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IFAC-PapersOnLine 49-7 (2016) 514–519

Implementation of advanced control in the process industry without the use of MPC

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Abstract: In process industry, such as chemical, pulp and paper or petrochemical industry there are plenty of processes that require multivariable control. Classical control structures that handle this, for example cascade control, feedforward, ratio control, and parallel control have been used at least since the 1930s. Today, much focus in academia is on model predictive control (MPC). In this paper we discuss the comparative advantages and disadvantages of classical control structures and MPC. We also briefly discuss some related topics in plant-wide control.

Keywords: Process control, multivariable systems, control structures, decentralized control, control specifications, MPC, plant-wide control, consistency.

1. INTRODUCTION

The processes that we study here have more than one manipulated variable (MV) and/or more than one controlled variable (CV), or at least a measured disturbance. Classical control structures, such as feedforwards, cascade control, ratio control etc, handle these by combining simple SISO controllers, in more or less clever ways, or slightly extending their functionality. In a way, they can be thought of as "extended decentralized control".

Classical control structures are described extensively e.g. in Marlin (2000), Smith and Corripio (2006), as well as in Skogestad and Postlethwaite (2005), where also many other approaches are presented. Some structures can be seen as "performance boosters" compared to PI-control, while others handle truly multivariable or non-linear control problems.

MPC, on the other hand, treats all MIMO processes in the same way, as truly multivariable systems with multiple interactions.

There is no doubt that MPC is a very useful and powerful paradigm for process control. In practical applications, the vast majority of implementations are in petrochemical industry whereas it is not as common in other industries, like pulp and paper, specialty chemicals or metals.

MPC is superior to classical methods in the sense that it represents a unified systematic procedure to control design for multivariable processes. It is also superior in handling complex interactions and multiple logical constraints.

However, it also has some disadvantages. Compared to classical control structures, the cost for an MPC controller is in many cases higher, e.g. when it comes to

- costs for making process models, and finding appropriate optimization criteria and constraints
- costs for licenses
- costs for maintaining the process models

Furthermore, for a new process being commissioned, it can be very hard to design an MPC controller in the design phase of the plant. It may not be obvious how the constraints should be set for them to be non-conflictive. If there is a process simulation model available, then of course this is less of a problem, but that is not always the case.

Classical control schemes, on the other hand, are fairly straightforward to design provided that the control specifications are reasonably clear.

Some also argue that the classical control schemes are more transparent for the operators, so that they can discover mistakes and suboptimal solutions in a classical structure, whereas that is harder for an MPC solution.

2. CONTROL SPECIFICATIONS

In process control, getting accurate and complete control specifications or optimization criteria is a complication that is often underestimated. This is valid both for MPC and classical structures. It frequently happens that neither operators nor process engineers or production management can specify how they want overall controls to work in great enough detail for the control engineers to be able to design the control structures.

There are sometimes misunderstandings in the form of implicit demands that are not communicated. A common example of this is related to multiple operational regimes: there are almost always "special" situations, e.g. start-ups or grade changes, which require dedicated control solutions. It can make a big difference if a controller should be possible to run in Auto during start-up or if it is acceptable that the operator runs the process manually then.

Peer review under responsibility of International Federation of Automatic Control.
10.1016/j.ifacol.2016.07.394

Another example is when there are several process streams, and depending on process capacity requirements it should be possible to run the streams in different combinations. Should the controllers, e.g. buffer level controls, handle all combinations of streams?

A simple approach is just to say that "the controllers should handle all possible cases", but that will almost always result in overly complicated control structures.

To illustrate this point, consider a simple flow split process, as showed in Figure 1.

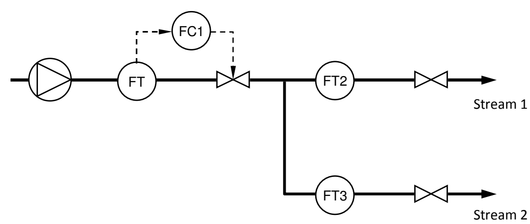

Fig. 1. Flow-split process.

The total flow is controlled by flow controller FC1. Obviously it is not consistent to also control the individual flows in stream 1 and 2.

A common and simple control specification for this scenario is that it that one of the flows is free and the other one is controlled in ratio against the free flow. This is often called ratio control, showed in Figure 2. FC3 gets it setpoint as a factor of the stream 1 flow, measured by FT2.

In this case the valve FV2 in stream 1 is not used for control. It is adjusted manually by the operator.

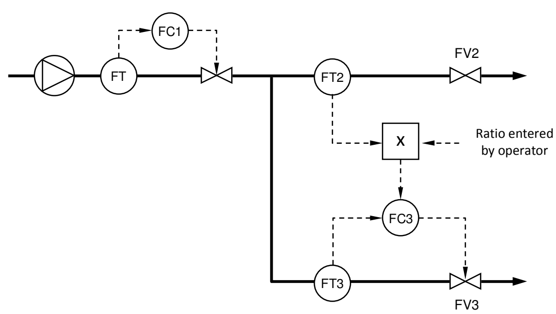

Fig. 2. Ratio control for flow-split.

However, operations staff may want to freely select which flow is the master flow. Traditional ratio control does not allow that. In Figure 2 FT2 is always master.

This could be solved by introducing selector functionality, where the operator can choose which structure to use. Implementation-wise, this is a fairly complicated solution, though.

Instead the control structure showed in Figure 3 solves the problem in an elegant way. Both valves are manipulated by the controller.

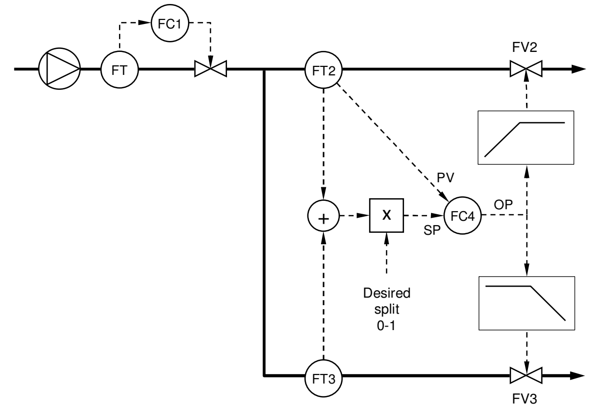

Fig. 3. Flexible flow split control.

The output (OP) of flow controller FC4 is sent to two tables, similar to a split range scheme. FC4 has action such that OP increases when CV is low.

The split is defined to be the share of the flow that goes to stream 1. E.g. if it is 0 all flow should go to stream 2. The desired split is entered by the operator.

When FC4.OP is between 0% and 50% valve FV3 is fully open, and when the output is between 50% and 100%, FV2 is fully open. This desired split is multiplied by the total flow to become the setpoint for FC4. The CV for FC4 is the flow in stream 1, FT2.

Note that at all times, one of the valves FV2 and FV3 is fully open. So the scheme also has the advantage of minimizing pressure loss, and thereby saving energy.

The scheme in Figure 3 was suggested by P. Sivertsson. I have not seen it in any publication. It has the advantage that it can be used in all operational scenarios. A disadvantage is that its inner workings may be hard to understand for users.

3. SYSTEMATIC ANALYSIS OF CONTROL STRUCTURES

Traditional expositions of classical control structures often lack a systematic and holistic perspective. The step from control specifications to choice of control structure is seldom obvious, and it is often unclear if the problem at hand could be solved by other structures than the one presented.

As a consequence it is not easy for an inexperienced user to design a new control structure that solves a given problem, or to combine several structures. In comparison, MPC design is definitely more systematic.

There are a few things that could be said about traditional structures, though. Table 1 classifies the structures in terms of the number of PVs, CVs etc for each structure.