Iterative predictive non-linear control in an evaporator to obtain bioethanol.

Ipanaqué William[1], De Keyser Robain [2], Dutta Abhishek [3], Oliden José[4], Manrique José[5].

[1]Universidad de Piura, Dpto. de Electrónica y Automática, Urb. San Eduardo, Piura (e-mail: William.ipanaque@udep.pe)
[2]Department of Electrical Energy, Systems and Automation, Ghent University, 9000 Gent, Belgium (e-mail: Robain.DeKeyser@UGent.be)
[2]Department of Electrical Energy, Systems and Automation, Ghent University, 9000 Gent, Belgium (e-mail: dutta.abhishek@ugent.be)
[4]Universidad de Piura, Dpto. de Electrónica y Automática, Urb. San Eduardo, Piura (e-mail: josecarlosoliden@gmail.com)
[5]Universidad de Piura, Dpto. de Electrónica y Automática, Urb. San Eduardo, Piura (e-mail: jose.manrique@udep.pe)

Abstract: The production of biofuels has a positive impact on the environment, so the modelling and control of these processes warrant the attention of the scientific community. This article presents the model, simulation and non-linear control applied to a double effect evaporation process that is used to extract bioethanol from sugar cane juice. Its purpose is to control the juice concentration at the discharge of the last stage. A NEPSAC non-linear predictive regulator is used as controller because of its important features that facilitate the development of its algorithm and cut down on excessive computation times. The paper presents the model results and the comparison with the plant data as well as the behaviour of the non-linear control based on the developed model.

Keywords: evaporation, bioethanol, sugar cane juice, modelling, non-linear predictive control, NEPSAC, EPSAC

1. INTRODUCTION

The various processes to obtain biofuels have garnered considerable attention in scientific journals because of their positive impact on the environment and the economy (Gupta et al, 2010). The production of bioethanol from biomass is one of the alternatives to reduce the consumption of traditional fuels (Mustafa and Ballat, 2009); the use of sugar cane as biomass offers significant advantages for the production of ethanol (Laser et al, 2002, Limtong et al, 2007) and applying advanced control techniques therefore becomes particularly relevant.

One of the intervening processes to obtain ethanol is the multistage evaporator. A study of its control based on an adequate model is presented in this article.

Several papers, such as Gomolka, 1985; Rousset, *et al*., 1989 a,b, have focused on the simulation of the evaporator model and its control. Villar et al, 1993, presented an interesting paper showing a real facility and comparing their results. Juice and vapour chambers were modelled considering the heat exchange with the outside and the existing delays (Tonelli et al. 1990). These papers establish a relationship between the physical-thermal properties, and some of its specific parameters have been assumed as constant, including the global heat transfer coefficient and the latent vaporisation heat.

The advanced model-based control methodologies include the predictive control based on linear or non-linear models (Quin and Badgwell 1998; De Keyser 2003). While methodologies based on linear models can be considered to be ripe, research and new proposals are still underway on the algorithms

based on non-linear models (Patic *et al.*, 2010; Genceli and Nikolaou 1995; Camacho and Bordons 2007).

The model used in this article corresponds to an evaporator whose simulation data have been compared with those of a plant in northern Peru. This non-linear model is used in a non-linear predictive control strategy known as NEPSAC (De Keyser and Donald 2007; De Keyser and Lazar 2004). The main advantage of this algorithm is that it is based on breaking down the output into two parts: one is identified as 'basic' and the other one is based on the effect of current and future operable variables, and it can therefore deal with the non-linear problem without huge processing demands. The results that are applied to the evaporator model for ethanol production are presented. The process to obtain the controller parameters is highly intuitive and empirical. The results are compared with a linear predictive controller, GPC (Clarke et al. 1987) and a PI regulator.

2. MODEL OF THE TWO-STAGE EVAPORATOR

Two stages to obtain the ethanol have been considered in the evaporator modelling. Each stage is divided into a heating or condensing chamber and a concentration or vaporisation chamber (Ipanaqué and Manrique, 2011; Cadet *et al.*, 1999). These chambers transfer heat to each other through a set of pipes used to transfer vapour (Robert-type evaporators). Figure 1 shows the diagram of a stage of a Robert-type evaporator.

Fig. 1. Diagram of a Roberts-type single-stage evaporator

The following assumptions were considered for the modelling of the evaporator:

- No properties of two-phase flow are considered in the heating chamber, so the transfer of energy from the vapour to the juice shall only be calculated with the differential statuses at the input and outlet
- The mass of vapour within the concentration chamber (vapour inside the evaporator) and the mass of vapour within the condensing chamber (within the shell and tube heat exchanger) is constant
- The juice content level in each effect is constant
- The heat losses that are transferred to each effect represent 20% of the total heat value provided by the vapour (estimated average value)
- The generated vapour and the cane juice of each stage are under temperature balance. The rise in the juice boiling point due to the level of concentration is considered
- The heating vapour is considered to be saturated, and it is condensed at the outlet

The output vapour of stage 1 is the input vapour of stage 2, and the outlet juice of stage 1 enters the following stage as input flow. The condensing and concentration chambers are analysed separately. Below are the variables that are used in the system model.

$\dot{m}_{vapor\ ent}$: mass flow of the input vapour [kg/s].

$\dot{m}_{cond}$: mass flow of the condensate [kg/s].

$H_{vapor\ ent}$: enthalpy of the input vapour [J].

H_{cond} : enthalpy of the condensate [J].

$\dot{q}_{trans}$: flow of transferred heat [J/s].

$\dot{m}_{\text{jugo ent}}$: mass flow of the input juice [kg/s].

$\dot{m}_{vapor\ sal}$: mass flow of the outlet juice [kg/s].

$\dot{m}_{jugo\ sal}$: mass flow of the outlet juice [kg/s].

$H_{jugo\ ent}$: enthalpy of the input juice [J].

$H_{jugo\ sal}$: enthalpy of the outlet juice [J].

$H_{vapor\ sal}$: enthalpy of the outlet vapour [J].

$C_{jugo\ ent}$: concentration of the input juice [°Brix].

$C_{jugo\ sal}$: concentration of the outlet juice [°Brix].

p_0 : pressure of the input vapour [Pa].

p_1 : vapour pressure at the first stage [Pa].

p_2 : vapour pressure at the second stage [Pa].

2.1. Heating or condensing chamber:

Mass balance:

$$\frac{dm_{vapor}}{dt} = \dot{m}_{vapor\ ent} - \dot{m}_{cond} \qquad (1)$$

Energy balance:

$$\frac{dm_{vapor}H_{vapor}}{dt} = \dot{m}_{vapor\ ent}H_{vapor\ ent} - \dot{m}_{cond}H_{cond} - \dot{q}_{trans} \tag{2}$$

2.2. Concentration or vaporisation chamber:

Mass balance:

$$\frac{dm_{jugo}}{dt} = \dot{m}_{jugo\ ent} - \dot{m}_{vapor\ sal} - \dot{m}_{jugo\ sal} \tag{3}$$

Energy balance:

$$\frac{dm_{jugo}H_{jugo}}{dt} = \dot{m}_{jugo\ ent}H_{jugo\ ent} - \dot{m}_{jugo\ sal}H_{jugo\ sal} - \dot{m}_{vapor\ sal}H_{vapor\ sal} + \dot{q}_{trans} \tag{4}$$

Concentration balance:

$$\frac{dm_{jugo}C}{dt} = \dot{m}_{jugo\ ent}C_{jugo\ ent} - \dot{m}_{jugo\ sal}C_{jugo\ sal} \tag{5}$$

The model is simplified by assuming that the mass and energy in both the heating chamber and the concentration chamber show no variation over time.

The system is considered to have two manageable variables: vapour mass flow at the input ($\dot{m}_{vapor\ ent}$), which has a bearing on the transferred heat ($\dot{q}_{trans}$)and the juice mass flow at the input ($\dot{m}_{jugo\ ent}$). The enthalpy of the juice depends on its concentration C_{jugo} and temperature (Hugot, 1986), as shown in the following formula (6):

$$H_{jugo} = (4.1868 - \left(0.0297 - 4.6\times10^{-5}q\right)C_{jugo} + 3.75\times10^{-5}C_{jugo}T_{jugo})T_{jugo} \tag{6}$$

Constant *q* is the purity of the cane juice.

In addition, the vapour input pressure and the pressure in the concentration chambers are assumed to remain constant.

3. NEPSAC ALGORITHM

The NEPSAC algorithm is based on the EPSAC control strategy (De Keyser and Lazar, 2004) that is explained below. The following form of a process model is considered:

$$y(t) = x(t) + n(t) \tag{7}$$

Where $y(t)$ is the measured output of the process, $x(t)$ is the real output of the process and $n(t)$ is the disturbance. Control requires a prediction of future outputs, with a prediction horizon of N_2. Established by:

$$y(t+k\,|\,t) = x(t+k\,|\,t) + n(t+k\,|\,t) \tag{8}$$

The future output can be described as the contribution of two parts:

$$y(t+k\,|\,t) = y_{base}(t+k\,|\,t) + y_{óptimo}(t+k\,|\,t) \tag{9}$$

$y_{base}(t+k\,|\,t)$ is the effect of the past inputs and future control sequence $u_{base}(t+k\,|\,t)$ and the disturbances.

$y_{óptimo}(t+k\,|\,t)$ is the effect of control actions $\delta u(t+k\,|\,t) = u(t+k\,|\,t) - u_{base}(t+k\,|\,t)$ in a control horizon N_u.

The optimised output can be expressed as a convolution equation in discrete time for the impulse response. Its matrix formula is as follows:

$$Y_{óptimo} = GU \tag{10}$$

Where:

$$Y_{óptimo} = [y_{óptimo}(t+N_1|t)\ldots y_{óptimo}(t+N_2\,|\,t)]^T$$

$$U = [\delta u(t|t)\ldots\delta u(t+N_u-1\,|\,t)]^T$$

$$G = \begin{bmatrix} h_{N_1} & h_{N_1-1} & h_{N_1-2} & \cdots & h_{N_1-N_u+1} \\ h_{N_1+1} & h_{N_1} & h_{N_1-1} & \cdots & h_{N_1-N_u+2} \\ \vdots & \vdots & \vdots & \cdots & \vdots \\ h_{N_2} & h_{N_2+1} & h_{N_2+2} & \cdots & h_{N_2-N_u+1} \end{bmatrix}$$

The matrix equation of the projected output is as follows:

$$Y = \bar{Y} + GU \tag{11}$$

Where:

$$Y = [y(t+N_1|t)\ldots y(t+N_2\,|\,t)]^T$$

$$\bar{Y} = [y_{base}(t+N_1|t)\ldots y_{base}(t+N_2\,|\,t)]^T$$

If the output prediction is available, control signal *U* can be optimised while minimising cost function *J*:

$$J = \sum_{N_1}^{N_2}[r(t+k|t) - y(t+k\,|\,t)]^2 + \lambda\sum_{0}^{N_u-1}[u(t+k|t)]^2 \tag{12}$$

Where λ is the weight of the control variable and $r(t+k|t)$ is the reference. The matrix equation of *U* is obtained by minimising the cost function.

$$U = \left[G^T G - \lambda I \right]^{-1} G^T [R - \bar{Y}] \tag{13}$$

$$R = \begin{bmatrix} r(t+N_1|t) \\ . \\ . \\ . \\ r(t+N_2|t) \end{bmatrix} \tag{14}$$

3.1 NEPSAC non-linear predictive controller

The EPSAC algorithm has been extended for non-linear processes, and the result is the NEPSAC algorithm. The available model is non-linear, so the strategy involves performing an iterative approximation of the model predictions, based on a sequence of future outputs, so that these predictions converge into an optimum solution. Future control actions are therefore expressed as the sum of a base sequence $u_{base}(t+k \mid t)$ and an optimised sequence $\delta u(t+k|t)$.

$$u(t+k|t) = u_{base}(t+k|t) + \delta u(t+k|t) \tag{15}$$

In the linear case, the initial value of $u_{base}(t+k|t)$ is unimportant, unlike in a non-linear case. Since it is based on iterations, the value of u_{base} must be as close as possible to $u(t+k|t)$, so $\delta u(t+k|t)$ is smaller than a tolerance.

In the case of non-linear control, matrix *G* is obtained from an impulse input and a step input. The following order of the elements of matrix *G* is obtained for a N_2 prediction horizon:

$$G = \begin{bmatrix} h_{N_1} & h_{N_1-1} & h_{N_1-2} & \cdots & g_{N_1-N_u+1} \\ h_{N_1+1} & h_{N_1} & h_{N_1-1} & \cdots & g_{N_1-N_u+2} \\ \vdots & \vdots & \vdots & \cdots & \vdots \\ h_{N_2} & h_{N_2+1} & h_{N_2+2} & \cdots & g_{N_2-N_u+1} \end{bmatrix}$$

The NEPSAC algorithm can be structured in the following steps:

1. Measuring process output $y(t \mid t)$
2. Selecting a vector U_{base} .
3. Calculating $\bar{Y}$ in the model with U_{base} .
4. Calculating matrix *G* with an impulse input and a step.
5. Calculating *U* minimising cost function *J*. If $U < \varepsilon$ then $u(t) = u_{base}(t|t) + U(1)$ and return to step 1 for the following sampling delay. Otherwise $U_{base} = U_{base} + U$ and return to step 3.

This structure is followed for every sampling period. The number of iterations depends on the proximity of the value of U_{base} to the optimum U.

4. OPEN LOOP TESTS

Open loop evaporator simulations were implemented in the EcosimPro software. The operating parameters in table 1 were taken into consideration for the simulation:

Table 1. Evaporator operating parameters

Parameters	Rated values
$C_{jugoent}$	13.91 °Brix
$T_{jugoent}$	93.97 °C
$\dot{m}_{jugo\ ent}$	33.02 kg/s
$\dot{m}_{vaporent}$	5.42 kg/s
$C_{jugosal}$	16.08 °Brix
$\dot{m}_{vaporsal}$	2.39 kg/s
$\dot{m}_{jugo\ sal}$	28.56 kg/s
p_0	374.81 kPa
p_1	276.55 kPa
p_2	177.28 kPa

The rated values can vary due to factors such as ambient temperature, fluctuations in the temperature, pressure or juice input concentration. Table 2 provides a comparison of the concentrations in the model and plant, with a favourable result.

Table 2. Comparison of values obtained in the model and the real Agrícola del Chira plant.

	Input ºBrix	Outlet of stage 1	Outlet of stage 2
Real plant value	13.91	14.95	15.44
Model in EcosimPro	13.91	14.84	16.086

Figure 2 shows the response of the concentration in degrees Brix in each evaporator stage for an increase in the stepped input from 4.00 kg/s to 5.42 kg /s of vapour in the first stage ($\dot{m}_{vaporent}$).

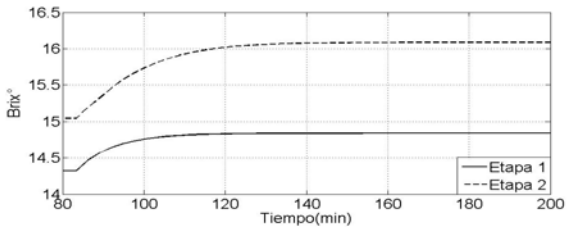


Fig. 2. Response to step input. Degrees Brix in the first and second stage.

The output value is between 15 and 16.5 ºBrix for the second stage. It is important to note that its dynamics are little slower than in the first stage (tests have been run with different vapour inputs). Figure 3 shows the

output of the second stage upon changes in the vapour flow. Its concentration increases as the vapour flow increases.

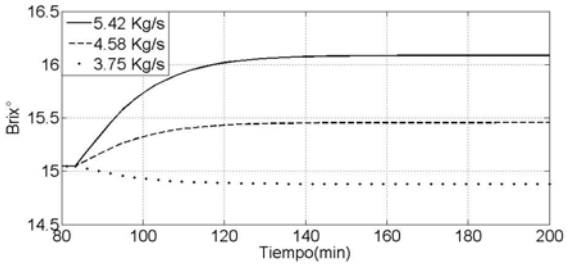


Fig. 3. Output in ºBrix from the second stage for different vapour flow inputs.

5. CLOSED LOOP TESTS

Several tests were performed with the NEPSAC control and with linear predictive controller GPC and PI so that their behaviour could be compared. Regulators NEPSAC and GPC (*Clarke et al. 1987*) employed parameter λ= 0.01 and a prediction horizon of 3 and 4 respectively. The PI controller was tuned by assigning $-6.3*10^{-3}$ and $-7.3*10^{-3}$ poles (Ipanaqué, W., 2012). Figure 4 shows the response of the system under a disturbance of between 13.91 and 13.6 ºBrix in the concentration of supply juice to the first stage in t=6 minutes. In the case of NEPSAC, the output signal returns to the reference value in approximately 50 minutes, which is an acceptable delay for this type of process. In the cases of GPC and PI, the output is further away from the reference when the disturbance occurs and its resetting also takes longer.

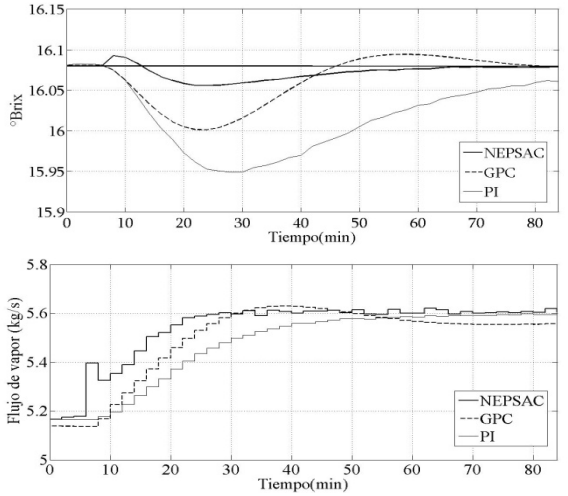
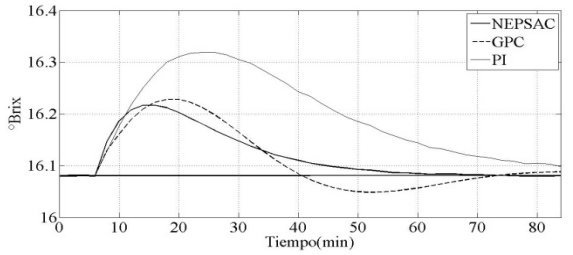


Fig. 4. Response of the closed loop system upon a disturbance in the concentration of supply juice



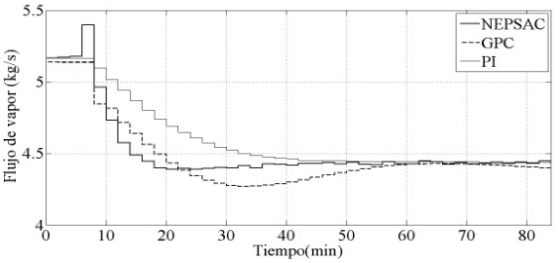


Fig. 5. Response of the closed loop system upon a disturbance in the flow of supply juice.

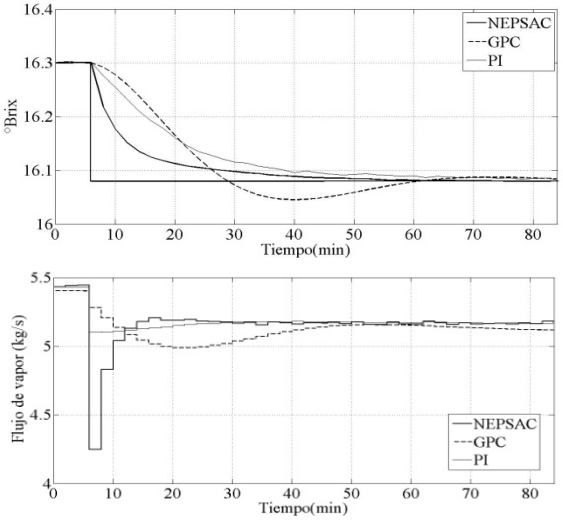


Fig. 6. Response of the closed loop system upon a reference change from 16.3 to 16.08 ºbrix.

Figure 5 includes a test for a disturbance in the flow of supply juice in the first stage. The disturbance was from 1981 kg/min to 1700 kg/min and occurred at instant t=6 min. The NEPSAC controller reaches the reference faster, after approximately 50 minutes, and does not move as far away from the setpoint as in the case of the GPC and PI controller.

A test involving a change of setpoint was done with the three controllers. Figure 6 shows that the NEPSAC controller reaches the reference in 50 minutes, whereas the GPC and PI controllers reach it in approximately 70 minutes. The behaviour of the NEPSAC controller for this system is better upon reference changes because it reaches the setpoint earlier and without oscillations.

6. CONCLUSIONS

A non-linear predictive control application with intuitive parameters has been presented. The controller performance has been tested against variations due to a disturbance and due to variations in the reference. The NEPSAC control strategy yields an optimum performance for values close to the rated values.

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