Control Structure Selection for an Evaporation Process

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A systematic procedure for control structure selection is applied to the evaporation process of Newell and Lee (1989). First, promising sets of controlled variables are selected, based on steady-state economic criteria. The objective is to find sets of controlled variables which with constant setpoints keep the process close to the economic optimum ("self-optimizing control") in face of disturbances and implementation errors. Second, stabilization and controllability analysis is performed for the most promising sets of economic controlled variables.

1. Introduction

Control structure selection consists of selecting controlled variables, manipulated variables, measurements and links between them. A poor choice can give both dynamic and steady-state problems, such as instability, input saturation, operation outside constraints and non-optimal operation. This can be partly counteracted by using logic, model predictive control and on-line optimization, but the control system then becomes more complicated and costly than necessary. Selecting a good control structure is a precondition for getting a simple control system with good control behavior.

Inspired by the many methods presented in the area of control structure selection, both from a steady-state economic point of view and from a dynamic analysis point of view, a procedure is proposed and applied to a simple, but illustrating example of the evaporation process of Newell and Lee (1989). The control structure selection is mainly based on a plant-wide control procedure presented by Larsson and Skogestad (2001). Back-off, Perkins (1998), from the nominal operation is sometimes needed in order to obtain feasible operation. We define back-off (or setpoint adjustment) as the difference between the nominally optimal setpoint and the actual setpoint, used e.g. to achieve feasible operation when there are disturbances. There are two cases: (a) Control of variables at active output constraints. The required back-off is given by the implementation (including measurement) error. (b) Control of unconstrained variables. We adjust the setpoints to achieve feasibility when there are disturbances which otherwise move the active constraints or include a new one. Note that the required back-off and corresponding economic loss depends on the selected controlled variables. Thus, the primary issue is to select the right control structure (variables), whereas the back-off is just a setpoint *adjustment* to deal with nonlinearities and in particular constraints. Optimal back-off can be determined by

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robust optimization, which minimizes the nominal steady-state economic criteria, given that the constraints are satisfied for all expected disturbances and implementation errors, see Glemmestad et al. (1999).

2. Control Structure Selection Procedure

A short summary of methods and rules recommended in control structure selection, Larsson and Skogestad (2001):
1) Formulate a steady-state economic objective. 2) Define control objectives, steady-state and dynamic degrees of freedom, manipulated variables with allowed variations, disturbances with expected variations and possible controlled variables with expected implementation errors. 3) Steady-state optimization should be performed at the nominal point and, if possible, for extreme values of expected disturbances and implementation errors. 4) In order to reduce the number of sets of controlled variables to be evaluated in detail, use active constraint control, see Maarleveld and Rijnsdorp (1970), and the minimum singular value rule, see Skogestad (2000). 5) For the remaining combinations of controlled variables the economic loss imposed by keeping the sets of controlled variables constant (rather than at their optimal values) should be evaluated for expected disturbances and implementation errors. 6) To avoid infeasibility, try back-off from nominally optimal setpoints. 7) For stabilization select the variables to control and manipulate based on computing the poles and their associated input and output directions, see Havre (1998). 8) Perform controllability analysis for the most promising sets of economic controlled variables. Tools are provided by Skogestad and Postlethwaite (1996). 9) Design controllers and run nonlinear simulations for the most promising alternatives.

3. Evaporation Process Case Study

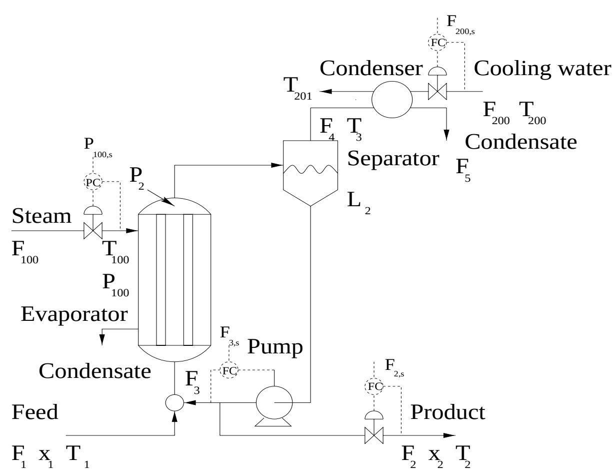

Figure 1. Evaporation process

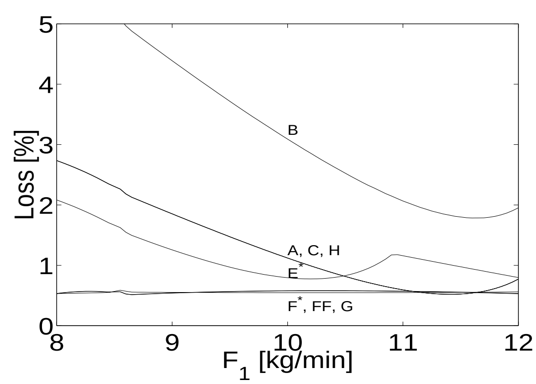

Figure 2. Loss as function of feed flowrate for different alternatives with robust setpoints and simple logic

In the evaporation process of Newell and Lee (1989) the concentration of dilute liquor is increased by a vertical heat exchanger with recirculated liquor, see figure 1. The steady-state economic objective is to minimize the operational cost ($\$/h$) related to steam, cooling water and pump work, see Wang and Cameron (1994):

$$J = 600F_{100} + 0.6F_{200} + 1.009(F_2 + F_3) \tag{1}$$

Process constraints related to product specification, safety and design must be met:

$$x_2 \geq 35\% \tag{2}$$
$$40\text{kPa} \leq P_2 \leq 80\text{kPa} \tag{3}$$
$$P_{100} \leq 400\text{kPa} \tag{4}$$
$$F_{200} \leq 400\text{kg/min} \tag{5}$$
$$0\text{kg/min} \leq F_3 \leq 100\text{kg/min} \tag{6}$$

There are four manipulated variables; steam pressure, cooling water flowrate, recirculating flowrate and product flowrate:

$$u^T = [F_{200}\ P_{100}\ F_3\ F_2] \tag{7}$$

One degree of freedom is purely dynamic (the condenser level which needs to be stabilized), hence there are three steady-state degrees of freedom. The major disturbances are feed flowrate, feed concentration, feed temperature and cooling water inlet temperature, with expected variations about $\pm 20\%$:

$$d^T = [F_1\ x_1\ T_1\ T_{200}] = [10 \pm 2 \text{ kg/min} \quad 5 \pm 1\% \quad 40 \pm 8^{\text{o}}\text{C} \quad 25 \pm 5^{o}\text{C}]. \tag{8}$$

Controlled variable candidates are all possible measurements and manipulated variables:

$$y^T = [F_2\ F_3\ F_4\ F_5\ X_2\ T_2\ T_3\ L_2\ P_2\ F_{100}\ P_{100}\ Q_{100}\ F_{200}\ T_{201}\ Q_{200}] \tag{9}$$

Expected implementation error associated with each variable is based on the following rules: Flowrates ($\pm 10\%$), compositions ($\pm 1\%(abs)$), temperature ($\pm 1^{o}C$) and pressure ($\pm 2.5\%$). The model equations are given in appendix A.

3.1. Steady-state economic evaluation and optimization

The steady-state optimal values at the nominal point for the objective function is $6162\$/h$ and corresponding to the following optimal values:

$$y^T_{opt} = [1.4\ 27.7\ 8.6\ 8.6\ 35.0\ 90.9\ 83.4\ 1.0\ 56.2\ 10.0\ 400.0\ 365.6\ 230.2\ 45.5\ 330.0] \tag{10}$$

Steady-state optimization at the nominal point and for extreme values of the disturbances yield that two of the constraints, product composition (x_2) and steam pressure(P_{100}), are always active. Active constraint control then consumes two steady-state degrees of freedom. The last degree of freedom is optimally unconstrained for most disturbances (There is one exception, for low feed flowrate the last degree of freedom is consumed by the minimum operating pressure constraint). The best economic choice for this controlled variable is related to the self-optimizing control properties. There are 13 candidate variables. The minimum singular value rule, see Skogestad (2000), is applied to eliminate some of them. For one single input the rule is to select the controlled variable with the largest absolute process gain ($|G|$), when the variables are scaled with respect to the sum of the variation in their optimal value and the expected implementation error. The seven most promising combinations of economic controlled variables are listed in table 1. Here we have also included a feed-forward improvement of alternative F (denoted FF). In addition we study alternative E, proposed and used by Newell and Lee (1989). In this alternative the recirculating flowrate (F_3) is not used as a manipulated variable in the basic control layer. They do not use active constraint control, because the last available manipulated variable, cooling water outlet flow (F_{200}), has too small effect on the product composition (x_2). The steam pressure (P_{100}) is not kept on its constraints, but is used to control the product composition (x_2). The cooling water flowrate (F_{200}) is used to control the operating pressure (P_2).

Table 1
Most promising alternative sets of controlled variables based on the minimum singular value rule

Rank	Alt.	$y_{s,1}$	$y_{s,2}$	$y_{s,3}$	$\|G\|$
1	G	x_2	P_{100}	T_{201}	0.0150
2	FF	x_2	P_{100}	F_{200}/F_1	0.0135
3	F	x_2	P_{100}	F_{200}	0.0108
4	C	x_2	P_{100}	P_2	0.0044
5	A	x_2	P_{100}	T_2	0.0042
6	H	x_2	P_{100}	T_3	0.0042
7	B	x_2	P_{100}	F_3	0.0018
–	E	x_2	P_2	F_3	–

Table 2
Average economic loss with optimal back-off to avoid infeasibility

Rank	Alt.	$y_{s,1}$ (x_2)	$y_{s,2}$ (P_{100})	$y_{s,3}$	Average loss
1	F^*	36	390.0	231.2	0.55%
2	G	36	390.0	48.0	0.59%
3	FF	36	390.0	20.6	0.60%
4	E^*	36	57.3	41.0	0.91%
5	H	36	390.0	90.3	1.24%
6	C	36	390.0	69.6	1.24%
7	A	36	390.0	98.7	1.24%
8	B	36	181.0	134.4	3.24%
–	F	–	–	–	Infeasible
–	E	–	–	–	Infeasible

*=With use of logic

To achieve feasibility, we compute the optimal back-off from the nominal optimum setpoints by robust optimization(Glemmestad et al. (1999)). The losses related to keeping the operation at these robust setpoints are given in table 2 and in figure 2. There exists no feasible setpoint adjustment (back-off) for alternatives E and F. Alternative F has feasibility problems only at low feed flowrates when the operating pressure (P_2) gets too low. To avoid infeasibility a simple logic scheme can be introduced: If the feed flowrate becomes lower than 8.5 kg/min, the cooling water flowrate is reduced to avoid that the operating pressure becomes too low. For the proposed structure of Newell and Lee (1989) (alternative E) several constraints may be violated, and the logic becomes more complicated. For large feed flowrates and a high inlet cooling water temperature the cooling water flowrate should be kept constant instead of the operating pressure (P_2).

3.2. Stabilization and controllability analysis

The rest of the analysis is based on a scaled linear, dynamic model. The holdup in the separator must be stabilized. Based on computing the poles and their associated input and output directions we find as expected, that the separator level must be controlled by

Table 3
Control structure used in decentralized control

Alt.	*Loop* 1	*Loop* 2	*Loop* 3	*Loop* 4**
E^*	$x_2 \leftrightarrow P_{100}$	$P_2 \leftrightarrow F_{200}$	F_3	$L_2 \leftrightarrow F_2$
F^*	$x_2 \leftrightarrow F_3$	P_{100}	F_{200}	$L_2 \leftrightarrow F_2$
FF	$x_2 \leftrightarrow F_3$	P_{100}	F_{200}/F_1	$L_2 \leftrightarrow F_2$
G	$x_2 \leftrightarrow F_3$	P_{100}	$T_{201} \leftrightarrow F_{200}$	$L_2 \leftrightarrow F_2$

* = With use of logic, ** = Stabilizing loop

the product flowrate. Controllability analysis is performed for the three most economic promising alternatives F, FF and G, plus alternative E. The alternatives are largely controllable. The relative gain array (RGA) is used to select links between the controlled and manipulated variables, see table 3. Alternative FF and G are shown in figure 3 and 4. Decentralized controllers were designed, and nonlinear simulations were performed to

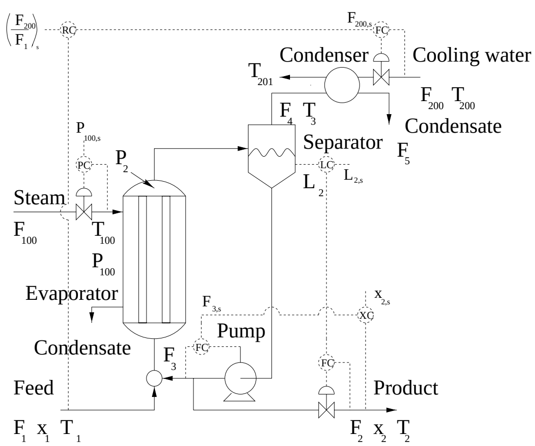

Figure 3. Evaporation process with control structure FF

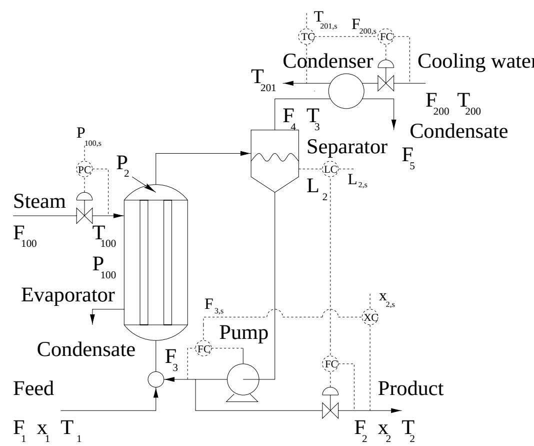

Figure 4. Evaporation process with control structure G

verify the controllability of the designs.

4. Conclusion

A systematic procedure for control structure selection, based on both steady-state economic evaluation and controllability analysis, has been demonstrated. The evaporation process example has been used to illustrate how important the selection of control structure is and how rewarding it may be. By putting serious effort in selecting a good control structure, it is possible to avoid significant control problems and reduce the complexity of the control system. Both alternative FF and G have small economic loss, good control behavior and no need of logic. In the end alternative G is proposed, because it is based on pure feedback control.

A. Model equations

$$20dL/dt = F_1 - F_4 - F_2 \quad (11)$$
$$20dx_2/dt = F_1x_1 - F_2x_2 \quad (12)$$
$$4dP_2/dt = F_4 - F_5 \quad (13)$$
$$T_2 = 0.5616P_2 + 0.3126x_2 + 48.43 \quad (14)$$
$$T_3 = 0.507P_2 + 55.0 \quad (15)$$
$$T_{100} = 0.1538P_{100} + 90.0 \quad (16)$$
$$Q_{100} = 0.16(F_1 + F_3)(T_{100} - T_2) \quad (17)$$
$$Q_{100} = 36.6F_{100} \quad (18)$$
$$Q_{100} = 38.5F_4 + 0.07F_1(T_2 - T_1) \quad (19)$$
$$Q_{200} = 6.84(T_3 - 0.5(T_{200} + T_{201})) \quad (20)$$
$$Q_{200} = 38.5F_5 \quad (21)$$
$$Q_{200} = 0.07F_{200}(T_{201} - T_{200}) \quad (22)$$

Newell and Lee (1989) assumed the time lag connected to the slave controllers to be 1.2 min. This seems too large, and we use 0.1 min.

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