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A General Method for Defining and Structuring Buffer Management Problems

Anna Lindholm, Krister Forsman and Charlotta Johnsson

Abstract— **In an industrial plant, availability is an important factor since increased availability often gives an increase of final production, which in many cases means an increased profit for the company. The purpose of using buffer tanks is to increase the availability either by separating production units from each other or by minimizing flow variations. However, the methods for achieving this goal is not trivial, and depend on the specific characteristics of the problem. This paper contributes to structuring the general buffer management problem for continuous chemical plants and suggests methods for solving some specific problems, presented as a case study at Perstorp AB, Sweden.**

I. Introduction

Production processes can generally be classified as continuous, discrete or batch. Briefly, processes that have a continuous outflow, e.g., production of energy or paper, are classified as continuous processes. Discrete processes have a discrete output whereas the outcome of a batch process is neither continuous nor discrete, instead it consists of a set of items, i.e., a batch.

Continuous chemical plants are normally composed of one or several production units such as reactors, evaporators and distillation columns. It is common to place buffer tanks between the production units for material balance control [1]. The buffer tanks serve several different purposes. The most common use of buffer tanks are [2]:

1. To separate production units from each other. The aim is to ensure that the production units can operate independently from each other. A buffer tank placed before the bottlenecking production unit[1] could be used to ensure that this production unit does not have to reduce its operating speed, even if a preceding production unit suffers a shutdown. Ensuring that the bottleneck unit operates at its maximum speed as much as possible can make a significant gain in productivity.
2. To minimize flow variations in the in- or outflow of the buffer tank. Flow variations often cause poor behavior or failure of sensitive production units. A buffer tank placed before or after a sensitive production unit can improve its behavior and increase its availability.

The two purposes described above often give contradictory demands on the level controller of the buffer tank. For purpose 1, a setpoint close to the top or bottom of the tank is often desired, which requires aggressive level control, while purpose 2 is to minimize flow variations, i.e. to have as non-aggressive control as possible.

The introduction of buffer tanks between production units can increase the productivity and availability of a plant and this motivates the need for good buffer management strategies. Tuning of simple controllers, such as P- or PI-controllers, for automatic level control is a common approach for improving buffer management. The objective is keeping the manipulated flow as smooth as desired while keeping the level within some high and low limits.

This paper will focus on the conditions and criteria that have to be known and/or selected in order to be able to make efficient tuning of buffer tank level controllers. PI-controllers, which are the most commonly used controllers for process control [3], have been used in the case study in Section IV. PI-controllers have also shown to have good performance for some buffer management problems [4], [5]. The paper will show where research has been done and in which areas further work could be done. A case study example at Perstorp is included to describe possible methods for tuning level controllers at an industrial plant.

II. Problem Structure

To be able to make efficient tuning for the control loops of the buffer tanks, the production process itself, the disturbance model, the optimization criteria, and the constraints for the optimization have to be known. The key elements are:

A. Production process
- Topology
- Tank dimensions
- Maximum and minimum limitations on flows
- Start-up times

B. Disturbance model
- Stochastic or deterministic

C. Optimization criterion
- Definition of production
- Definition of flow variations

D. Constraints
- Freedom in instrumentation
- Local or global problem
- Set of solutions to be considered

A. Production process

The production process that the optimization problem should be solved for, referred to as the relevant production

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[1]The production unit with the slowest maximum production speed.

process, can consist of one or several buffer tanks. If the relevant production process includes more than one buffer tank its topology has to be considered. The buffer tanks could be connected in series, or they could have a setup with recycle flows from one tank to another. Depending on the location of the recycle flow the solution to the buffer management problem is affected in various ways. The tank dimensions of all buffer tanks in the topology are of relevance for the solution to the buffer management problem. In most cases there are also constraints on maximum and minimum flows in the relevant production process. Limitations on maximum flows can often be related to maximum production speed, whereas minimum flow limitations are given by the functionality of the equipment. Almost all chemical apparatuses have a minimum rate at which they can be run. Different production units in the relevant production process could also have different start-up times. If the start-up times are included in the problem formulation, the optimal solution to the problem could be different from the solution to the original problem, where start-up times were not considered.

B. Disturbance Model

The choice of disturbance model will affect the optimal solution to the buffer management problem. A deterministic model could consider for example step disturbances at a specific location in the relevant production process. Stochastic modeling of disturbances could be models such as random walk or stationary random processes.

C. Optimization criterion

The choice of optimization criterion is an important part in the definition of the problem. Two parameters that affect the optimization are the definition of production and the definition of flow variations.

A common choice to define production is as final production (output) of the relevant production process. However, in some cases it is preferable to define production as throughput in the bottleneck unit. In the long run these two optimization criteria will be approximately equal since all units will be able to catch up with the bottleneck unit in the sense of production speed. If the difference between the capacity of the bottleneck unit and the unit with second slowest maximum production speed is small, the time for the bottleneck to catch up will be very long though.

Common definitions of flow variations are $\mathrm{Var}[\dot{u}(t)]$, which penalizes frequent flow variations, and $\max_t |\dot{u}(t)|$, which penalizes large variations in the flow[2].

D. Constraints

Before solving the buffer management problem some constraints must be set. One issue is the freedom in changing the instrumentation of the relevant production process. Are the sensors and actuators fixed or can their location be changed to achieve better performance? In most cases the placement of the actuators are given by the design [6].

[2]This definition is often denoted Maximum Rate Of Change, MROC.

The level of the solution also has to be selected. Should the problem be solved for one single unit (local problem) or for several units (global problem)?

In addition, the set of solutions to be considered must be specified. The set can consist of for example PI-controllers, PID-controllers or all linear controllers. The set can also be limited to include only controllers tuned using a specific tuning method.

III. Previous work

Several researchers have considered the averaging level control problem, i.e. developed a control strategy to solve the buffer management problem where the objective is to have small variation of the outlet flow of a buffer tank. To mention some:

Cheung and Luyben [7] developed a strategy for tuning P- and PI-controllers with peak level height and maximum rate of change of outflow as specifications, for both single tanks and cascades of tanks.

Ogawa [4] considered averaging level control of one tank with $\mathrm{Var}[\dot{u}(t)]$ and $\mathrm{Var}[u(t)]$ as definitions of flow variations and disturbances modeled as random walk and a stationary random process. The optimization is performed over all linear controllers.

Lee and Shin [5] considered averaging level control of one tank using a PI-controller with flow variations defined as maximum rate of change of the control signal and a deterministic disturbance model; a step disturbance in the unmanipulated flow of the buffer tank.

The wide variety of options when choosing disturbance model, optimization criterion and constraints makes the general buffer management problem very complex. Today the problem has been solved for a number of combinations but there are still many gaps to be filled. To our knowledge little work has for example been done on solving buffer management problems for production processes including more than one tank (global problem), and for production processes where recycle flows or start-up times are considered. Further, only limited amount of work has been done regarding general methods for defining and structuring buffer management problems. This paper contributes with such a method and it presents an industrial case study where this method is applied.

IV. Case study at Perstorp AB

A case study at Perstorp UK Ltd was performed to suggest some methods for buffer management at an industrial plant using the problem structure defined in this paper. According to this structure the problem investigated in the case study could be stated as:

A. Production process
- Topology: Without recycle flows
- Tank dimensions and maximum limitations on flows are given.
- Minimum limitations on flows and start-up times are considered in an ad hoc manner.

B. Disturbance model
   - Deterministic: Step disturbances in the uncontrolled inflow of the tank

C. Optimization criterion
   - Production defined as maximum throughput in the bottleneck unit.
   - Definition of flow variations: $\max_t |\dot{u}(t)|$

D. Constraints
   - Instrumentation may not be changed.
   - The local problem is considered (one buffer tank).
   - PI-controllers with T_a-tuning with $T_i = T_a$ are considered. More about T_a-tuning in Appendix.

Motivations of the choices are given below.

A. Production Process

The production process studied in the plant at Perstorp UK Ltd in Warrington produces per-acetic acid from 70 % hydrogen peroxide. The simplified model of the production process that has been studied is shown in Figure 1. The production process is already existing and can not be modified.

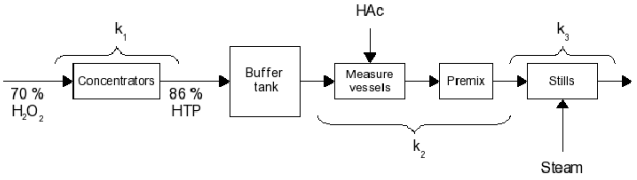

Fig. 1. Studied production process.

The production process includes one buffer tank, containing High Test Peroxide (HTP). This buffer tank was manually level controlled as the study began. The other production units in the production process have been approximated with a constant ratio to the incoming flow based on process data from normal operation (see Figure 1).

The master flow of this production process is the incoming flow of 70 % hydrogen peroxide, which is concentrated to 86 % HTP. HTP at these high concentrations can at certain conditions cause hazardous explosions and therefore a lot of safety switches control the operation of the concentrators. One of them redirects the flow of 86 % HTP if the level in the buffer tank rises above 55 %[3]. The concentrators are bottlenecking the production process, and consequently a redirect is equivalent with a loss of production in the production process. It is desired to keep the level of the buffer tank within 10% to 55% of the tank.

B. Disturbance Model

The major disturbance acting on the process was found to be changing of the feed to the concentrators due to changed production rate. This change is step-like and motivates the choice of step disturbances in the inflow as the disturbance model. By studying data from normal operation during a longer time-period, amplitudes of the maximum expected disturbances upwards and downwards can be found in in terms of percent of the manipulated variable. These disturbances represent the worst-case scenario of disturbances.

[3]A level of 55 % corresponds to the maximum amount of HTP allowed to be stored at the same location.

C. Optimization criterion

The main objective in the case study is to increase the average production rate of the production process by increasing the availability of the bottlenecking unit, the concentrators. This can be seen as separating the concentrators from its downstream processes, i.e. purpose 1 of the buffer tank as defined in the introduction of this paper. Production is thereby defined as maximum throughput in the bottleneck unit. The availability is increased by minimizing the redirect periods by keeping the level of the buffer tank within specified limits.

Flow variations are defined as $\max_t |\dot{u}(t)|$ and a maximum tolerated value is set considering process data.

D. Constraints

The problem is to be solved with the present instrumentation, i.e. the locations of sensors and actuators are considered to be fixed.

The suggestion is to control the level of the buffer tank by manipulating the steam to the stills, to ensure that the level stays within 10 % to 55 % of the tank. This level control is possible since a certain amount of steam to the stills will give a certain outflow of the buffer tank. A larger steam flow will cause the stills to work faster and thus increase the outflow of the buffer tank.

PI-controllers tuned with T_a-tuning with $T_i = T_a$ are considered since they are easy to implement and have shown to yield good behavior for similar production processes [5]. It is also an intuitive tuning method since it consists of only one tuning parameter. Other tuning methods could also be applied, giving a solution in the same manner as the one presented below.

Solution

To solve the problem the deviation from the setpoint for the level controller of the buffer tank due to the specified disturbances must be computed. In Appendix it is shown that the maximum level deviation due to an inlet flow step disturbance of C % of the manipulated variable is $e_{max} = \frac{1}{\sqrt{2}} e^{-\frac{\pi}{4}} k_v T_a C$, for an integrating process with transfer function $G_P(s) = \frac{k_v}{s}$ if T_a-tuning with $T_i = T_a$ is applied. In this case the maximum tolerated deviation depends on the setpoint. If for example the setpoint, SP, is 30 % we can tolerate a positive deviation, e_{max}^{up}, of 25 % and a negative deviation, e_{max}^{down}, of 20 %.
This can be expressed as

$$e_{max}^{up} = 55 - SP$$

$$e_{max}^{down} = SP - 10$$

Upper bounds on the arrest time are given by

$$T_a^{up} \leq \frac{\sqrt{2}}{e^{-\frac{\pi}{4}}} \frac{e_{max}^{up}}{k_v C^{up}}$$

$$T_a^{down} \leq \frac{\sqrt{2}}{e^{-\frac{\pi}{4}}} \frac{e_{max}^{down}}{k_v C^{down}}$$

In Appendix it is also shown that the maximum rate of change of the control signal due to an inlet flow step disturbance of C % of the manipulated variable is $\max_t |\dot{u}(t)| = \frac{2}{T_a} C$. This gives a lower bound on the arrest time according to

$$T_a \geq \frac{2}{\max_t |\dot{u}(t)| \max(C^{up}, C^{down})}$$

The bounds can be illustrated in a plot showing arrest time, T_a, as a function of setpoint, see Figure 2. The allowed values of the combinations of setpoint and arrest time are located inside the triangle given by the three lines. The optimal choice of arrest time and setpoint among these allowed combinations depends on the probability of even larger or more frequent disturbances than the specified versus the wear on the equipment and the effect on downstream processes when having rapid changes in the control signal. Another aspect is start-up times of upstream or downstream processes. In the real plant the setpoint is usually chosen as high as 45 % by the process operators to ensure that the reactors, located downstream of the stills, with start-up time up to 4 hours, do not have to be shut down.

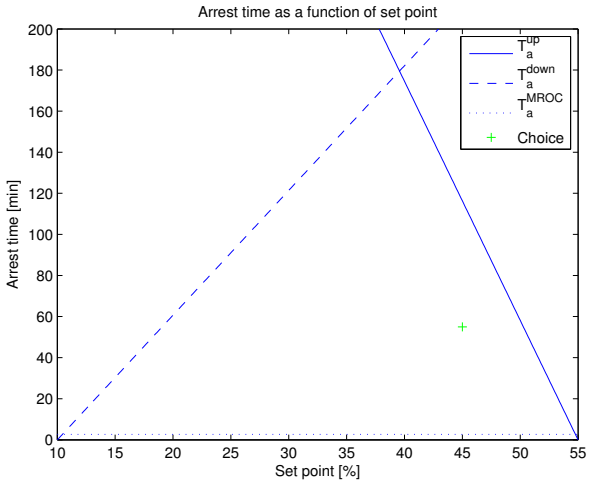

Fig. 2. Arrest time as a function of setpoint for tuning of the buffer tank level controller.

Evaluation

A level controller for the buffer tank was implemented in October 2008. The choice of arrest time was 55 minutes and is marked as a cross in Figure 2 for the commonly used setpoint 45 %. Data from operation during November of 2008 can be viewed in Figure 3. The redirect periods, when the level of the buffer tank reaches 55 %, that previously appeared around 10 times a month now very seldom occur, since a reliable level controller is implemented which successfully separates the concentrators from its downstream processes. This improves the availability of the production process, where the concentrators is the bottleneck unit. It can be seen in Figure 3 that the chosen arrest time gives quite tight level control and not much of the buffer tank volume is utilized, resulting in a not so smooth control signal, i.e. steam flow. The choice of tight level control is an initial choice to investigate the behavior of the controller during a longer time period and make the process operators familiar with the new controller. After some time of operation, the controller could gradually be detuned until optimal behavior is achieved.

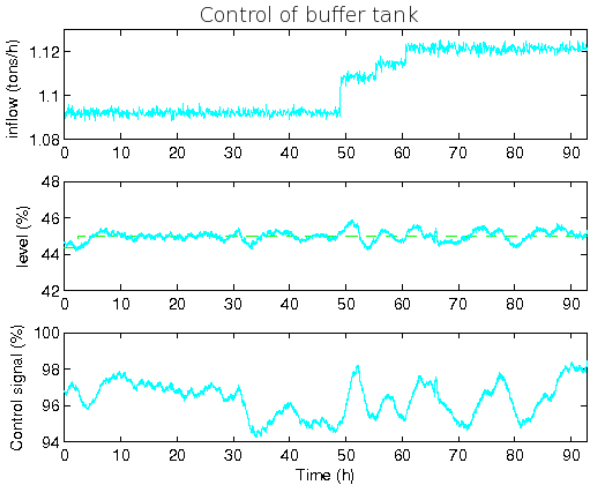

Fig. 3. Operation of the level controller for the buffer tank with $T_a = 55$ min

When using the method described above, a tuning of the level controller to handle worst-case disturbances is obtained. However, these disturbances might rarely occur and the controller will then most of the time be too aggressive. A possibility that allows better utilization of the buffer tank is to introduce some kind of gain scheduling.

V. Conclusions and Future Work

Buffer management has shown to be an important factor within the process industry and good strategies for buffer handling can possibly increase both the availability of the plant and the quality of the end product. The purpose of the buffer tanks, separating production units from each other or minimizing flow variations, will decide the goal, the optimization criterion, of the problem. The strategy for achieving the goal has to be decided in terms of control strategies to be considered as possible solutions. Before being able to solve the problem there are many constraints and model parameters to be defined. In this paper a method for defining and structuring a buffer management problem is suggested. When having defined the parameters of the process model, the disturbance model, the optimization criterion and the

constraints, the specific buffer management problem can be solved. A case study performed at Perstorp AB confirms that the methodology could be useful for industrial applications.

The challenge in solving the general buffer management problem is in taking the complexity of the problem, given by the numerous possible combinations of constraints, into account.

What has been done today is to solve the problem for some specific sets of constraints but there are still many problem formulations left to be solved. Taking start-up times into account or including recycle flows in the problem formulation would be two interesting areas of research. Buffer management for production processes with multiple tanks are also an interesting topic that is currently investigated. This paper contributes to structuring the multi-dimensional map of buffer management problems with different constraints and giving an example of a strategy for solving a specific buffer management problem at an industrial plant.

APPENDIX

T_a-tuning for integrating processes

A method for tuning PI controllers which has shown to yield good result for averaging level control is the T_a-tuning method for integrating processes. The parameter T_a represents the arrest time, which is the time elapsed before the process value starts to return to the setpoint after the occurrence of a step disturbance, d, in the unmanipulated flow (see Figure 4 and 5). For an integrating process with transfer function $G_P(s) = \frac{k_v}{s}$ and a PI controller with transfer function $G_C(s) = K_c(1 + \frac{1}{sT_i})$, the choice of parameters when using T_a-tuning is

$$K_c = \frac{2}{k_v T_a}, \quad T_i = 2T_a$$

which will give a double pole in $-\frac{1}{T_a}$ and a zero in $-\frac{1}{2T_a}$. A common choice is to choose the integral time to be half of that in original T_a-tuning, i.e. $T_i = T_a$, yielding two complex conjugated poles in $-\frac{1}{T_a} \pm \frac{1}{T_a}i$ and a zero in $-\frac{1}{T_a}$. The behavior of T_a-tuning with $T_i = 2T_a$ and $T_i = T_a$ respectively is shown in Figure 4.

Below, the maximum deviation from the setpoint and the maximum rate of change of the control signal when the process suffers a step disturbance of C % of the manipulated variable is derived for T_a-tuning with $T_i = T_a$.

With T_a-tuning with $T_i = T_a$ the transfer function from a step disturbance d entering before the process to the control error e becomes:

$$G_{ed}(s) = \frac{G_P(s)}{1 + G_C(s)G_P(s)} = \frac{k_v s}{s^2 + \frac{2}{T_a}s + \frac{2}{T_a^2}}$$

When d is a step disturbance of C % of the manipulated variable we get

$$E(s) = G_{ed}(s)\frac{C}{s}$$

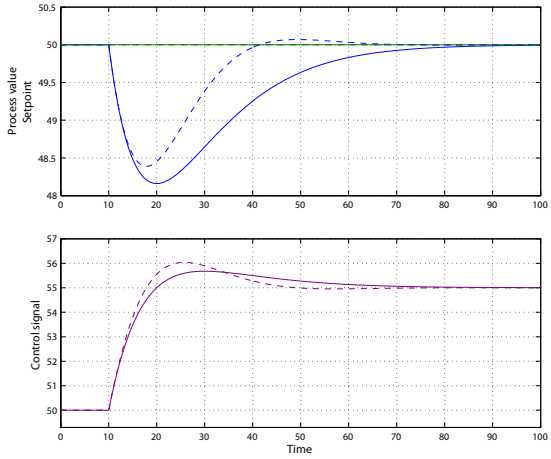

Fig. 4. The solid lines represent original T_a-tuning with $T_i = 2T_a$, the dashed T_a-tuning with $T_i = T_a$. The arrest time is here T_a =10 time units.

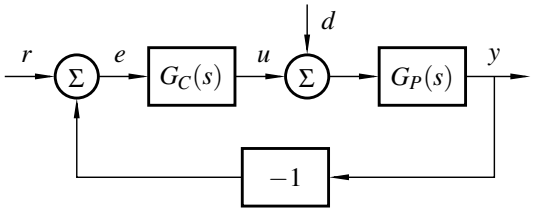

Fig. 5. System setup.

Inverse Laplace transform gives

$$e(t) = k_v T_a e^{-\frac{1}{T_a}t} \sin(\frac{1}{T_a}t)C$$

with the derivative

$$\dot{e}(t) = k_v e^{-\frac{1}{T_a}t}[\cos(\frac{1}{T_a}) - \sin(\frac{1}{T_a})]C$$

The derivative is equal to zero if and only if $t = (\frac{\pi}{4} + n\pi)T_a$ where n is an integer. The maximum control deviation is obtained for $n = 0$ and has the value

$$e(\frac{\pi}{4}T_a) = \frac{1}{\sqrt{2}}e^{-\frac{\pi}{4}}k_v T_a C \approx 0.3224 k_v T_a C$$

The closed loop transfer function from d to the control signal, u, is

$$G_{ud}(s) = \frac{\frac{2}{T_a^2}(T_a s + 1)}{s^2 + \frac{2}{T_a}s + \frac{2}{T_a^2}}$$

Given that d is a step disturbance of C % of the manipulated variable we get

$$U(s) = G_{ud}(s)\frac{C}{s}$$

Inverse Laplace transform gives

$$u(t) = [1 + e^{-\frac{1}{T_a}t}(\sin(\frac{1}{T_a}t) - \cos(\frac{1}{T_a}t))]C$$

with the derivative

$$\dot{u}(t) = \frac{2}{T_a} e^{-\frac{1}{T_a}t} \cos\left(\frac{1}{T_a}t\right) C$$

The second derivative of the control signal becomes

$$\ddot{u}(t) = -\frac{2}{T_a^2} e^{-\frac{1}{T_a}t} \left[\cos\left(\frac{1}{T_a}t\right) + \sin\left(\frac{1}{T_a}t\right)\right] C$$

The second derivative is equal to zero if and only if $t = (\frac{3\pi}{4} + n\pi)T_a$ where n is an integer. The expression for $\dot{u}(t)$ has its maximum value for $n = 0$ which gives

$$|\dot{u}(\frac{3\pi}{4}T_a)| = \frac{2}{\sqrt{2}T_a} e^{-\frac{3\pi}{4}} C$$

This is less than $|\dot{u}(0)| = \frac{2}{T_a}C$ which proves that

$$\max_t |\dot{u}(t)| = |\dot{u}(0)| = \frac{2}{T_a}C$$

due to a step disturbance of C % of the manipulated variable.

For more information about T_a-tuning, see [8] (in Swedish).

REFERENCES

[1] P. S. Buckley, *Techniques of Process Control*. John Wiley & Sons, Inc., 1964.

[2] A. Faanes and S. Skogestad, "Buffer tank design for acceptable control performance," *Ind. Eng. Chem. Res*, vol. 42, pp. 2198–2208, 2003.

[3] F. G. Shinskey, *Process Control Systems: Application, Design and Tuning*. McGraw-Hill, 1996.

[4] S. Ogawa, *Stochastic Averaging Level Control and Its Application to Broke Management in Paper Machines*. PhD thesis, University of British Colombia, 2003.

[5] M. Lee and J. Shin, "Constrained optimal control of liquid level loop using a conventional proportional-integral controller," *Chem. Eng. Comm.*, vol. 196, pp. 729–745, 2009.

[6] S. Skogestad, "Control structure design for complete chemical plants," *Computers and Chemical Engineering*, vol. 28, pp. 219–234, 2004.

[7] T.-F. Cheung and W. L. Luyben, "Liquid-level control in single tanks and cascades of tanks with proportional-only and proportional-integral feedback controllers," *Ind. Eng. Chem. Fundam.*, vol. 18, no. 1, pp. 15–21, 1979.

[8] K. Forsman, *Reglerteknik för processindustrin*. Studentlitteratur, 2005.