Plantwide control of fruit concentrate production

Mark van Dijk, Sander Dubbelman, Peter Bongers *

**Unilever Research Vlaardingen Netherlands, Process Systems Engineering.*
(E-mail: mark-van.dijk@unilever.com).

Abstract: Fruit concentrates are key ingredients in many fruit based Unilever products. We have designed a novel continuous fruit concentration process involving a decanter, an evaporator and a recombination process. In order to ensure best product quality and highest capacity a methodology for control structure design for complete processing plants (plantwide control) was applied. This included defining the control objectives, degrees of freedom analysis, definition of inventory and production rate control and the development of a non-linear dynamic model. Using the methodology, control alternatives were systematically analyzed and eliminated. The chosen control structure was successfully applied and implemented in a Unilever factory.

Keywords: Control structure design; Plantwide control; Dynamic control model; Decanter; Evaporator.

1. INTRODUCTION

Plantwide control is viewed as a strong methodology to design effective control structures in complex food plants.[1]

This paper presents the conceptual process control design of a novel fruit concentration method using a separation and recombination step. It involves the development of a dynamic control model of the system, the design of a suitable control strategy with the help of this model, the implementation of the control strategy in an actual factory process control system and evaluation of its performance.

1.1 Background

Traditionally fruit purees are produced by concentrating the fruit pulp in a forced recirculation evaporator, which are known to be detrimental to product quality because of their large residence time and high temperatures.

We have developed a novel fruit concentration process involving the use of a separation step to separate the fruit pulp into a cake fraction and a serum fraction. Because of the absence of fibres in the serum fraction, the serum fraction can be concentrated at lower temperatures, thus maintaining its freshness. Simultaneously, the fibres are not exposed to high shear and will therefore better maintain their water-binding properties.

A draw-back of the novel fruit concentration process is that the cake fraction and the concentrated serum fraction need to be recombined into a homogeneous fruit paste in a controlled way. At the same time, the Brix and viscosity of the resulting fruit concentrate need to be maintained within normal food standards. For this reason, a systematic approach is required in order to design an effective control strategy.

1.2 Approach

Plantwide control is a concept or methodology to systematically build control structures of large, continuous processes with complex interactions [1, 2, 3]. Process modelling and simulation form an important aspect of the method. It follows a series of logical steps which are easy to apply in practice:

1. Definition of Operational Control Objectives
2. Manipulated Variables and Degrees Of Freedom
3. Primary Controlled Variables
4. Production Rate
5. Regulatory Control Layer
6. Supervisory Control Layer
7. Optimization Layer
8. Validation

We applied most steps of this methodology in conjunction with a conceptual process design methodology [4] to ensure that the integration of process, equipment and control design is optimal. This paper will focus on the first five steps. In chapter 2 we will go systematically through steps 1 to 5 given above based on the process given in figure 2.1. Next, we will present the assumptions behind the underlying mathematical model that we used to determine the best control strategy. Finally we will present the results from the industrial process that we implemented in one of our factories.

2. PLANTWIDE CONTROL PROCEDURE

2.1 STEP 1: Overall control objective. Identify operational constraints.

The process control aims at producing high grade, constant quality fruit paste. Good quality paste has a nice red

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[1] This is in contrast with plantwide automation which addresses the design and implementation of Industrial Process Control and Automation Systems (IPCAS).

colour, constant Brix (sugars) and Bostwick (viscosity) values and also good taste and flavour. In order to achieve this, it is necessary to control the product flow throughout the plant smoothly and to minimise the hold-up time of product in the system. The following process and control system requirements were considered:

Process Requirements

Maximization of assets. Maximization of plant output is a key objective since fruit concentration processes are mostly seasonal continuous operations. It is therefore important that the bottleneck is 100% utilised. In our case, the evaporator is the bottleneck of the process.

Process reliability. There are many factors in the process not directly related to the control system or strategy, which influence the control performance. A major factor is the overall factory reliability (equipment and utilities). The novel process is more complex and especially vulnerable for unplanned stoppages. Both factory reliability and the control system robustness towards unplanned stoppages are essential aspects to consider in the design process.

Equipment restrictions. It is industry practice to operate decanters with a constant in-flow in order to get a stable liquid/solid separation. It is recommended that feed flow variations are less than ±20%. It is also industry practise to operate forced recirculation evaporators with a constant steam supply (constant evaporation rate).

Fruit quality fluctuations. The fibre- and sugar content of raw fruit juice is known to vary up to ±20% on an hourly basis. The fibre- and sugar content vary independently.

Control system requirements

End-product variability Brix. The Brix boundaries should be less than 1.0 Brix from controller setpoint for commercial purposes.

End-product variability Bostwick. The Bostwick should be the same or lower as normal variations in standard paste (typically +/- 0.5 cm). This implies that there should not be *more* variation in the ratio of sugars to fibers in the puree as in the original fruit.

Simple control. The control strategy must be able to be implemented in the SIEMENS Control System and must be relatively easy to understand for the operators. For this reason, the control strategy is based on standard PID-controllers.

2.2 STEP 2: Manipulated variables and degrees of freedom.

According to the principles by Skogestad, we will identify dynamic and steady-state Degrees Of Freedom (DOF) based on the process given in figure 2.1 according to the equation below:

$$N_{ss} = N_m - (N_{0m} + N_{0y})$$

In which:

N_{ss}	Number of steady state DOFs
N_m	Number of dynamic (control) DOFs (valves, pumps)
N_{0m}	Number of manipulated input variables with no steady state effect
N_{0y}	Number of output variables with no steady state effect (product levels)

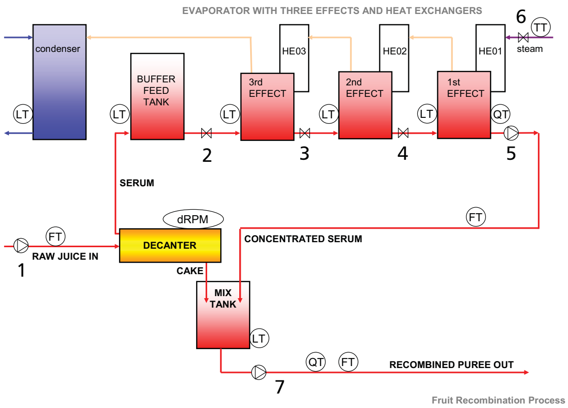


Figure 2.1: Process Flow Diagram of novel fruit concentration process with Degrees Of Freedom (DOFs).

As can be seen from figure 2.1, 7 dynamic DOFs (N_m) can be identified. Since the process consists of 5 tanks, there are 5 tank levels that need to be controlled (N_{0y}), but that do not contribute to the steady state mass- and composition balance. This means that there are two remaining DOFs (N_{ss}) to control the whole process. This is identical to an existing fruit concentrate process in which the two DOFs are used to control the Brix of the puree and the production rate. However we know that the dynamics of the novel process have changed, because we have combined a slow evaporation process with a fast separation process in one additional process step (the mix tank). Therefore a dynamic model is essential to evaluate the new process dynamics and possible process control structures.

2.3. STEP 3: Primary controlled variables.

Skogestad states that the primary variable to control is the active constraint. According to our control objectives, this is overall production rate as set by the evaporator. Another key primary controlled variable is the Brix of the resulting fruit puree. The Brix is given by the following equation:

$$\phi_{puree} \cdot Brix_{puree} = \phi_{cake} \cdot Brix_{cake} + \phi_{concentrated\ serum} \cdot Brix_{concentrated\ serum}$$

in which:

ϕ_{puree}	Flow of puree [kg/s]
$Brix_{puree}$	Brix of puree [Brix]
ϕ_{cake}	Flow of cake [kg/s]
$Brix_{cake}$	Brix of cake [Brix]
$\phi_{concentrated\ serum}$	Flow of concentrate serum [kg/s]
$Brix_{concentrated\ serum}$	Brix of concentrate serum [Brix]

The flow rate of cake is mainly driven by the amount of fibres in the fruit juice as well as the way the decanter is operated (level of drying of the cake), whereas the Brix of the cake is driven by the fruit variety, ripeness and other agronomical factors. Both Brix and flow rate of concentrated serum are highly dependent on the way that the evaporator is operated and controlled. The chance of product that is out of specification is high since it is dependent on four variables.

2.4 STEP 4: Set of production rate.

The choice where to set the production rate determines the structure of the remaining inventory (level) control system. The production rate should be set at the (dynamic) bottleneck of the process as explained in section 2.3. For the traditional fruit concentrate process, the production rate is determined by the slow evaporator (dynamic bottle-neck). The evaporator is operated at a fixed steam pressure (constant evaporation rate). The up-stream and down-stream processes follow the evaporation rate. Since the end product Brix is controlled through manipulating the evaporator out-feed pump, this means the product flow through all unit operations is changing constantly dependent on the incoming Brix of the raw juice.

On the other hand, it is industry experience that decanters require a constant feed (within 20% of the main flow) to have the best performance. This requires controlling the flow towards the decanter and setting the production rate here, through a flow controller.

In order to control the level in the evaporator feed tank, therefore two options do exist:

A. **Control the level via manipulation of the steam supply** (thus via manipulation of the evaporation rate). No constant steam supply means that the evaporator cannot be operated at constant evaporation rate and temperatures. The potential negative impact should be evaluated and preferably minimized.

B. **Control the level via manipulation of the inflow to the decanter.** It is clear that in this scenario no constant inflow to the decanter can be guaranteed and that the potential negative impact on separation performance should be evaluated and minimized. It is noted that controlling the level through decanter serum outflow will disturb the separation process and therefore is not feasible.

Either scenario clearly forced us to deviate from the industry practice w.r.t equipment operation.

Already at this stage of the methodology we decided to develop a dynamic model based on the following arguments:

· The choice of setting the production rate highly influences the dynamics of the overall process
· There is a direct correlation between productivity and end-product composition in concentration processes
· The recombination process is highly vulnerable to variations
· We deviate from the industry standard w.r.t. equipment operation for either the decanter or the evaporator.
· The design of the control system will influence the process design, like the design of the recombination tank.

The model must involve the complete regulatory control layer in order to decide on the best control strategy. In section 2.5, we will provide the results from the model simulations and we will demonstrate how we used this to evaluate the various control strategies.

2.5. STEP 5, 6 and 8: Regulatory and Supervisory Control layers and Validation.

Section 2.4 showed that the production rate could not be defined because of the strong interactions between production rate, inventory control and product composition. It was also shown that there are two DOFs. With these two DOFs, a steady state analysis demonstrates that we can control the production rate and the Brix of the fruit puree in the following way:

1. Puree Brix control through manipulation of the concentrated serum Brix. Steady state mass- and composition balances show that an increase in concentrated serum Brix will increase the Brix of the resulting puree. Based on this, a cascade control loop can be designed. In this way, no extra DOF needs to be created and no additional equipment is required.

However, if the dynamics of this control structure prove to be unsuitable to control the Brix of the fruit puree, one additional DOF will be required in order to make the process controllable. This DOF should have sufficiently fast response time and a sufficient process gain. We have designed two alternative ways to create one extra DOF:

2. Puree Brix control through manipulation of decanter settings. By changing the differential speed of the screw inside the decanter [5] the amount of juice incorporated inside the cake can be controlled. Reducing the differential speed will send less water via the cake to the recombination tank and the Brix of the resulting puree will increase.

3. Puree Brix control through addition of juice to recombination tank. A third stream of juice can be added to the recombination tank to dilute a slightly over-concentrated puree to the specified composition.

All scenarios are summarized in table 2.1.

Table 2.1: Overview of control scenarios.

Production rate / Brix control	Extra DOF created?	A. Control the level via steam supply	B. Control the level via inflow to the decanter.
1. Puree Brix control through manipulation of the concentrated serum Brix	No	A1	B1
2. Puree Brix control through manipulation of decanter settings.	Yes	A2	B2
3. Puree Brix control through addition of juice to recombination tank	Yes	A3	B3

Despite the fact that scenarios A1 and B1 are feasible based on steady state, an inverse response occurs initially: Upon increasing of the concentrated serum Brix (see figure 2.2 at t=10 hours), the outflow pump first slows down in order to increase the residence time in the first effect. The result is an inverse response of the puree Brix. This clearly demonstrates that in concentration processes, flow and concentration level are inversely proportional.

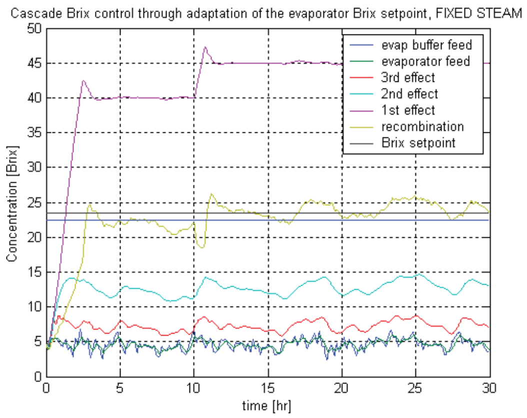


Figure 2.2: Dynamic model simulation of scenario A1. The inverse response in puree Brix (denoted in yellow-green as 'recombination') at t=10 hours is clearly visible.

The conclusion is that the additional DOF is required (Scenarios A2, B2, A3 and B3). In order to decide on the optimal control strategy, we will first evaluate the A2 and B2-strategies against the A3 and B3-strategies.

For strategies A2 and B2 the dynamic behaviour and operating window of the decanter needs to be known. Based on step response measurements on the serum exiting the decanter (see figure 2.3), we estimated that the response time of the cake fraction upon changes in differential speed is in the order of 300 seconds (three times the response time of the serum fraction).

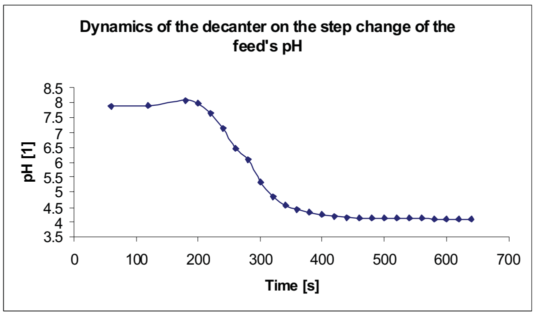


Figure 2.3: Response time of decanter serum exit on a step change in pH via citric acid addition (at 170 s).

Figure 2.4 shows the response behaviour of the solids concentration in the cake exiting the decanter as a function of changes in the differential speed (courtesy of GEA Westfalia). It can be seen that there is a strong response, but only in a small operating window. Also can be seen that the machine operation is highly non-linear.

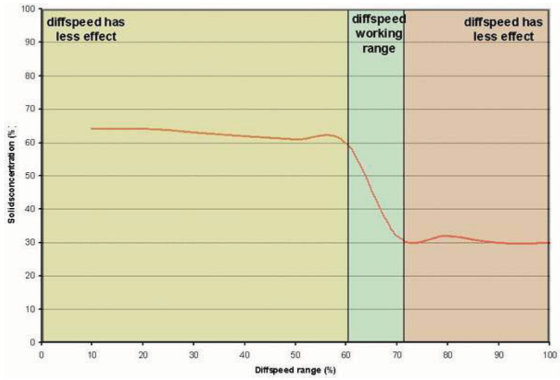


Figure 2.4: Response of solids concentration in the cake from a decanter. Courtesy of GEA Westfalia GmbH.

The results of the simulations with the above time delay and response behaviour are shown for scenario A2 in figure 2.5. As can be seen the fruit puree Brix cannot be kept within the desired range. Scenario B2 (not shown) behaved in a similar way. This means that the decanter is too slow to be used in a control loop, thus rendering strategies A2 and B2 ineffective.

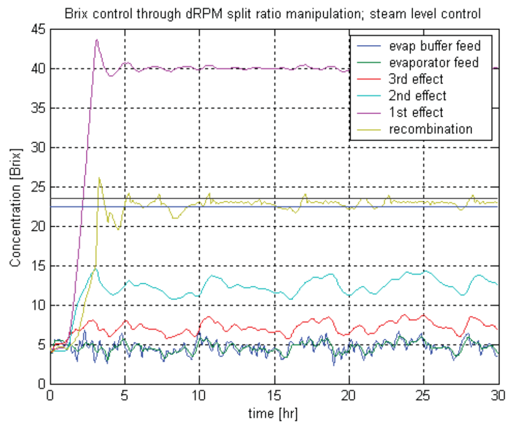


Figure 2.5: Dynamic model simulation of scenario A2. Fruit puree Brix control through manipulation of decanter settings. It can be seen that the Brix of the fruit puree (denoted in yellow-green as 'recombination') cannot be maintained within the desired range of ±1 Brix around setpoint.

We now need to decide between the scenarios A3 and B3. For a stable process, it is important that both concentrated serum and cake are recombined according to its natural ratio. This means that there cannot be an excess flow of either cake or concentrated serum in the mix tank. In this way no changes in hold-up volume of either the cake fraction or the serum fraction should occur. The only location where significant changes in hold-up volume can occur is the evaporator feed tank.

Figure 2.6 shows the level in the evaporator feed tank for scenarios A3 and B3. It can be seen that both control strategies can control the level in this tank, but control strategy A3 requires a much longer time upon start-up to reach a stable level (> 5 hours). This means that this strategy is very slow due to the slow response of the evaporator upon changes in steam supply. For this reason, this control strategy is very vulnerable to disturbances and not recommended.

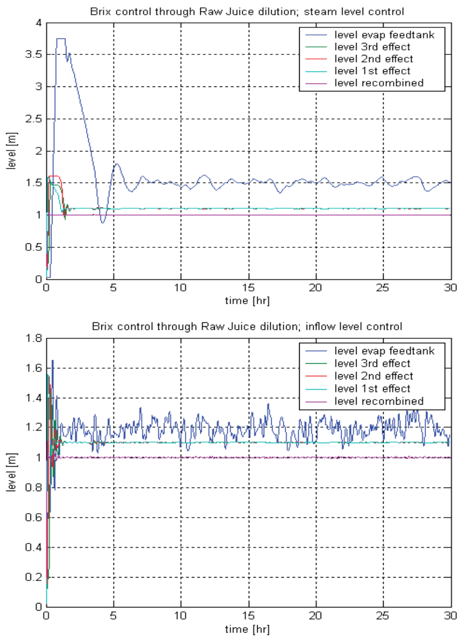
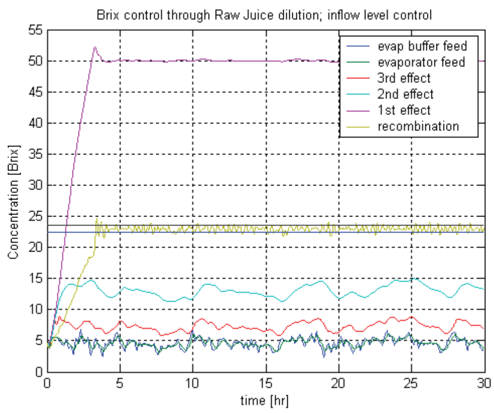


Figure 2.6: Dynamic model simulation of scenarios A3 (top) and B3 (bottom). It can be seen that the time delay to reach a setpoint of the evaporator feed tank is in the case of scenario A3 is in the order of 5 hours and in scenario B3 in the order of 1 hour.

Scenario B3 gave the right dynamic behaviour to control the evaporator feed tank, allowing for a proper recombination of concentrated serum and cake. Figure 2.7 demonstrates that this strategy is also able to control the Brix of the fruit puree. Also, it can be seen that the variation of the ratio of sugars to fibers is within variations found in the raw material. Thus control strategy B3 is most suitable for this process.



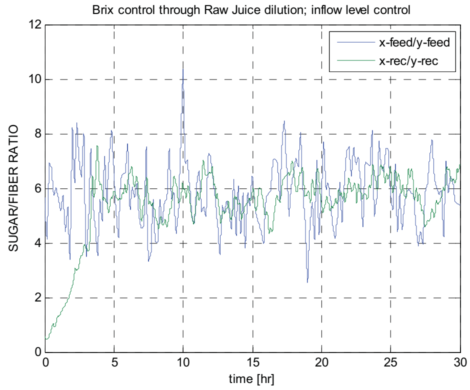


Figure 2.7: Dynamic model simulation of scenario of control strategy B3. It can be seen that the Brix of the fruit puree (denoted in yellow-green as 'recombination') can be maintained within the desired range of ±1 Brix around setpoint (left bottom). The ratio of sugars to fibers in the feed (x-feed/y-feed) and in the fruit puree (x-rec/y-rec) are shown above.

Proper tuning of the level control loop of the evaporator feed tank further contributes to the overall process stability. In order to minimize fast fluctuations in decanter inflow, it was decided to use a P-algorithm and to set the gain value relatively low. In this way, the evaporator feed tank can be used as a real buffer without impacting on the ratio of sugars to fibers. Also feed fluctuations to the decanter are minimised.

The resulting process flow diagram for strategy B3 with all sensors, actuators and control loops is given in figure 2.8.

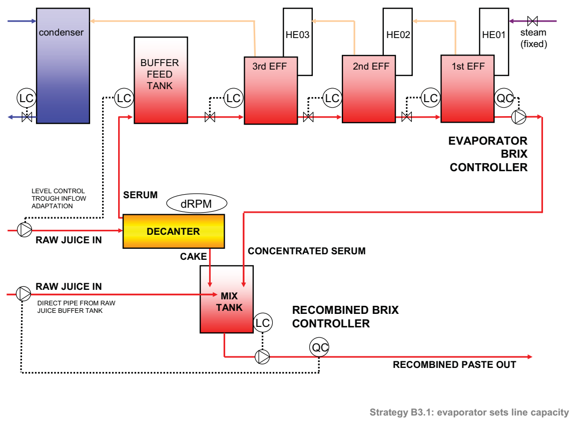


Figure 2.8: Process flow diagram with for strategy B3 with sensors, actuators and the final control structure.

3. DYNAMIC MODEL

We built a non-linear dynamic model in Matlab/Simulink [6] based on:

- Dynamic mass- and composition balance for all tanks
- Dynamic heat balance for all heat exchangers with the heat transfer coefficient estimated based on real-time factory data
- Steady state equations for the decanter with a first order transfer function with lag time
- Band-limited white noise in feed sugar and fibre levels of ±20% around the average
- PID-feedback control

4. IMPLEMENTATION INTO THE FACTORY

4.1 IPCAS implementation

The process and control strategy as defined before was implemented in a Unilever factory. To this purpose a "User Requirements Specification of the Industrial Process Control and Automation System (IPCAS)" was written. It specifies the Industrial Process Control System (PLC/SCADA process computer hardware and software), instrumentation and installation infrastructure.

4.2 Evaluation of control strategy B3 under factory conditions

The performance of the selected control strategy that was implemented in the SIEMENS PLC/SCADA system was evaluated during actual production.

Figure 4.1 shows that for the given day, the whole production was within specifications. During one whole season, the average Brix was 22.92 (with a setpoint of 23) with a standard deviation of only 0.3.

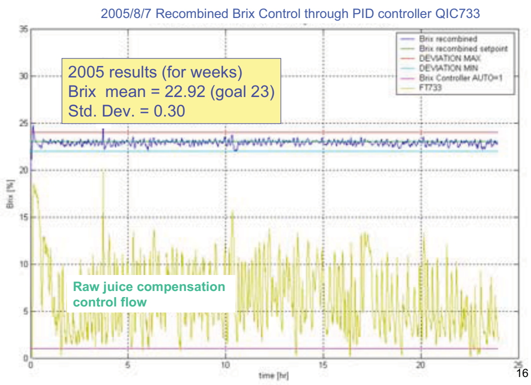


Figure 4.1: Factory data of fruit puree Brix (named QIC 733 in factory control system). Also the flow rate (named FT733 in the factory control system; raw juice compensation control flow) of the added juice is depicted. This flow rate should be multiplied by 100 and is expressed in kg/hour.

5. CONCLUSIONS

In this paper we presented a novel fruit concentrate production process involving a decanter, an evaporator and a recombination process. The choice for this process results in complex, non-linear, process dynamics. Such processes can be difficult to control and a systematic methodology was required.

We demonstrated that a strategy in which the Brix is controlled via addition of a third juice stream was the best choice for the given process. Evaluation of the control strategy under real factory conditions showed that the control strategy is very robust and that end-product specifications are met.

This case demonstrated to us that it is important to integrate control strategy design in an early stage with process and equipment design. Furthermore we required a non-linear dynamic model to understand the complex dynamics of the process and to design an appropriate control strategy.

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