SUGAR FACTORY BENCHMARK

Rogelio Mazaeda [(a)], Alexander Rodríguez [(b)], Alejandro Merino [(c)], César de Prada[(d)], Luis F. Acebes [(e)]

[(a)(b)(c)(d)(e)] Systems and Automation Department, School of Industrial Engineering, University of Valladolid
Center of Sugar Technology (CTA)

[(a)]rogelio@cta.uva.es, [(b)] Alexander.rodriguez@autom.uva.es, [(c)]alejandro@cta.uva.es, [(d)]prada@autom.uva.es
[(e)]felipe@autom.uva.es

ABSTRACT
This paper describes a simulated benchmark specifically designed for trying out and comparing different decentralized control strategies applicable to large scale complex process plants. The benchmark represents a reduced version of a typical beet sugar factory and basically corresponds to the interrelation of the Evaporation and the Sugar End sections. The underlying dynamic model is full of realistic details and has been derived from first principles, stating all the involved mass, energy and population balances. The ready to be used executable is offered to the interested parties as a standard and documented package accessible from any system implementing the widely used OPC process communication protocol.

Keywords: Sugar production, hybrid models, industrial process, hierarchical control, distributed control

1. INTRODUCTION
The typical beet sugar factory consists of several sections in series, which are respectively concerned with the extraction of the sucrose out of the sliced beets by a diffusion in water process, the elimination of as many impurities as possible in a purification plant, the concentration of the resulting still impure sucrose solution in a cascade of evaporators and finally, the Sugar End or House, where the crystallization of the dissolved sucrose in batch and continuous crystallizers is carried out to deliver the white sugar grains with commercial value.

The preceding cursory description is deceptively simple. Each of the above mentioned sections is a full fledged plant on its own right, hosting many process units, some of them of difficult individual operation (Poel, Schiweck and Schwartz, 1998). Additionally, each specific unit participates in a complex layout with numerous mass and energy recycles implying a tightly integrated environment that makes the overall management of the factory an arduous task.

The Evaporation section (fig. 2.a), for example, is made up of several serially connected units in an arrangement that seeks to improve the efficiency of the factory by reusing the steam served by the boiler. The primary objective here is the elimination of the extra water contained in the incoming fresh juice, but the obtained steam, with important energy content, is reused in the same section and in other departments of the plant.

On the other hand, the downstream Sugar End, is a very complex installation organized in several stages (in figure 2.b the main "A" stage is depicted), with many individual batch and continuous pans performing the highly uncertain and poorly measured process of sugar crystallization.

The Sugar House and the Evaporation sections (see figure 1) interacts strongly and not only due to the interchange of the stream of concentrated syrup to process. In addition, the crystallizer pans are heavy consumers of the steam which is served by the evaporators cascade.

The overall management and control of these two specific departments constitutes, then, an important challenge. The difficulty of the task is very much compounded by the need to coordinate continuous and semi-batch type units.

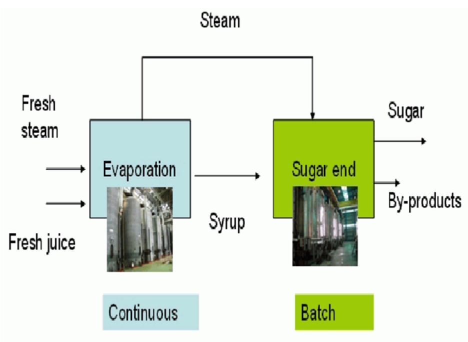

Figure 1: Top level view of the benchmark

The efficient and smooth conduction of the Sugar Factory, and this is true for most process plants, is a complex undertaking. The required solution goes well beyond the regulation of some variables to fixed setpoints at the individual unit level.

Nowadays, in most factories, the needed coordination is heuristically tackled by the operators and technical personnel. Automatic solutions are, of course, possible. In the specific case discussed, where the scheduling of the different batch pans is critical, the control scheme described in Prada, Sarabia, Cristea and Mazaeda (2008) does the job. They use a simplified model of the section in a coordinating central MPC controller implementing a novel control signal continuous re-parametrization of an otherwise discrete or hybrid problem, to find the optimal solution with a reasonable utilization of computer resources.

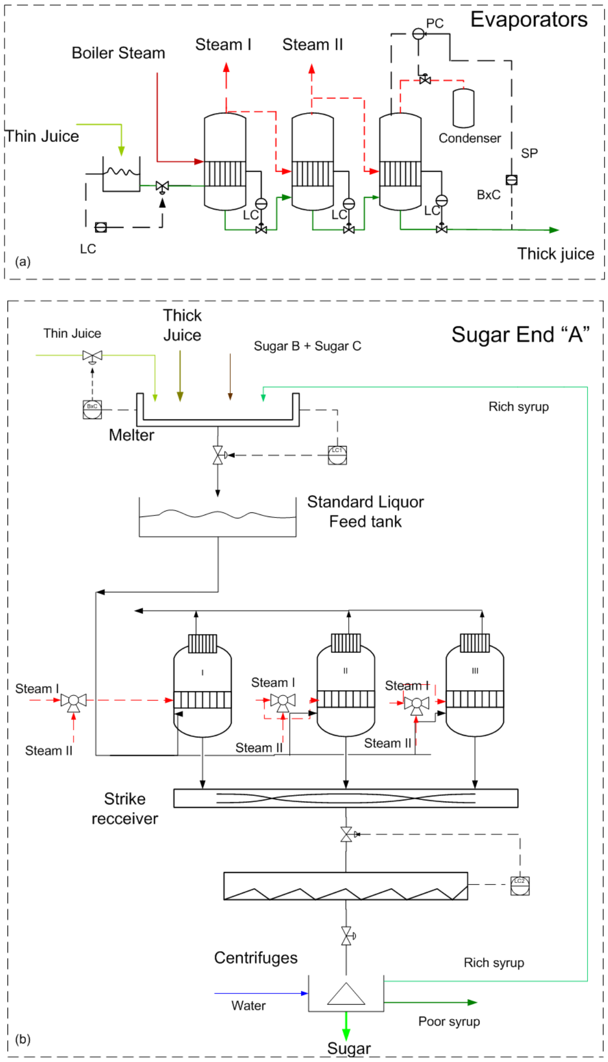

Figure 2: Benchmark detailed view. a) Evaporators cascade. b) Sugar End A stage

In any case, there is an emergent notion in the control community stating that centralized optimizing schemes, in spite of their unquestionable mathematical properties concerning, for example, the rigorous optimality of the solution, should not be the only way to go, especially in the case of large plants. It is argued, for example, that the burden of maintaining a central good enough model is too exacting; and that the associated numerical problem scales badly with the size of the plant.

On the other hand, the structure of the typical process plant, where the main product stream travels serially from installation to installation, with occasional recycle loops, seems to suggest that a decentralized but coordinated or hierarchical control arquitecture would be a reasonable, maybe suboptimal, way to go. The intuition being that the disturbances appearing, for example, at the entrance of a plant of several hours processing time, would not imply too much of a difference for the control actions to be currently adopted at the output. It is then possible for the local control to work, and this would mean important advantages like the possibility of a scalable distribution of the control computer job. Additionally, in the long run, it would result in a conceptually simpler problem. A recent review of the subject is to be found in Scattolini (2009).

In this context, a dynamic realistic first principle model of a reduced scaled down version of the Evaporator and Sugar End is going to be offered as a benchmark for testing and comparing different plant-wide control strategies. The benchmark to be described has officially been adopted by the European Network of Excellence entitled Highly-complex and networked control systems, **HYCON2**, (Framework Program 7 Network of Excellence HYCON2, 2011).

In section 2 a description of the process represented in the benchmark is given, section 3 briefly explains the main assumptions and characteristics of the underlying model and gives references to more detailed descriptions, while section 4, for the sake of conclusions, discusses some of the control and plant-wide coordination challenges posed by the proposed simulated plant.

2. BENCHMARK DESCRIPTION

The benchmark proposed consists of the interrelation between the evaporation section and the first stage of a Sugar House. The scaled down version incorporates detailed first principle models of three evaporators in series and the same number of parallel batch crystallizers, a very simplified, static representation of the A stage centrifuges and along with the necessary auxiliary equipment such as buffer tanks.

The first evaporator receives the stream of technical sucrose solution, the thin juice, with a certain flow rate, purity and concentration to deliver the thick juice at a much greater concentration or Brix. In the real situation the purity and Brix would depend on the beet quality and the workings of the previous sections, but here are considered as given. The concentration of the thick juice is enforced by the PID loop controlling the vacuum pressure in the chamber of the last evaporation unit.

The energy for heating the juice in the first evaporator comes from the live steam delivered by the factory boiler but the second and third units re-uses the vapors obtained from the evaporation of part of the water conforming the juice which is processed in the previous one. This reutilization scheme increases the overall factory's efficiency and is possible since each successive unit is operated at a lower pressure, and so temperature, than the upstream unit.

Steam reuse is not limited to the evaporation cascade. An important fraction of the water evaporated from the syrup is diverted to provide heating energy at other sites in the factory. The subsequent Sugar House department is a particularly heavy steam consumer and the steam demand it represents on the evaporation section is very severe due to the large amounts required and to its intermittent character.

The vapours obtained from the last evaporator are sunk in a barometric condenser. It is common knowledge of the sugar industry that the flowrate of this stream should be minimized since it represents wasted energy.

The difference between the boiler steam pressure at the input and the one that is enforced in the last evaporator chamber by controlling the flowrate to the condenser, drives the downstream vapour flow.

A real factory Sugar House has an architecture consisting in various, usually three, stages: the first or A stage is dedicated to the production of the commercial white sugar crystals and the rest to the exhaustion of the remaining syrup.

The benchmark only represents the first stage. It consists of the parallel array of semi-batch crystallizers followed by a similar disposition of batch filtering centrifuges. The crystallizers receive the so called standard liquor and deliver the massecuite: a viscous slurry consisting of the grown sugar crystal population which is suspended in the resulting syrup or mother liquor.

At the end of their respective strike, each pan discharges the massecuite into a common strike-receiver tank. Next the downstream centrifuges perform the required step of separating the sugar grains form the mother liquor. The purity quality requirements of the white sugar product determine the need of applying water during the centrifuging process for improved filtering. Water helps in expelling the traces of mother liquor from the crystal faces but re-dissolves part of the sugar crystal mass. As a result of this technological setup, each centrifuge offers two type of syrups classified according to their purity: first the so called poor or green syrup, with purity similar to the original mother liquor and then the rich or wash run-off syrup, of higher purity, enriched with the re-dissolved sucrose. The poor syrup gets processed in the following stages and the rich syrup is directly recycled in the A stage.

The standard liquor is conformed in the melter. The main part is the thick juice coming from the evaporators but it also receives the contribution of the above mentioned A rich syrup and of the recovered sugar crystallized in the B and C stages. In the benchmark the melter is simply modelled as a level controlled open tank which assumes the instantaneous and full dissolution of the crystal streams.

The interplay between the workings of the crystallizers and those of the centrifuges is very important for the efficiency of the factory: a bad strike with a not uniform population of crystals would have a poorer filtering capacity in the centrifuges, so it would demand more water with the associated negative impact of the efficiency. To keep the complexity of the benchmark under check, the mentioned compromise is not modelled. In any case, the simplified model of the centrifuge, allows, as a perturbation, the introduction of more water, always a prerogative of the human operator, to simulate the mentioned effect. It is to be noted that the efficiency impact associated with this action, should be understood also in the sense that the obtained syrups are more diluted, so it would imply and extra evaporation effort in the crystallizers and more processing time to each cycle or alternatively a greater demand of steam from the evaporation section.

The differences in rhythm between continuous and batch operated equipment determines the existence of buffer tanks of the appropriate size in the flowsheet of the plant. The standard liquor tank, in particular, which serves the feed syrup to the pans, should accommodate the peaks in demands from the crystallizers with the continuous supply of standard liquor from the melter. In the typical plant, the operator schedules the workings of the parallel array of pans to keep the syrup inventory in the container between safe limits. Observe that the long and uncertain processing times of the batch units make the task difficult. A similar situation exist in the strike receiver, whose level is controlled by modifying the working rhythm of the batch centrifuges, and so their throughput, but the problem here is easier due to the short cycle time and predictability of their time controlled cycles.

2.1. Evaporator

The evaporators considered are of the Robert type. Each unit has two chambers. The heating chamber or calandria encloses a set of vertical tubes that contains the boiling juice. The heating steam enters the shell of the calandria and the energy needed for boiling is transferred to the juice inside the tubes. The heating steam condenses on the shell walls and finally leaves as condensate water. The juice to concentrate enters at the bottom of the chamber, is heated and rises in the interior of the tubes driven by the resulting vigorous bubbling effect produced while boiling. The more concentrated syrup goes over the rim of the tubes and falls into the central downtake and to the output of the unit. The steam produced from the water evaporation emerges from the juice phase, and reaches the upper space containing the vapour. There is a pipe at the top which leads the vapour resulting from evaporation to the calandria of the next station or to the condenser. There is a complex steam delivery circuit that distributes the vapour stream to other units, especially to the batch crystallizers in the Sugar House.

The juice level in the central downtake must be regulated to assure that the generated vapours are safely sealed in the upper part of the unit and to guarantee the height difference with the input of the downstream evaporator which is needed to drive the juice flow.

2.2. Batch crystallizers

The process of crystallization is possible when the concentration of the solute to crystallize, sucrose in this case, exceeds the concentration defining the solubility of the substance at the given temperature. The supersaturation, which is conventionally defined in the sugar industry as the ratio of the two mentioned concentrations, should be greater than unity for crystallization to occur. For moderate values of supersaturation, a metastable zone is defined where the growing of existing crystals is possible but the probability of the creation of new grains out of the solution, is negligible. If the supersaturation, however, is increased so that it trespasses the fuzzy defined labile zone frontier, then the nucleation phenomenon turns explosive, a situation to be avoided in industrial crystallization processes. The solubility of the sucrose increases with temperature and with the presence of impurities. In sugar industry supersaturation can be obtained by cooling or by evaporation of the water in excess. In the present benchmark the batch crystallizers use the latter mechanism.

The batch sugar industrial crystallizer serves the purpose of creating the conditions of supersaturation of the technical solution of sucrose, so that a tiny initial population of sugar crystals may steadily grow until it achieves the commercial average size. It is an important technological requirement that the spread of the distribution of sizes in the population is kept as narrow as possible and this implies that supersaturation should be always kept at moderate values in the so called metastable region. The supersaturation conditions are created by striking the right balance between the rates of water evaporation and of the standard liquor which is supplied to replenish the solution of the sucrose that has migrated to the faces of the crystals.

The evaporation is carried out at low, vacuum pressures so as to keep the temperature of the mass at reduced values so avoiding the quality impairing caramelization of sugar. The process is conducted in a semi-batch fashion, and this means that the prevalent conditions are continuously varying along the strike. The impurities, for example, get accumulated along the cycle, so the mother liquor purity gets progressively lower and this implies a greater difficulty in keeping the right supersaturation. The amount of massecuite in the pan grows from an initial value of roughly half to the full capacity of the pan at the end; and this fact has an important negative impact on the circulation of the mass and on the the heat transfer efficiency of the unit. The difficulties in the conduction of the pan are compounded by the absence of important on line measurements. The supersaturation, for example, that is critical, depends of other variables like the concentration, the temperature and the purity of the solution. Purity is not measured on-line but it is periodically reported by the factory laboratory. The on-line measurement of the concentration of the solution or Brix is problematic and the temperature of the mass is not homogeneous, especially at the end of the strike because of poor circulation, so its determination is also uncertain.

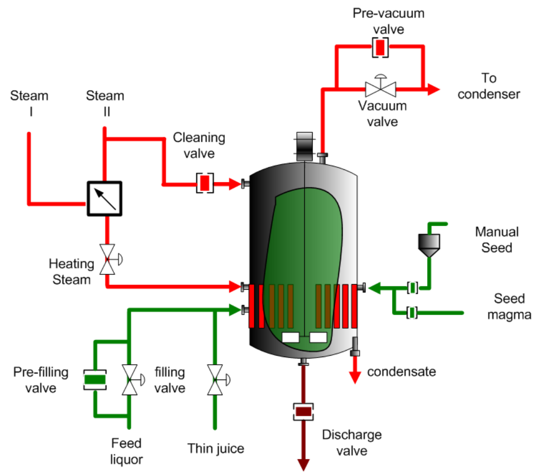

Figure 3: Vacuum pan crystallizer

The vacuum pan is constructed as a cylinder with a diameter and height of comparable dimensions. At the bottom, there is a floating calandria type of heat exchanger. The mass circulates inside the tubes and the central downtake, and the heating steam goes inside the shell.

The calandria is designed so as to bolster the circulation of the mass. The massecuite rises in the tubes as the water component is evaporated. The steam rises through the existing mass and emerges at the surface to enter the vapour occupied phase. The massecuite, which is driven over the tubes rim by the rising bubbles, returns to the bottom via the central downtake. In order to intensify the circulation, a mechanical stirrer is placed at the bottom of the downtake. The stirrer contribution is important mostly at the end of the strike, when the achievable evaporation is lower and the natural agitation provided by the bubbles are probably not enough.

The pan is provided with the necessary valves to regulate the heating steam input to the calandria, the feed syrup to the chamber and the seed magma containing crystal initial population. There are valves that allow the discharge of the mass at the end of the strike, and the control of the evaporated steam out of the camber to the condenser. There is also a cleaning valve for inputting steam after the product evacuation with the purpose of removing the traces of massecuite.

Note that in the specific unit modelled (fig. 3), it is allowed to individually choose the evaporator effect which is going to be the source of heating steam. Normal practice dictates the use of lower pressure steam (II effect) at the beginning of the cycle when the heating process is more efficient and then, to switch to higher pressure steam (effect I) as the cycle progresses and the mass transfer coefficient rapidly diminishes.

The instrumentation used in the pan allows to measure the following variables: chamber mass temperature and steam space pressure, the level attained by the mass and the steam pressure in the calandria. The unit has a radio frequency (RF) sensor whose on-line readings can be calibrated to somehow represent the concentration of the slurry: solution plus growing crystals. The electrical current which is drawn by the stirrer motor, could be taken as a indication of the consistency of the mass, and this fact is put to use at the end of the strike, when the RF transmitter measurements are less reliable.

2.2.1. Batch crystallizer program

Each batch crystallizer cycle follows a recipe implemented by sequential program (fig. 4.a) whose main stages are the following:

1. **Loading**: A high capacity valve is fully opened to start the introduction of standard liquor in the chamber. The objective is to load enough syrup so as to completely cover the calandria in such a way as to maximize the circulation process of the mass and improve the heat transfer coefficient. The two PID based loops controlling the pressures in the chamber and in the calandria are both put in automatic mode.
2. **Concentration**: The heating of the mass continues to concentrate the mass with the purpose of reaching the required supersaturation. The heat transfer coefficient and the steam consumption are both, at this moment, very high. The stage ends when the syrup Brix reaches a value, which in view of the most recent standard liquor purity report from the laboratory, would correspond to the right supersaturation.
3. **Seeding**: The seeding stage proceeds by automatically introducing the amount of seed magma which is considered adequate for obtaining the final correct average size.
4. **Growing of the grain**: This is the longest and most important stage of the cycle. The population of crystals in the seed should be made to grow as the sucrose in the solution migrates to their faces along the strike. The supersaturation would tend to decrease as the impurities accumulate in the solution, so standard liquor should be introduced in a controlled way to add the dissolved sucrose needed for compensating this effect. The amount of syrup to introduce would depend on the rate of evaporation, and on the purity currently existing in the pan. In the absence of on-line supersaturation estimation, the stage is conducted by establishing a curve of massecuite Brixes that gives, at each instant in the evolution of the process, the total mass concentration that should be enforced to obtain the right supersaturation. It should be noted, that the readings of the RF sensor takes into account not only the dissolved substances but also the mass of growing crystals. The level attained by the mass in the chamber, which should grow from the value initially loaded to the pan full capacity, is taken as a measure of the evolution of the strike. So, the feed syrup input flowrate is controlled in the stage by the scheme shown in fig. 4.b, where the setpoint for mass concentration is given by a Brix vs. level curve that should be adjusted by the operator to reflect the changes in standard liquor purity.
   The considerable reduction of the heat transfer coefficient in the heat exchanger, taking place as the mass level rises, is compensated, in some degree, by modifying the setpoints of the calandria and of chamber steam pressure regulators.
5. **Tightening Up**: The purpose here is to increase the consistency of the massecuite in preparation for the discharge. There is no further introduction of syrup but the evaporation continues until the electric intensity consumed by the stirrer motor attains a configurable value. The setpoint of the calandria pressure controller is raised to accelerate the process.
6. **Discharge**: The heating steam input is shut down and the vacuum is broken in the chamber by opening the cleaning valve. When the pressure reaches an appropriate high value, the discharge output gates are opened.
7. **Cleaning Up**: Cleaning valve is kept opened with discharge gates closed to get rid of the traces of massecuite which remain contaminating the interior walls.

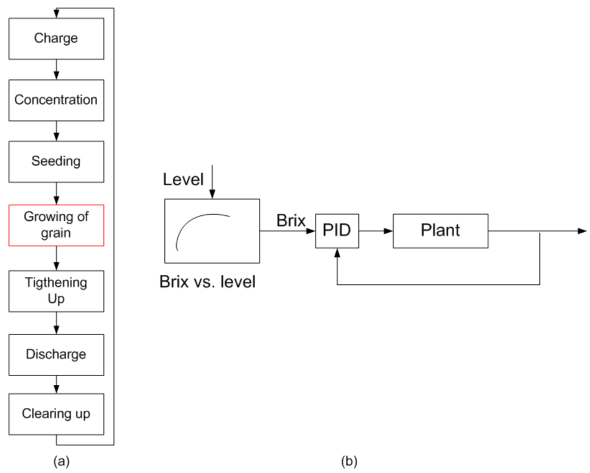

Figure 4: Crystallizer program. a) Main stages. b) Brix controlling loop in growing state

In figures 5 and 6 the evolution along one cycle of the level of the mass and of the mass concentration of crystals are respectively shown, highlighting in each case the instant of activation of some important events.

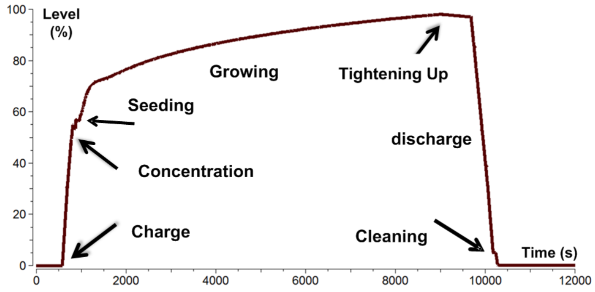

Figure 5: Level evolution

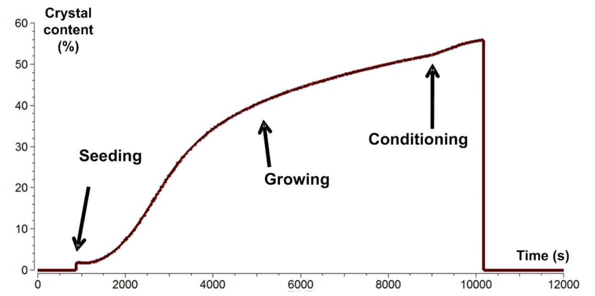

Figure 6: Crystal content evolution

3. BENCHMARK MODEL

The model has been created assembling objects instantiated from a previously existing library of sugar factory components using the Object Oriented (OO) concepts implemented by the **EcosimPro** modelling and simulation tool (ESA International, 2008).

The original purpose of the library was to provide the elements for constructing realistic dynamic simulators specially dedicated to the training of the beet sugar factory control room operators.

The library contains a representation of the main units to be found in the factory. It obviously include classes representing evaporators and batch evaporative crystallizers but it also hosts the auxiliary equipment needed in any process industry such as valves, pumps, tanks and even PID regulators (Merino, Acebes, Mazaeda and Prada, 2009). These ancillary elements are used to create the topology of the specific plant but are also deployed in the definition of the main process units to define its internal structure.

All the models are derived from first principles and are coded in a generic way, exposing numerous parameters, making possible the adaptation of the resulting overall instantiated model to the specific situation at hand.

The models exhibit a hybrid character. They are predominantly made of continuous time dynamical equations but must also respond correctly to discrete events fired, for example, as the program of the crystallizers move form on stage to the next.

The original motivation as a training aid has permeated the modelling effort, guiding the election of the general assumptions of the library. Since all variables and actuators were in principle susceptible of being accessed by the trainee, the models should meticulously state all the involved mass and energy balances. In the case of the crystallizers the evolution of the mass of grains is tracked by means of the formalism of the population balance equations. The need to be able to simulate whole large factories determined the use of globalized models, wherever possible, to facilitate the numerical integration effort. There is an ample use of non dimensional relations used in the general chemical engineering literature and in sugar studies, and mass and energy transfer rates are put in relation with the characteristics of the processed streams. This provides reasonable starting values for the physically related parameters that can be further tuned to adapt to each specific case. The physico-chemical properties of the main products such as syrup or juice and massecuite had been taken from the existing specialized literature (Bubnik, Kadlec, Urban, and Bruhns, 1995). The characteristics of typical utilities such as liquid water and steam are readily available.

The capacity of the generic model to faithfully reproduce real plant data and to meet the informed qualitative demands of sugar experts had been described elsewhere (Mazaeda, 2010; Merino, 2008; Mazaeda, Prada, Merino and Acebes, 2012). The assembled model here proposed does not exactly emulate any existing plant; but the deployed individual units are the ones that have been calibrated and validated with real data.

The Sugar Benchmark has a very different purpose from the one that originally motivated the design of the underlying OO library. In any case, the model proposed, with its abundance of realistic details, full of special situations, which are needed for training, but that that would be considered as non-essential in almost any other type of application, would stand with respect to the designer of the overall control strategies, in a situation approximately similar to the one he/she would encounter when facing a real world problem.

Detailed descriptions of the evaporator and the vacuum pan crystallizer models can be found in Merino, Alves and Acebes, (2005) and Mazaeda and Prada (2007) respectively. The centrifuges model has been specifically created for the benchmark. It is less involved that the previous models and has a static character. It simply consists of the necessary balances to each component, taking into consideration the dissolution provoked by the amount of water introduced. The exact composition of the poor and rich syrup streams and the humidity of the separated sugar grain product are decided by adjustable parameters.

4. THE BENCHMARK CHALLENGE

The benchmark main purpose is to serve as a testing platform to develop high level strategies with the capacity of guaranteeing an optimal economical behaviour simultaneously dealing with the management of the continuous processes and the scheduling of the batch units. The solutions proposed should be able to cope with the perturbations and uncertainties represented by the variability of the working conditions and the lack of a complete knowledge of the state of the process.

More specifically, the problem of conducting the whole plant in a smooth way is complicated due to the uncertainty in the batch crystallizers processing times which are very dependent on the characteristics of standard liquor. As an example, in figure 7, the effect that a modification of the feed purity, keeping the Brix vs. level curve fixed, has on the cycle time and on the evolution of other important variables, is shown.

A smoothly managed plant should schedule the workings of the batch pans in such a way as to guarantee the consumption of all the syrup delivered by the upstream continuous section. Figure 8 shows the evolution of the levels in the feed syrup tank and in the strike receiver in this ideal situation.

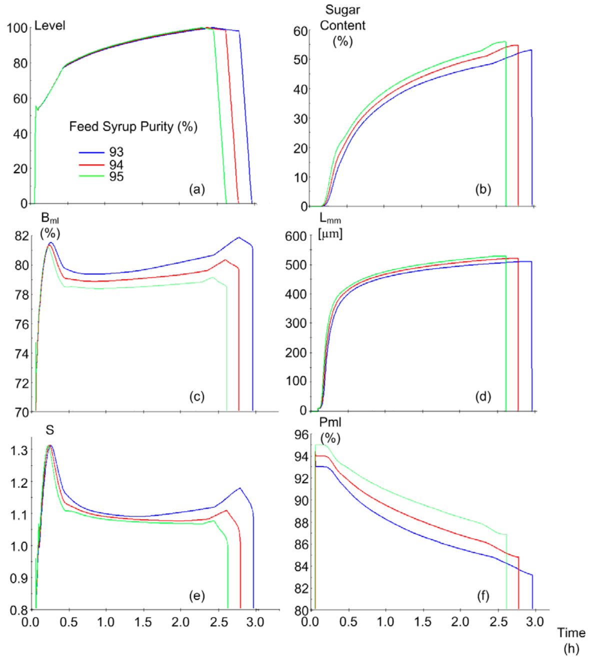

Figure 7: Effect of feed syrup purity on the performance of a crystallizer. a) level. b) Sugar content. c) Brix of mother liquor. d) Average crystal size. e) Supersaturation. f) Purity of mother liquor

In figure 9, a decrease of the flow rate of thick syrup or an increase of purity could lead to the depicted situation, where the level in the feed syrup tank is clearly diminishing with the risk of violating the safety restriction on the required inventories. It should be noted that the availability of steam and the crystallizer cycle time are both very dependent on the Brix enforced at the output of the evaporation section. On the other hand, the amount of water to the centrifuges alters the concentration of feed syrup and also its purity.

More formally, the objectives to be achieved in conducting the plant simulated in the benchmark are the following:

- Minimize the consumption of boil steam while serving the demands of the crystallizers.
- Guarantee the processing of all the incoming syrup. The inventories of buffer units, strike receiver and, fundamentally, the standard liquor tank should be kept between safe limits.
- The quality standards of the produced white sugar should be met.

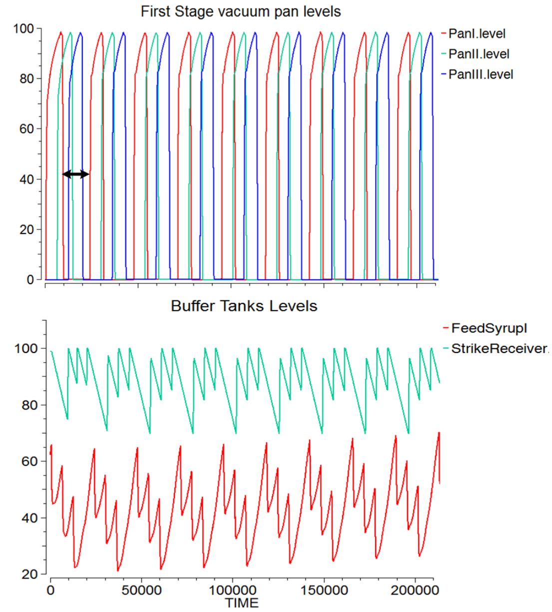

Figure 8: Level in buffer tanks over several cycles

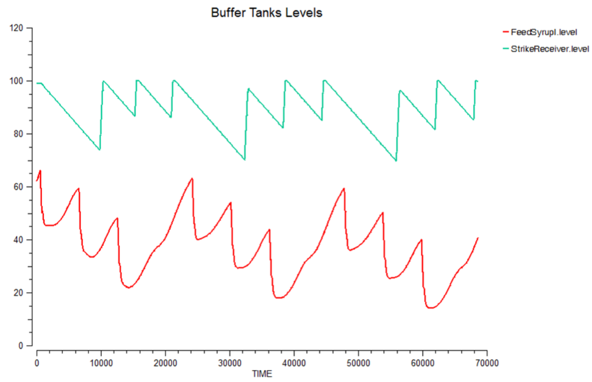

Figure 9: Level in buffer tanks with problems

Table 1: Allowed range for perturbation variables

Parameter	Description	Range
W_{evap}	Mass flow rate into evaporation	10-16 kg/s
B_{mm_evap}	Brix of juice into evaporation	40-45 %
P_{mm_evap}	Purity of juice into evaporation	91-95 %
W2mc	Mass flows rate of water to massecuite in centrifuges	0.02-0.025

Table 2: Degrees of freedom for plant-wide control

Variable	Description
SP_{evap_Brix}	Setpoint of Brix control loop at Evaporation output
P_{boiler}	Pressure form boiler to Evaporation effect I
$Load_{VP}[k]$	k= 1-3. Load command for each vacuum pan
$Valv_{cent}$	Valve controlling centrifuge throughput

The objectives must be attained in the presence of perturbations, whose range of variation is shown in table 1.

The minimum set of variables which are at the disposal of the design solution for the management of the plant are described in table 2.

The variables that can be read form the benchmark are the same ones which are typically sensed in the real factory, namely:

- The levels of all evaporators, vacuum pans and tanks.
- The temperatures of all evaporators, vacuum pans and tanks.
- The Brixes of all syrups involved.
- The evolution of the Brix of the massecuite in each vacuum pan.
- The purity of the syrups involved.
- The values of all the pressures involved: in the chambers of the evaporators and vacuum pans and in the heat exchangers.

It is possible to consider the problem proposed at several levels. In the simpler approach, the crystallizers could be considered to be reasonably well controlled and the plant-wide coordinator should be simply concerned with guaranteeing the availability of steam and syrup as demanded. The third objective of keeping up with the quality requirements of the sugar product would be considered as automatically enforced by the pan's program if the steam demand is served.

But it is also possible a more involved strategy, which deals directly with the control of each crystallizer. This would imply the handling of the values of the Brix vs. level curve, of the setpoints for the pressure of the calandria and of the chamber, among other details. Of course, this other approach would surely be able to achieve a more efficient solution, but would also imply a greater responsibility concerning the quality of the end product.

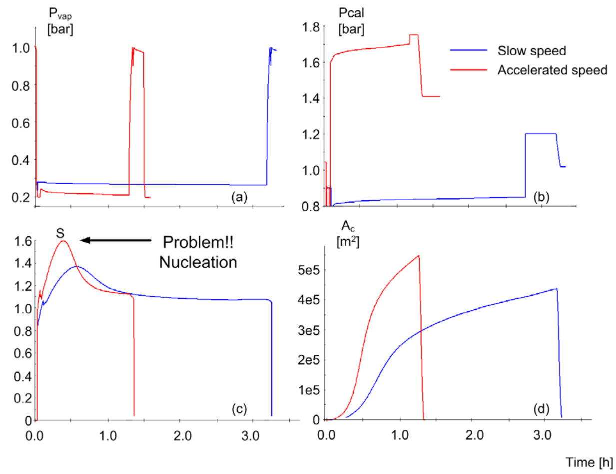

Figure 9: Cycle time acceleration by increasing evaporation rate. a) Vacuum pressure setpoint. b) Calandria pressure setpoint. c) Supersaturation. d) Aggregated area of crystal population

An example of the complex issues involved in applying the second, lower level strategy can be discussed analyzing the figure 9. The setpoints of the controllers of heating pressure to the calandria and of vacuum in the chamber influences the duration of the strike. So it would be legitimate to consider the use of these references as additional degrees of freedom for achieving the plant-wide objectives. It should be bore in mind, however, that an immoderate use of this kind of acceleration is somehow artificial and is limited by the purity of the feed liquor and so should be performed with caution. An unreasonable increase in the calandria pressure would imply an excessive increase of the supersaturation. The reason being, that as the crystallization kinetics is basically unaffected, the existing aggregated crystal area of the sugar population had not reached the value that would be able to sustain a flowrate of crystallization capable of compensating the new rhythm of water evaporation. As a consequence, supersaturation gets inside de labile region with the consequent prejudice to the quality of the product: a wider and reduced average size population. It goes without saying that the new setpoint pressure to fix would also depend of the availability of steam from the evaporation section.

The benchmark gives a suitable platform for testing many other types of interesting applications like, for example, the hybrid identification of the discrete-continuous units, or the explicit handling of the uncertainty by embedding the numerical optimization procedure in a stochastic framework.

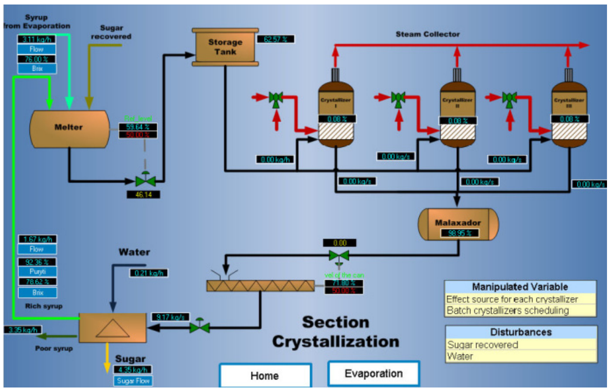

Figure 10: HMI interface to the sugar benchmark

Finally, it should be said that the simulated plant is going to be made available as an executable (Alves, Normey-Rico, Merino, Acebes and Prada , 2005) which implements the **OPC** protocol (Iwanitz and Lange, 2002; Zamarreño, 2010) and with a graphic interface (Alves, Normey-Rico, Merino, Acebes and Prada, 2006) making possible its standalone operation. In figure 10 a screenshot of the user interface is shown. The use of **OPC** will additionally facilitate the access to the simulated data from any of the many clients currently supporting that widely adopted standard.

ACKNOWLEDGMENTS

The research leading to these results has received funding from the European Union Seventh Framework Programme [FP7/2007-2013] under grant agreement n°257462 HYCON2 Network of excellence.

REFERENCES

Alves, R., Normey-Rico ,J., Merino, A., Acebes, L.F., Prada, C.d., 2006. Edusca (educational scada): Features and applications. In: *7th IFAC Symposiumin Advances in Control Education*.

Alves, R.A., Normey-Rico, J., Merino, A., Acebes, L.F., Prada, C.d., 2005. *OPC based distributed real time simulation of complex continuous processes*. Simulation Modelling Practiceand Theory, 13, 525-549.

Bubnik, Z., Kadlec, P., Urban, D., Bruhns, M.,1995. *Sugar Technologist Manual. Chemical and Phys-ical Data Manufacturers and Users*. Dr. Albert Bartens.

ESA International, 2008. *EcosimPro User Manual, EL Modelling Guide*. EA International and ESA.

Framework Program 7 Network of Excellence HYCON2, 2011. *Highly-complex and networked con-trol systems*. www.hycon2.eu.

Iwanitz, F., Lange,J., 2002. *OPC: fundamentals, implementation and application*. Hüthig.

Mazaeda, R., 2010.Librería de modelos del Cuarto de azúcar de la industria azucarera para entrenamiento de operarios. Ph.D. Thesis, University of Valladolid.

Mazaeda, R., Prada, C. d., 2007. Dynamic simulation of a sucrose batch evaporative crystallizer for operators training. In: *19th European Modeling and Simulation Symposium (Simulation in Industry), EMSS 2007*.

Mazaeda, R., Prada, C. d., Merino, A., Acebes, L.F., 2011. *Librería de modelos orientada a objetos para la simulación del Cuarto de azúcar: Cristalizador continuo por evaporación al vacío*. Revista Iberoamericana de Control Automático, 8, 100-111.

Merino, A., 2008. *Librería de modelos del cuarto de remolacha de la industria azucarera para entrenamiento de operarios*. Ph.D.Thesis, University of Valladolid.

Merino, A., Acebes, L.F., Mazaeda, R., Prada, C.d., 2009. *Modelado y simulación del proceso de producción de azúcar*. Revista Iberoamericana de Control Automático, 6, 54-60.

Merino, A., Alves, R., Acebes, L.F., 2005. A training simulator for the evaporation section of a beet sugar production process. In: *European Simulation Multiconference. ESM05*.

Poel, P.V.d., Schiweck, H., Schwartz, T., 1998. *Sugar Technology. Beet and Cane Sugar Manufacture*. Dr. Albert Bartens.

Prada, C. d., Sarabia, D., Cristea, S., Mazaeda, R., 2008. *Plant-wide control of a hybrid process*. International Journal of Adaptive Control and Signal Processing, 22, 124-141.

Scattolini, R., 2009. *Architectures of distributed and hierarchical Model Predictive Control-a review. Journal of process control*, 19, 723-731.

Zamarreño, J., 2010. *Acceso a datos mediante OPC*. Editorial Andavira.