ROBUST DEAD-TIME COMPENSATION OF A EVAPORATION PROCESS IN SUGAR PRODUCTION

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Abstract: This paper shows the application of a 2DOF robust discrete dead-time compensator to a evaporation process in a sugar factory. The proposed controller is tuned using performance and robustness specifications and it is specially designed for integrative plants as is the case of the level control of the principal unit of the evaporator. A comparative analysis with the PID installed in the unit is performed. The comparative analysis shows that the proposed controller gives better performance than the PID maintaining the simplicity of the tuning procedure. Some simulation results illustrate the good performance of the controller.

Keywords: Dead-time compensators, industrial simulators, sugar production, robustness.

1. INTRODUCTION

In industry there are a lot of processes that exhibit dead-times between the manipulated and measured variables and many of them also present an integral dynamics. In general, tuning a conventional PID controller for these processes is a difficult task. Because of this several studies have been proposed in literature to cope with integrative plus dead-time processes (Palmor, 1996; Normey-Rico and Camacho, 1999*b*; Normey-Rico and Camacho, 2002).

An example of these systems is the evaporation section in a sugar factory. In order to economically produce sugar, energy consumption must be minimized. This requires matching the evaporation and heating requirements in the process to re-use the energy as many times as possible, so a good control and design of this stage is essential to have a good heat economy. Note that a great amount of steam is needed in the evaporation and that this steam is produced in steam boilers which need gas or fuel to work. Evaporation can be carried out in one or several evaporators. When working with high flows and the cost of the steam is high, a chain of evaporators is usually used. In this configuration the product to be concentrated passes from one evaporator to another in series. The steam produced in the evaporation of one of them is used for heating the next evaporator in such a way that only the first effect receive

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steam directly from the boilers. This is known as a multiple effect system. In this configuration, pressure and temperature decrease along the section in every effect. The interest in this multiple effect system lies basically in the more efficient use of the steam (der Poel *et al.*, 1998).

Level in the evaporators and output product concentration are the main variables to be controlled. There are many posible control schematics but all of them suffer from problems related to the long delay times between flow in one end, which is an important manipulated variable, and level or concentration in the other, causing poor performance of the installed PID controller. This poor performance of the PID controller is due to the significant amount of detuning required to maintain closed-loop stability for long dead-time processes (Hagglung, 1996). Dead time compensators DTC seem to be an appropriate solution for these cases because they improve the system response without increasing to much the complexity (Watanabe and Ito, 1981; Matausek and Micic, 1999; Normey-Rico and Camacho, 1999*a*).

In industrial applications simple tuning procedures must be defined, the final number of tuning parameters must be reduced and the physical meaning of them must be maintained. To tune a DTC with these characteristics a simple model of the process must be used. In many cases, as in our case study, it is possible to describe the dynamic behaviour of a slow process using an integrator, a velocity gain and a dead-time. In this cases, as will be shown, the tuning of the controller can be done using simple and intuitive rules. So, this paper proposes a discrete robust DTC based on the results presented in (Normey-Rico and Camacho, 1999*a*) to control the buffer tank level of the evaporation section.

The paper is organized as follows. In section 2 a description of the plant and the modelization of the evaporation process are presented. Section 3 presents the two degree of freedom discrete DTC structure as well as the tuning procedure. A comparative analysis between the proposed controller and a PID is given in section 4, which includes some simulation results performed in a complete non linear industrial simulator. The paper ends with the conclusions.

2. PLANT DESCRIPTION AND PROCESS MODEL

2.1 The evaporation unit

As has been explained, evaporation is the stage in which the water contained in a juice with low sugar concentration is evaporated in order to obtain a juice with a higher sugar concentration. In the unit analyzed in this work the evaporation is carried out using eight evaporators in five effects where the product to be concentrated passes from one evaporator to another in series. In this multiple effect system only the first two evaporators receive steam directly from the boilers as shown in figure 1.

In this configuration of five effects, the first, second and third effects have two evaporators and the fourth and fifth effects have only one evaporator each. An important control problem in this configuration is to stabilise the level of the buffer tank in the inlet of juice. It must be noted that in each evaporator the juice input is used to control the evaporator level. As the output flow of the buffer tank is the input flow of the first evaporator, this flow cannot be used to control the level of the buffer tank, that is, the two controllers cannot use the same flow to control two different variables. This problem could be solved with a simple strategy, just using the juice output flow of the evaporators as the manipulated variables of the local controllers. However, for security reasons, most of the sugar plants use the controllers acting at the evaporator inlets. Note that with the outputs as manipulated variables it is possible for the evaporators to close the valves when the level goes down and in this situation, without juice circulation, the evaporator could produce a crust of caramel. Then, to achieve control of the level of the buffer tank, the extraction flow is controlled in the IV effect. This present a large delay because there are seven evaporators between the actuating point and the controlled level. Because of this, the traditional PID controller installed in the real process achieves a very slow closed loop response. Also, the most common dead-time compensator control structure used in industry, the Smith predictor plus a PI controller, cannot be used in this type of plant because does not reject step disturbances when controlling integrative processes (Palmor, 1996).

To simulate the evaporation section a library has been developed. The basic elements in the libraries are the components. A component represents a model by means of variables, topology, equations and event based behavior. Thus, models for evaporators, valves, pumps, heaters, principal pipes, etc. have been constructed. They have been modeled using first principles equations. This components include all the dynamics and non linearities of the real system. Once the individual components have been developed, the complete model has been builded by connecting the components obtaining a complex mathematical model with 1938 variables and 183 state variables.

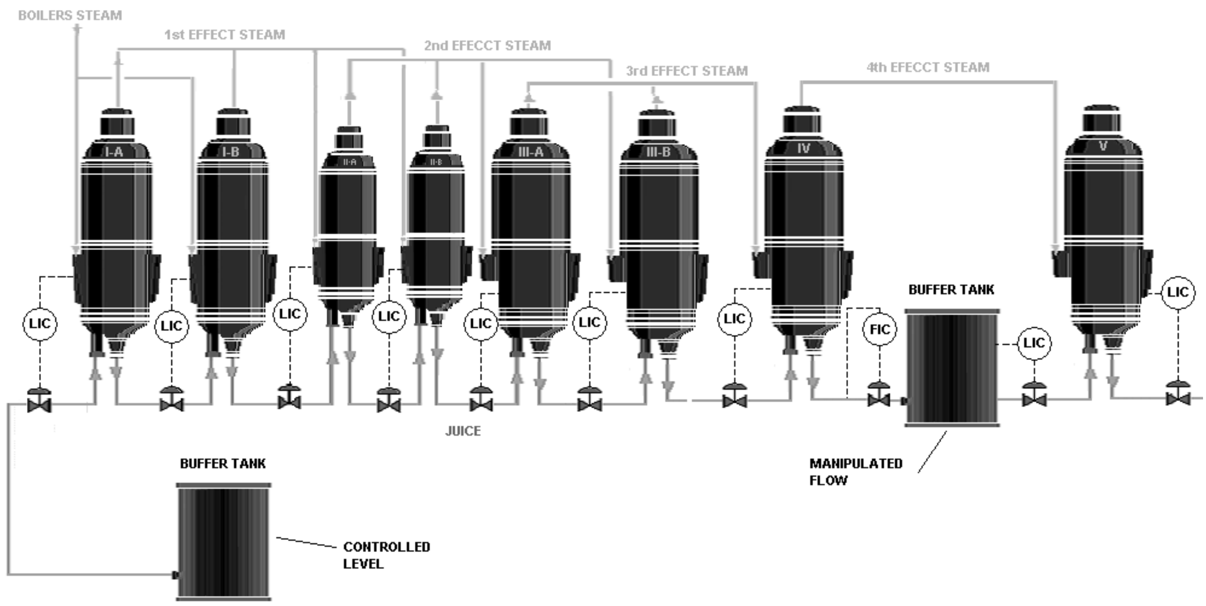


Fig. 1. Structure of the evaporator unit.

2.2 The industrial process simulator

The simulator was developed at the Universidad de Valladolid, Spain (de Prada *et al.*, 2002). For practical reasons, the real industrial process has been replaced by a dynamic simulator fully tested, that is been used for operator training in the factory. The training simulator has the double objective of extending the process knowledge of the plant operator and testing possible actions that he never would test in real working conditions. Due to this objective, the simulator is articulated in two big systems: a simulation program and a distributed control system, where one of the control units works as an instructor console. The objective of the simulation program is to reproduce in the most reliable way the global dynamic performance of the process of sugar production. The behaviour of the simulator has been compared to real data of Spanish beet-sugar factories and tested with sugar experts. The process consists of four main stages: diffusion, purification, evaporation and crystallisation. The simulation is made using the EcosimPro simulation language and the model is developed using libraries of elemental units developed with the object oriented modelling orientation (Zamarreno *et al.*, 2000). Additionally, the simulation code must work in real time and must be communicated with the distributed control system. This is done using the $C++$ simulation class generated by EcosimPro to build a real time data server that communicate with other application, mainly the distributed control system, using an OPC (Zamarreno *et al.*, 2000) (OLE for Process Control) interface. In this case study, the control algorithm has been implemented in $C++$.

2.3 Identification tests.

The complete model of the evaporator unit is a complex non linear one. Note that the changes in the extraction flow of juice at the end of the evaporation section cause a variation in the tank level after modify all the intermediate evaporators as each one has a local control. Also, as the tank has a integral dynamics, it is possible to describe the dynamic behaviour between the level and the set point of a extraction flow controller as a slow process with integral action. If a simple model must be used, a good approximation could be obtained using a transfer function with an integrator, a velocity gain and a dead-time. This allows, as will be seen, a simple tuning of the controller.

To obtain the equivalent dead-time and the velocity gain an identification step test has been performed near the operation point of the process. A step change in the flow juice from 41.8% to 45% has been applied at $t = 0$ seconds. The evolution of the level of the storage tank is shown in figure 2.a. From the figure the dead-time is computed as $L = 110$ seconds and the velocity gain as

$K_v = -0.002\ 1/s$. These two parameters are used in the next section to tune the controller.

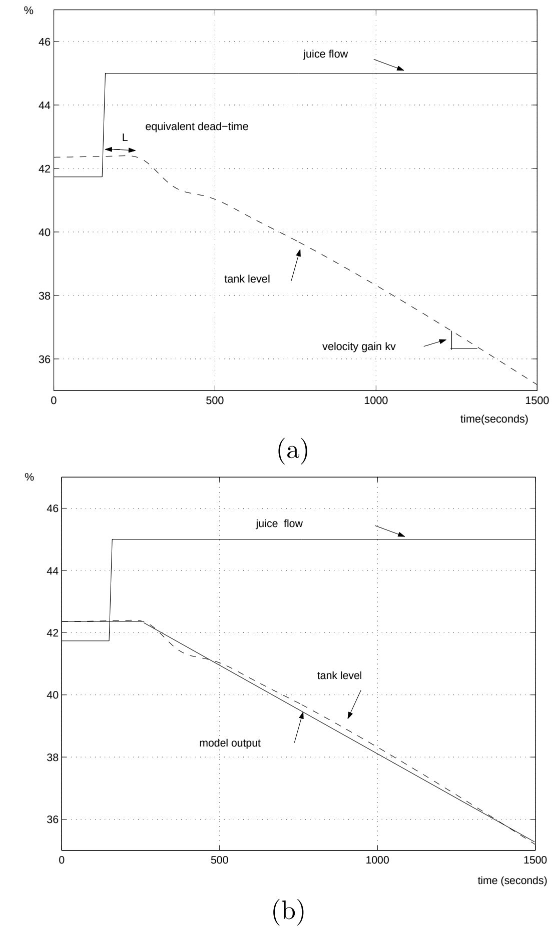


Fig. 2. (a) Identification step test with details of the computation of the estimated gain and dead-time. (b) Validation of the model

For the robust tuning of the controller is interesting to measure the time constant of the non-integrative part of the process (T_e). From figure 2.a it is possible to estimate the 95% settling time of the step response in 1200 seconds. Thus the time constant T_e can be estimated as $T_e = 400$ seconds.

To validate the identified model figure 2.b shows the behaviour of the process and the simple first order nominal model $P_n(s) = \frac{K_v}{s}e^{-Ls}$. As can be seen the nominal model captures the main dynamics of the process.

3. STRUCTURE AND TUNING OF THE PROPOSED DISCRETE 2DOF DTC

For control purposes this paper considers that the process has a transfer function given by:

$$P(s) = \frac{K_v}{s}e^{-Ls} = G(s)e^{-Ls} \qquad (1)$$

where K_v the velocity gain and L is the effective dead-time. As this simple model is obtained using a step test close to the operating point of the process the dead-time L considers not only the real dead-time but also the high order dynamics of the real process. Using a sampling time T ($T = 10$ seconds), the discrete nominal model of the process can be obtained:

$$P_n(z) = \frac{K_v T z}{z-1}z^{-d} = G_n(z)z^{-d} \qquad (2)$$

where d is the nearest integer to the value L/T. In this case $d = 11$ so there is no error between d and L/T. Note that in practice, this numerical approximation error is always small if compared to the error in the estimation of the dead-time of the process.

The proposed control scheme, shown in figure 3, is a two degree of freedom discrete DTC where $P_n(z)$ is the model of the process, $G_m(z)$ is a dead-time free model, $C(z)$ is a PI controller and $F(z)$ is a reference filter.

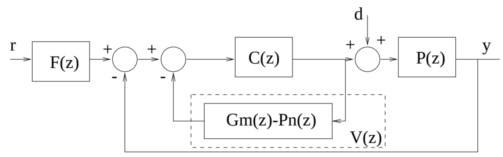


Fig. 3. Structure of the two degree of freedom DTC.

In this scheme, $V(z) = G_m(z) - P_n(z)$ is the predictor model and is given by:

$$V(z) = K_v T\sum_{i=0}^{d-1} z^{-i} - K_v dT.$$

With this choice the system rejects step disturbances. Note that as the static gain of $V(z)$ is zero the integral action of the PI controller imposes the zero static error.

Finally the tuning of the two degree of freedom structure (C and F) is done for a compromise between robustness and performance. First the primary controller is defined as:

$$C(z) = k_c(1 + \frac{Tz}{T_i(z-1)})$$

with $T_i = 2T_0 + L$ and $k_c = \frac{2T_0+L}{K_v(T_0+L)^2}$ (Normey-Rico and Camacho, 1999*a*) where T_0 can be used as a tuning parameter to define the speed of the disturbance rejection response.

Second, to eliminate the overshoot of the set point response the reference filter is tuned in order to minimize the effect of the zero introduced by the primary controller. Also a second parameter T_1 is introduced to define the speed of the set point response:

$$F(z) = K_f\frac{(z-e^{T_0/T})^2}{(z-e^{T_i/T})(z-e^{T_1/T})},$$

where $K_f = \frac{(1-e^{T_i/T})(1-e^{T_1/T})}{(1-e^{T_0/T})^2}$ defines the unitary static gain of the filter.

With this choice the closed-loop relations are given by:

$$\frac{y}{r} = \frac{1 - e^{T_1/T}}{z - e^{T_1/T}} z^{-d}$$

$$\frac{y}{d} = \frac{K_v z}{z-1} z^{-d} \left(1 - \frac{(1 - e^{T_0/T})^2}{1 - e^{T_i/T}} \frac{(z - e^{T_i/T})}{(z - e^{T_0/T})^2} z^{-d}\right) \quad (3)$$

Note that T_0 also defines the robust stability of the closed loop system. Considering that the process can be described as $P(z) = P_n(z)(1 + dP(z))$ the robust stability condition is given by:

$$\mid dP(z) < \mid \frac{1 - e^{T_i/T}}{(1 - e^{T_0/T})^2} \frac{(z - e^{T_0/T})^2}{(z - e^{T_i/T})} \mid \qquad (4)$$

with $z = e^{j\omega}, \omega \in [0, \pi)$. This shows that the tuning must consider a compromise between performance and robustness.

The proposed controller has four tuning parameters (K_v, L, T_1 and T_0), and when a model of the process is obtained experimentally, as is the case of the present study, only two parameters must be tuned: T_1 and T_0.

In practice, to reduce the number of tuning parameters when the dead time is dominant, T_1 can be set equal to the time constant of the non-integrative part of the process. That is, the tuning could be reduced to the only one parameter T_0 that could be tuned manually using a compromise between robustness and disturbance rejection.

Therefore, the proposed controller gives a solution that is as simple to tune as a PID or a Smith predictor and achieves a performance, robustness and disturbance rejection response that cannot be obtained with these algorithms.

4. SIMULATION RESULTS.

As has been mentioned, the real process is simulated with a very complex model that has been validated with real data. Several simulation analysis shows that the real dynamic of the plant is of high order and also presents some oscillations in the open loop response, as can be seen in figure 4 where only the first part of the open loop step response of the system is shown. Thus, to cope with the high value of the modelling error a conservative tuning of the parameter T_0 of the controller must be used. In this particular case a manual tuning has been used to obtain a good compromise between performance and robustness.

4.1 Closed loop results with the DTC.

Two simulations will be shown in order to illustrate the use of the controller. In the first case the simulation test consists of changes in the set-point of the level in the buffer tank, being the manipulated variable the set point of a extraction flow controller (PI).

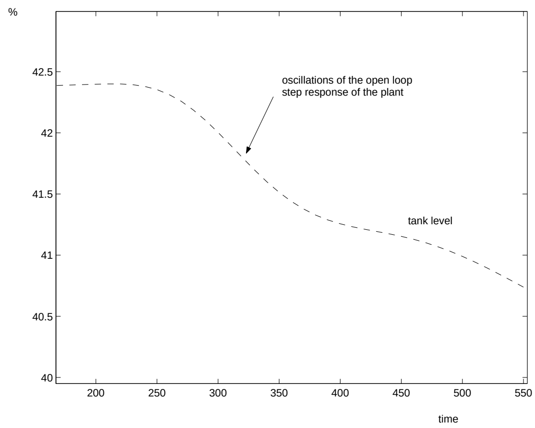


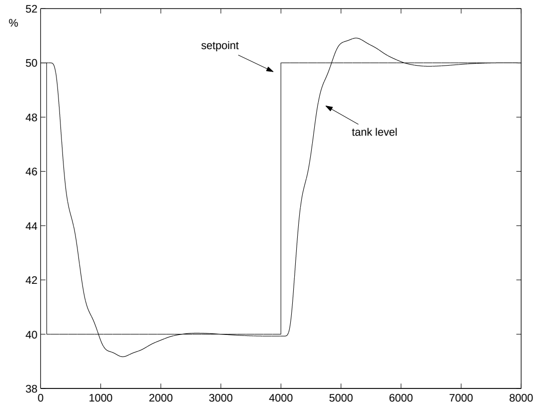
Fig. 4. Details of the open loop step response of the system (time in seconds).

The starting point for tuning is to use a value of T_1 close to the value of the dominant time constant of the non integrative part of the process: $T_1 = 400$. The first value for T_0 is defined as $T_0 = T_1$. This is a good initial choice because in dead-time process the closed loop time constant must be near the open loop one. The obtained time response was too oscillatory and a greater value of T_0 must be used (the time response is not shown). Finally, the response is plotted for $T_0 = 1000$ in figure 5. This high value of T_0 was necessary due to the modeling errors.

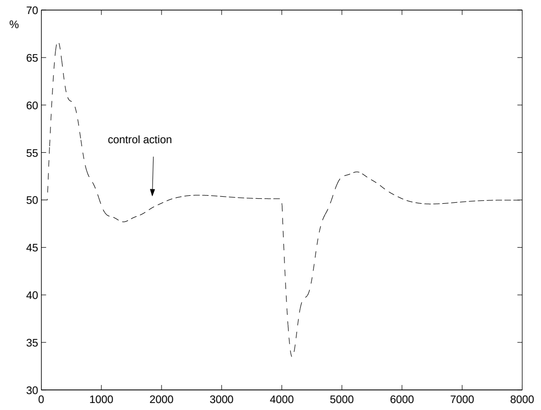
To show the improvement in performance obtained with the controller figure 6 shows a change in the set point using the DTC and the PID controller installed in the process. In the second case the simulation test consists of a disturbance at $t = 0$ produced by a change in the set point of one of the local evaporator controllers. As can be seen with the proposed controller the closed loop system has a small overshoot, less settling time and less oscillatory behaviour. It is important to note that the improvement in performance is limited by the modelling error that is important in this process.

5. CONCLUSIONS

This paper presented the application of a 2DOF robust discrete dead-time compensator to a evaporation process in a sugar factory. The controller performs well both for set point changes and disturbance rejection and the tuning procedure is simple. From the point of view of implementation the controller is as simple as PID controller but allows better performance. Because of the fidelity

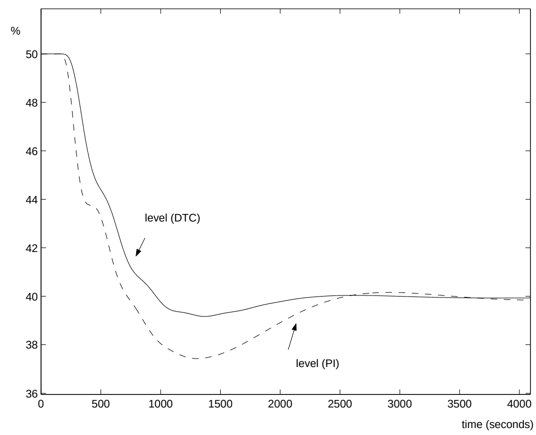


(a)

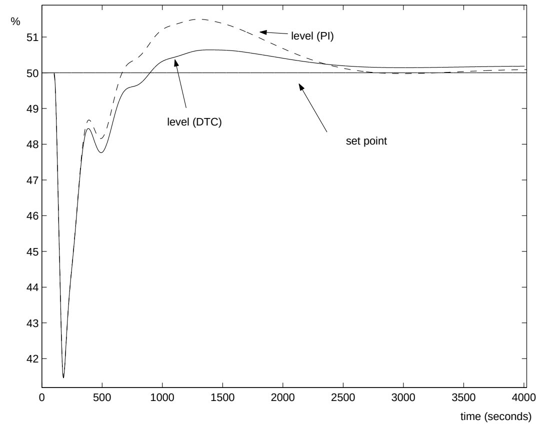


(b)

Fig. 5. Plant output and set point (a) and control action (b) for the proposed DTC controller for two changes in the set point (time in seconds).



(a)



(b)

Fig. 6. Comparative results between the PI controller and the proposed DTC for: (a) a change in the set point, (b) a disturbance rejection

of the process simulator, the controller could be used with success in the real plant.

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