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 inter-actions.

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.

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.

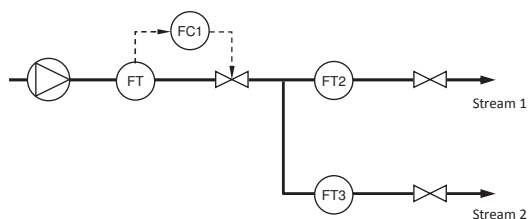To illustrate this point, consider a simple flow split process, as showed in Figure 1.

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.
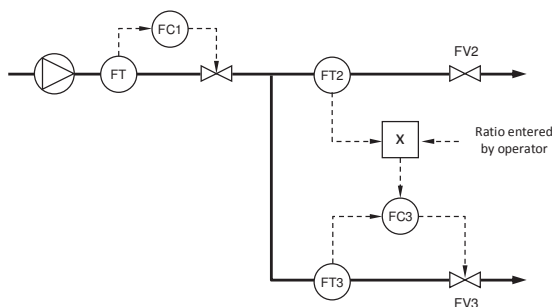

Fig. 2. Ratio control for flow-split.

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.
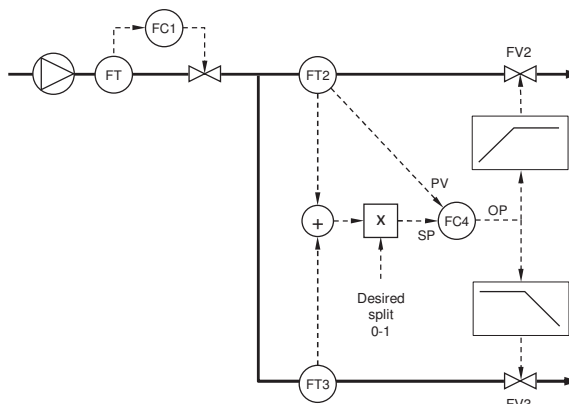

Fig. 3. Flexible flow split control.

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.

Structure	PVs	CVs	MVs	Controllers
Cascade control	2	2	1	2
Feedforward from external signal	2	1	1	1
Mid-ranging	1	2	2	2
Split-range	1	1	2	1
Selector control; Conditional control	2	2	1	2
Parallel controllers	1	1	2	2
Feedforward from setpoint	1	1	1	1
Ratio control, fixed ratio	2	2	2	1 or 2
Ratio in cascade	3	3	3	2 or 3

Table 1. Comparison of control structures.

Another way of structuring the understanding of control structures is by their functionality:

- Some structures are primarily used to improve the performance of a single-input single-output (SISO) control loop, e.g. cascade control and feedforward control.

- Other structures e.g. mid-ranging and ratio control are motivated by the presence of non-linearities. See the section below, about non-linearities, for a further discussion of this.

- It is also worth noticing that some structures rely on MV saturation to work at all. This is the case with conditional, or selector, control.

The "holy grail" in this area would be to have an algorithm that suggests a control structure or combination of structures, given a coarse process description, such as a process flow diagram, control specifications and a disturbance description.

It is very hard to find the Holy Grail, though. One step on the journey would be to have an algorithm for producing a block diagram description of a process given a flow sheet / P&I-diagram and some additional information.

A simple example could serve as an illustration (Fig 4). The process is just a gas pipeline where the pressure between the two valves is controlled by manipulating the first valve, and the flow through the second valve is controlled by manipulating that valve
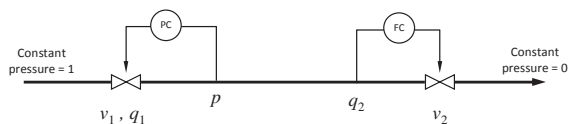

Fig. 4. Pressure controlled gas pipe line, with flow control.

The block diagram for this (controlled) process is showed in Fig 5. Here the process is represented by three blocks P_1, P_2, P_3, in order to clearly indicate dynamic interaction between the controlled variables
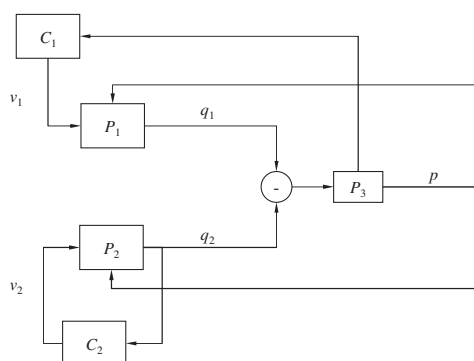

Fig. 5. Block diagram for pressure – flow control example.

The control scheme described above is completely decentralized. It may be quite hard to tune the individual controllers in this case. An example of an open loop step response (flow in manual – pressure in auto) is given in Fig 6. These observations suggest that a more advanced structure is called for. In this case, decoupling by a single feedforward makes the control tuning much easier.
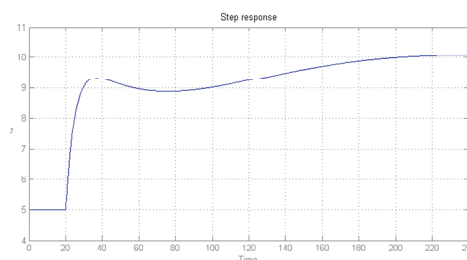

Fig. 6. Open step response for flow; pressure in auto.

More generally, the complexity of the dynamics of the "equivalent open process" Euzébio and Barros (2015), gives useful insight in the MV-CV-pairing problem for a multi-variable process.

The analysis of block diagrams is crucial in the study of control structures. Here we summarize some basic principles that are not new, but also not particularly well known.

The easiest way to determine all transfer functions in a block diagram is probably to write down one equation for each block, considering the internal variables (outputs of the blocks) as "unknowns", the scalar transfer functions of each block as "coefficients" and external inputs as "parameters". In this way we get a set of linear equations represented by a square matrix. The example below shows the equation for the textbook cascade control scheme. Here r_1 is the setpoint for the master controller, u_1 and y_1 are the master loop control signal and PV, and d_1 and d_2 disturbances entering at different points in the process.

$$x = \begin{bmatrix} u_1 \\ u_2 \\ y_2 \\ y_1 \end{bmatrix} \qquad x = \underbrace{\begin{bmatrix} 0 & 0 & 0 & -C_1 \\ C_2 & 0 & -C_2 & 0 \\ 0 & P_2 & 0 & 0 \\ 0 & 0 & P_1 & 0 \end{bmatrix}}_{F} x + \underbrace{\begin{bmatrix} C_1 \\ 0 \\ 0 \\ 0 \end{bmatrix}}_{b_1} r_1 + \underbrace{\begin{bmatrix} 0 \\ 0 \\ P_2 \\ 0 \end{bmatrix}}_{b_2} d_2 + \underbrace{\begin{bmatrix} 0 \\ 0 \\ 0 \\ P_1 \end{bmatrix}}_{b_3} d_1$$

The solution to this equation is of course

$$x = (I - F)^{-1} (b_1 r_1 + b_2 d_2 + b_3 d_1)$$

which gives us all the transfer functions. The adjugate matrix expression for the matrix inverse, tells us that the denominator for all transfer functions is $1/\det(I - F)$, except for possible cancellations. All these calculations are easy to make using some symbolic math software. These observations are captured, in quite a different formulation, by "Mason's rule" (Mason, 1956). It is worth noticing that the F-matrix is closely related to the adjacency matrix of the block diagram considered as a directed graph.

3. PLANT-WIDE CONTROL

Some issues in plant wide control could be considered as control structure selection. For example, in an inventory line such as the one in Fig. 7, we first have to choose a throughput manipulator (TPM). Then all levels should be controlled, so for every tank there is a corresponding MV. In the flow diagram the individual valves and flow meters are not indicated; instead each flow loop is symbolized by a circled FCx.
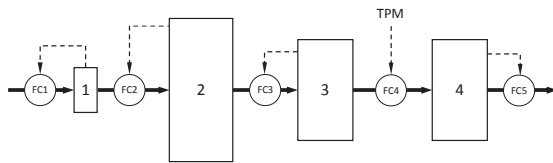

Fig. 7. Inventory line controls, following the "radiating rule."

Thus we get a special case of the MV-CV pairing problem. In most cases, the "radiating rule" as described in Aske and Skogestad (2009) gives a good PWC scheme.

However, there are cases where a non-locally consistent scheme gives advantages. The above scenario is taken from a Perstorp plant, and here the operator sometimes wants to give a constant setpoint to the first flow, FC1. When that happens we have in principle two "TPMs", and one of the levels is controlled manually. One can question if TPM is the right word here; maybe "independent flow" is more suitable.

As indicated in the figure, the first buffer level is very small compared to the second one. If we follow the radiation rule, the operator will have to manually control the level of a tank which is small in comparison to the flow. This is not a desirable situation. If instead we use the pairing showed in Fig. 8, the operator the operator gets an easier task.
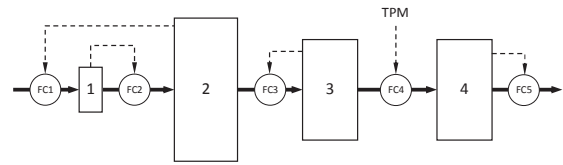

Fig. 8. Inventory line: non-radiating inventory control

He or she then keeps the level in tank 2 within limits by manually adjusting the SP of FC1, as indicated in Fig 9.
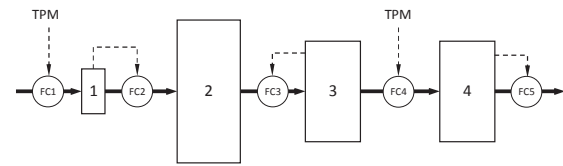

Fig. 9. Inventory line: non-radiating inventory control

There could be other advantages with a structure not obeying the radiating rule. E.g. the structure in Fig 8 will typically give less variability in FC2 than if we use the radiating rule (Fig 7).

An important question is whether the scheme in Fig 8 leads to problems in controller tuning. Some simple analysis shows that that scheme does not imply complicated tuning. It is easy to tune PI controllers for all levels, giving good performance.

There are many possible variable pairings in an inventory line, by pure combinatorics. Fig 10 shows a complicated scheme. It is unlikely that there are practical reasons for using this structure, but from a theoretical point of view we may note that all structures where the TPM is in one end of the line, it is possible to stabilize the whole line using properly tuned PI controllers.
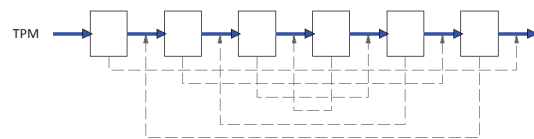

Fig. 10. Inventory line: Complex control structure.

The pairing in Fig 11 is even more unintuitive, but can be made to work. An observation that is a bit interesting is that this structure requires that one of the level controllers has derivative action in order for the whole system to be stable. (Proving this requires some fairly complicated calculations.)
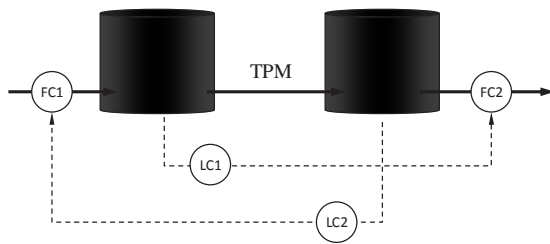
Fig. 11. Inventory control. Off-diagonal pairing.

Maybe this is a general principle for quantifying if a given structure is "appropriate" or not: if it is impossible to stabilize the entire process using only PI controllers, then that structure is not optimal. This statement assumes that the dynamics of the individual processes is reasonably simple.

In order to have a general method for control structure selection, and in particular MV-CV-pairing, we first need to have a method for analyzing the consistency of a suggested control scheme.

The question is: which flows can be set independently of the others, while ensuring that mass balances are maintained, i.e. that there is no accumulation or depletion of material in a single tank in steady state. This is a topic addressed in the classical theory of mass balances, e.g. Reklaitis (1983). From a classical control point of view it can be formulated in terms of internal stability of all states.

A solution to the problem is obtained by considering the matrix defining the topology, which may be considered as the incidence matrix of the corresponding graph. This matrix has one column per flow and one row per node (tank). The incidence matrix of the topology in Figure 12
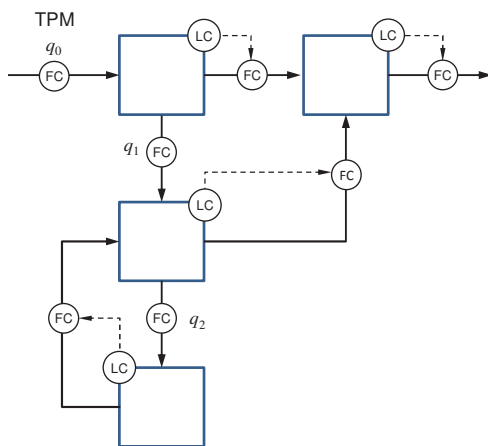

Fig. 12. TPM selection. Example of consistent structure.

is given by equation

$$x = \begin{bmatrix} 1 & -1 & 0 & -1 & 0 & 0 & 0 \\ 0 & 1 & -1 & 0 & 1 & 0 & 0 \\ 0 & 0 & 0 & 1 & -1 & 1 & -1 \\ 0 & 0 & 0 & 0 & 0 & -1 & 1 \end{bmatrix}$$

A configuration (choice of TPMs) is consistent if and only if it satisfies the following algebraic requirement: Suppose we assign the flows $k_1,...,k_n$ to be TPMs. Then that choice is consistent iff the matrix $\tilde{x}$ obtained by removing columns $k_1,...,k_n$ from x has maximum rank. Typically $\tilde{x}$ is a square matrix, so we can test consistency by calculating the determinant.

The motivation for the criterion is that the equation

$$\tilde{x}\tilde{q} = \begin{bmatrix} q_{k_1} \\ \vdots \\ q_{k_n} \end{bmatrix}$$

has a unique solution iff $\tilde{x}$ is non-singular (non rank-deficient). $\tilde{q}$ is the vector of remaining flows.

Using this observation we can also enumerate all consistent control schemes, using exhaustive search.

4. NON-LINEARITIES

Many of the classical control structures actually rely on non-linearities that may not be represented in the block diagrams or P&I diagrams. One such example is parallel control, as showed in Figure 13. In this application the idea is that the vent valve should normally be completely closed and only open when there is a large excess pressure.

This is achieved by having two controllers operating on the same process value, manipulating two different valves, and having distinct setpoints. PC2 in the figure is the "rescue controller" that is only active when venting is needed. It has a higher setpoint than PC1, which is the controller maintaining the pressure in the header when the plant is running normally.
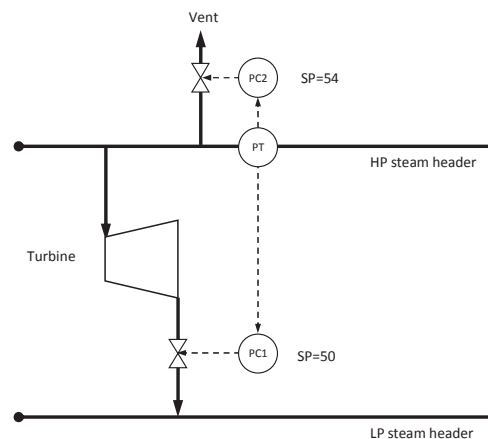
Fig. 13. Application of parallel control: steam pressure in high pressure header.

In some cases the whole motivation for using a certain control structure lies in the process non-linearities. For example "mid-ranging" (a.k.a. valve position control, input reset control, or habituating control) is often used when there are two MVs, one CV, and one of the MVs has a faster but less powerful effect on the CV. Unlike in parallel control, the two MVs are supposed to be used simultaneously in the normal case.

The purpose of the control scheme, showed in Figure 14, is to ensure that the "small valve" is kept away from its saturation limits so that it can be used as the manipulator for the CV. Using the "large, big" valve as a CV-manipulator would typically render inferior control performance, both because of the dynamics and quantization effects, such as valve stiction. A linear model of course does not explain any of that.
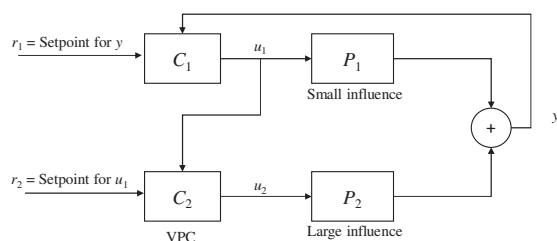
Fig. 14. Block diagram representation of valve position control.

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REFERENCES

E.M.B. Aske and S. Skogestad: *Consistent Inventory Control.* Ind. Eng. Chem. Res. Vol 48, pp 10892–10902. 2009.

T.A.M. Euzébio and P.R. Barros: *Iterative Procedure for Tuning Decentralized PID Controllers.* IFAC Papers OnLine 48-8, pp 1180-1185. 2015

T.E. Marlin: *Process Control.* 2nd edition. McGraw Hill, 2000.

S.J. Mason: *Feedback Theory – Further Properties of Signal Flow Graphs.* Proc. Institute of Radio Engineers, pp 920–926. July, 1956.

G.P. Rangaiah, and V. Kariwala (eds): *Plant-wide Control: Recent Developments and Applications.* Wiley, 2012.

G.V. Reklaitis: *Introduction to Material and Energy Balances.* Wiley, 1983.

S. Skogestad and I. Postlethwaite: *Multivariable Feedback Control: Analysis and Design.* 2nd edition. Wiley, 2005.

C.A. Smith and A. Corripio: *Principles and Practice of Automatic Process Control.* Third Edition. Wiley, 2006