variables which must be controlled in the system are identified and modelled: the reactor level and temperature. The model chosen to represent the dynamic variable is the FOPDT (*First-order Plus Dead-time*) because this model is widely used to tune industrial controllers.

3.1.1 Temperature Model

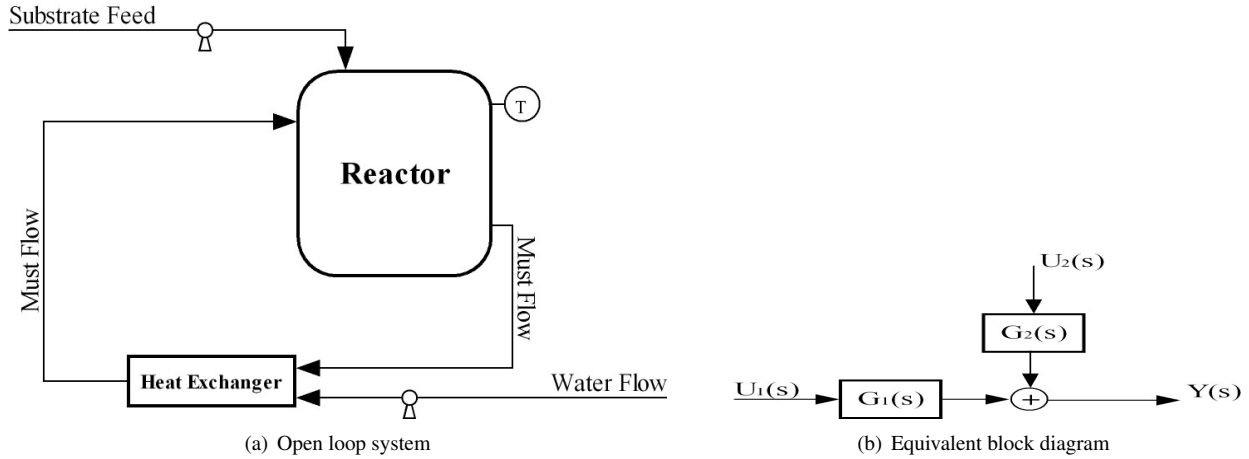
Consider the open loop system shown in Fig. 4(a). In general, after some manipulations, this type of system can be represented by an equivalent block diagram as shown in Fig. 4(b). In this representation, $U_1(s)$ is the opening of the water flow valve, $U_2(s)$ is the opening of the substrate feed valve and $Y(s)$ is the temperature of the system. $G_1(s)$ and $G_2(s)$ are transfer functions in the Laplace domain.

Thus, the temperature variable can be identified as the following model:

$$Y(s) = G_1(s)U_1(s) + G_2(s)U_2(s) \quad (13)$$

with $G_n(s) = \frac{K_n e^{-L_n s}}{\tau_n s + 1}$, $n = 1, 2$. Where K_n is the static gain, τ_n is the constant time and L_n is the delay time of the equation n .

Usually, a model is estimated by the step test data from the open loop process and it is used to design PI and PID controllers (Aström 1995). This procedure is widely used in industry. As the model is linear, it can be analyzed through the superposition principle, that is, the total response at a given time resulting from two or more signals is the sum of the responses which would have been resulted from each signal individually. For temperature modeling, a step signal is applied to the inputs of the system.



With $U_2(s) = 0$, several step tests are carried out to identify $G_1(s)$ as shown in Fig. 3.1.1. It is common that the reading and processing of signals inside an industrial environment are susceptible to noise. Therefore, a white noise signal is considered for the output system with a maximum magnitude of the power spectral density of 10^{-2} . The identified models as shown in Fig. 4(c) and Fig. 4(d) are $G_{11}(s) = \frac{-16e^{-0.015s}}{0.9s+1}$ and $G_{12}(s) = \frac{-8e^{-0.02s}}{0.71s+1}$, respectively.

Note that, according to the static gain concept, $K = \frac{\Delta Y}{\Delta U}$, the gain of the system is not constant because the temperature tends to the same final value which is approximately the water flow temperature, independently of the input variable $U_1(s)$. This is due to the fact the heat release is very small and tends to zero, since $U_2(s) = 0$, and there is coolant circulation in the system.

After some step tests, $G_1(s)$ can be modelled as:

$$G_1(s) = \frac{K_1 e^{-0.02s}}{0.8s + 1}, \quad -40 \leq K_1 < 0. \tag{14}$$

Similarly, $G_2(s)$ is identified with $U_1(s) = 0$. The models found as shown in Figs 5(a), 5(b) are $G_{21}(s) = \frac{88e^{-0.185s}}{0.615s+1}$ and

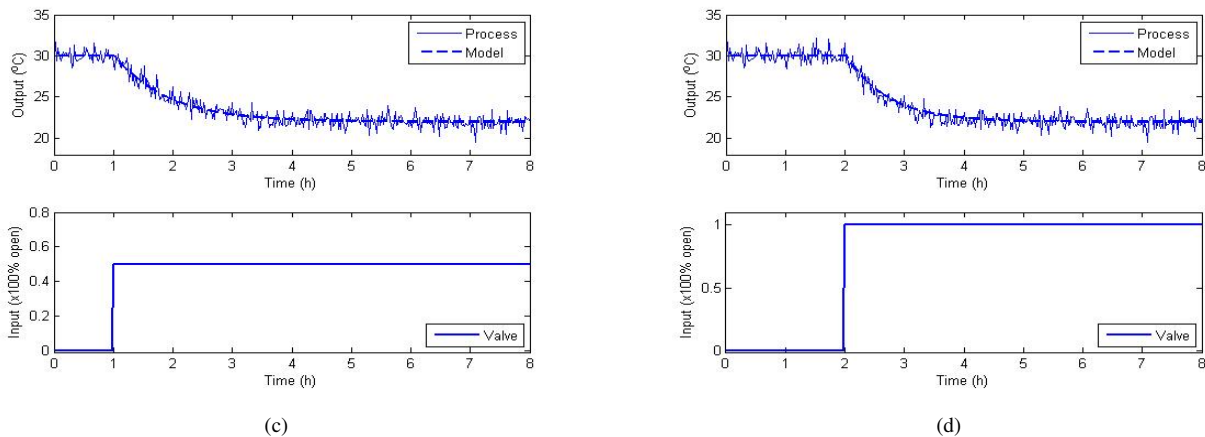


Figure 4: Step test to identify $G_1(s)$

$G_{22}(s) = \frac{46.5e^{-0.17s}}{0.63s+1}$, respectively. For the identification of $G_2(s)$, the static gain is also not constant. However, in this case, the temperature tends to a final value of approximately $39^\circ C$ when the substrate concentration reaches saturation, independently of the input variable $U_2(s)$.

After some step tests, $G_2(s)$ can be modelled as:

$$G_2(s) = \frac{K_2 e^{-0.18s}}{0.62s + 1}, \quad 0 < K_2 \leq 156. \tag{15}$$

In all step tests for the identification of $G_2(s)$, the limit of the level was considered in the reactor, as is the case in practice. The capacity of the reactor is $1000m^3$ which corresponds to a level of $16m$. Thus, if the operator opens arbitrarily the substrate feed valve for a sufficiently long time period, the must in the reactor can overflow.

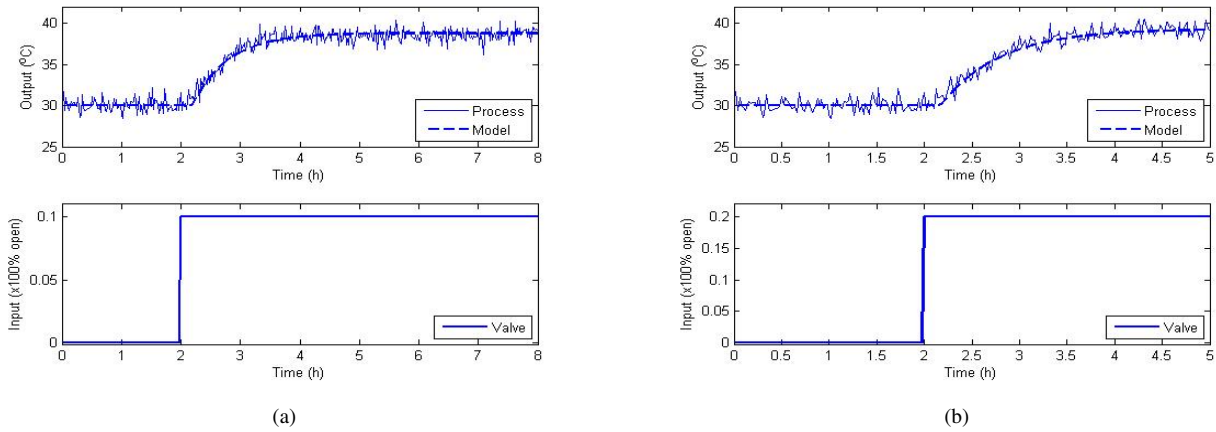


Figure 5: Step test to identify $G_2(s)$

3.1.2 Level Model

The model for the dynamic behavior of the level in the reactor is based on a mass balance of the must in the reactor which is dependent on the flow in:

$$\rho_o u_2(t) \dot{F}_{max} = \frac{dm}{dt} \quad (16)$$

$$m(t) = A_r \rho_i h(t) \quad (17)$$

where \dot{F}_{max} is the maximum substrate flow rate, m is the mass inside the reactor, A_r is the reactor cross-sectional area and h is the level. The substrate and must densities are ρ_o and ρ_i , respectively.

The dynamics of the valve and the sensor can be neglected because they are much slower than the dynamics of the level. Thus, assuming $\rho_o = \rho_i$ and substituting Eq. (17) in Eq. (16), it follows that:

$$u_2(t) \dot{F}_{max} = A_r \frac{dh}{dt} \quad (18)$$

Considering the initial level $h(0) = 4.8m$ (initial volume $v(0) = 300m^3$) and applying the Laplace transform in Eq. (18), the following transfer function for the level is obtained:

$$G_3(s) = \frac{H(s)}{U_2(s)} = \frac{\dot{F}_{max}}{A_r s} \quad (19)$$

3.2 PIDs Tuning

The block diagrams with the terminology used in this paper, shown in Fig. 6, are considered for the PIDs tuning. The computational elements are the reference filter, $F_r(s)$, and the controllers, $C_1(s)$, $C_2(s)$ and $C_{ff}(s)$. The process output variables selected as the controlled variables are $H(s)$ and $Y(s)$, and the process input variables selected for adjustment by the control system are the manipulated variables, $u_1(t)$ and $u_2(t)$.

3.2.1 Level Control

The implemented structure (see Fig. 6(a)) is known as 2DOF (Two Degrees of Freedom) and its transfer function in relation to the reference and the controlled variable is:

$$\frac{H(s)}{R(s)} = \frac{F_r(s)C_2(s)G_3(s)}{1 + C_2(s)G_3(s)} = \frac{(T_d T_i s^2 + T_i s + 1)(\gamma s + 1)}{\left(\frac{A_r T_i \alpha}{K_c \dot{F}_{max}} s^3 + \frac{A_r T_i + K_c T_d T_i}{K_c \dot{F}_{max}} s^2 + T_i s + 1 \right) (\beta s + 1)} \quad (20)$$

where $F_r(s) = \frac{\gamma s + 1}{\beta s + 1}$ and $C_2(s)$ is given by Eq. (12).

In this case, a fast closed-loop response is desired, with a constant time $T_0 = 0.5$. Thus, using of pole placement design, the tuning parameters are:

$$K_c = 0.25, T_i = 1, T_d = 0, \alpha = 0, \gamma = 0.5, \beta = 1$$

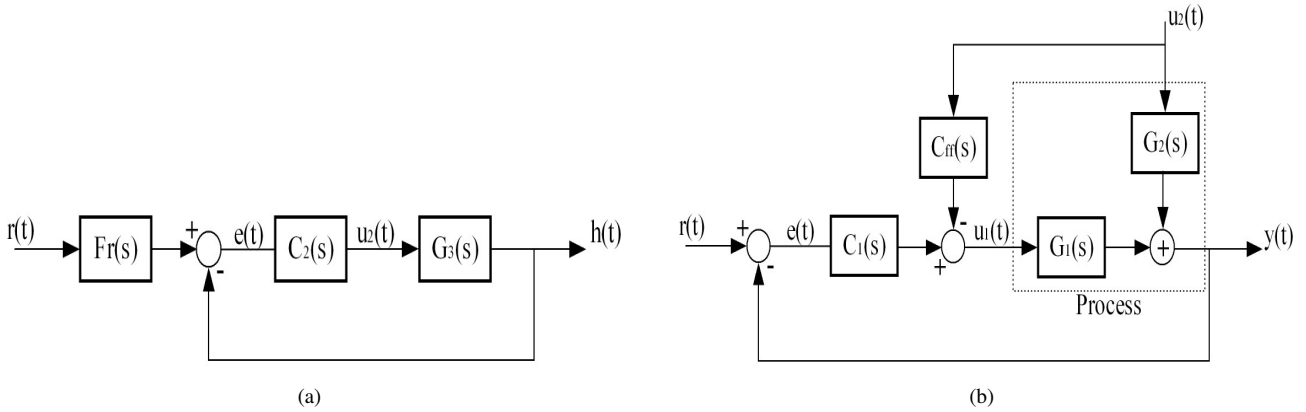


Figure 6: Block diagrams of the feedback control systems

Note that the controller is a PI and a low filter in cascade is not necessary. Therefore, the closed-loop transfer function is:

$$\frac{H(s)}{R(s)} = \frac{1}{0.5s + 1} \quad (21)$$

3.2.2 Temperature Control

To control the temperature, the two controllers shown in Fig. 6(b) can be used: the PID controller, $C_1(s)$, and the feedforward controller, $C_{ff}(s)$.

Using the ZN rule (Ziegler and Nichols 1942), the tuning parameters of the PID are:

$$K_c = 0.03K_1 = -0.9, T_i = 0.04, T_d = 0.01, \alpha = 0.1$$

As $-40 \leq K_1 < 0$, an estimated value for K_c is based on the transient response of the system. Therefore, a fine adjustment of this parameter is required to improve the behavior of the system. This can be carried out in several ways, but generally it is performed by trial-and-error, which requires experience by the operator.

The feedforward controller can be given by Coleman Brosilow (2002):

$$C_{ff}(s) = \frac{G_2(s)}{G_1(s)} e^{-(L_2-L_1)s} = K_{ff} \frac{0.8s + 1}{0.62s + 1} e^{-0.16s} \quad (22)$$

where $K_{ff} = \frac{K_2}{K_1} = -3.9$ is the gain of the controller.

When the process is well approximated as a minimum phase system it may not pay to implement a feedforward controller. Should the effect of the disturbance on the controlled variable be very fast, feedforward could not affect the output variable in time to prevent a significant deviation from the setpoint (Marlin 2000). In this case, it is usually possible to design a PID controller to suppress disturbances well enough so that there is little to be gained by the addition of feedforward control (Morari. and Zafiriou 1989). Thus, the authors prefer not to implement the feedforward controller.

4 SIMULATIONS AND RESULTS

To analyze the system, a simulator was developed in the Matlab[®] Mathworks environment where the PID controllers were implemented with their anti-windup structure. Some criteria were considered: a maximum fermentation process time of 8h and a maximum level of 14.4m (900m³). The initial conditions were: $V_0 = 300m^3$, $X_0 = 31kg/m^3$, $P_0 = 33kg/m^3$, $T_{r0} = 30^\circ C$ and S_0 (kg/m³), X_{d0} (kg/m³) are zero.

The performance of the system is shown in Fig. 3.2.2 where two situations were simulated. In the first, the system was considered without control in open-loop and, in the second, it was considered as subject to the control structure.

For the open-loop system simulation (see Fig. 7(a), 7(c) and 7(e)), the substrate feed valve is operated manually according to the data obtained from a real fermentation process, while the water flow valve remains closed. Note that the temperature becomes high enough to allow a considerable cellular death. Consequently, the fermentation has a low productivity, since the ethanol concentration is around 40.5kg/m³.

For the closed-loop system simulation (see Fig. 7(b), 7(d) and 7(f)), the substrate feed and water flow valves are operated automatically by the controllers. Two setpoint changes were selected for the level and temperature references. The controlled variables stabilize at desired points of reference, guaranteeing an ethanol concentration of approximately 72.5kg/m³. This allows the operator to decide the reactor level and temperature of the fermentation process which will give the best results.

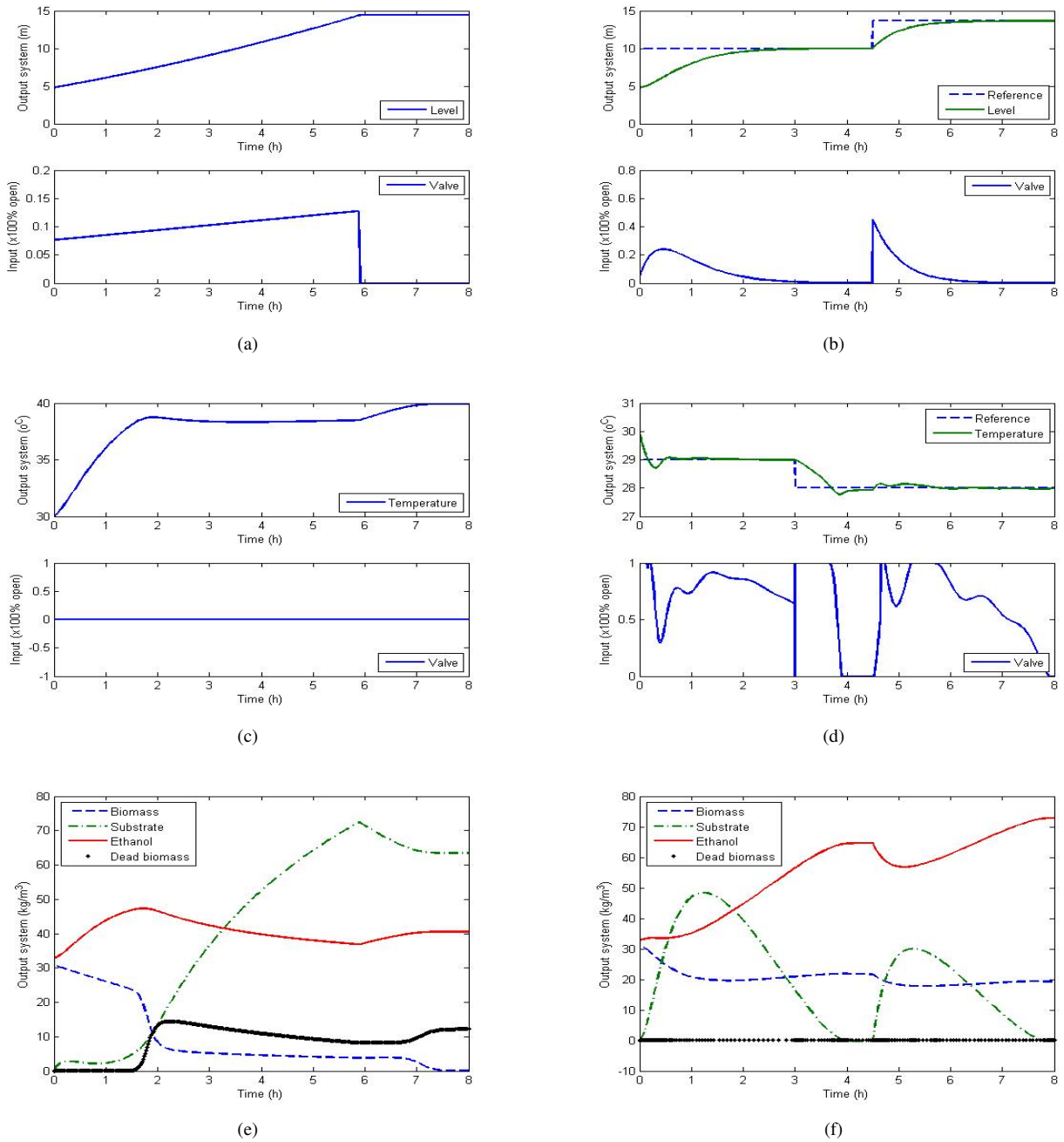


Figure 7: The behavior of the system in open-loop and closed-loop

5 CONCLUSIONS

In this paper, models of the fermentation process and heat exchanger in industrial scale were coupled generating a new system to be studied. PIDs, which have great potential for industrial application, were used to control the system.

The procedure for identification and design that should be used in industry is also presented. The proposed control structure implemented in this study is able to keep the controlled process variables at their setpoints regardless of load disturbances and/or setpoint changes. The simulations show that the ethanol production is dependent on the operation points defined by the operator.

Finally, from a control point of view this new detailed model of a fermentation process with temperature and feed control can be regarded as a very good approximation of a real system and the simulations can be used to analyze industrial situations. With the proposed structure, the engineer can determine and select the best feed control strategy and analyze the performance of the process using different tuning parameters to control the temperature.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Usina Cerradinho for making available the process data and the FINEP and PROABS project for the financial support of this study.

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