CONTROL CONFIGURATION SELECTION FOR DISTILLATION COLUMNS UNDER TEMPERATURE CONTROL

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1 Introduction

In most process control applications, the structural issues which precede the actual controller design are the most important. The problem of *control structure selection* involves the following decisions:

1. Selection of control objectives, actuators and measurements.

2. "Control configuration selection": Selection of controller structure (e.g., pairing of actuators for decentralized control.

Note that we define the last step as the control configuration selection, whereas the combination of the two steps is denoted the "control structure selection".

In practice, control systems are implemented in a hierarchical manner, with a regulatory control system at the lowest level. The two main objectives for the regulatory control system are

1. Take care of control tasks where fast response is needed.

2. Make the control problem seen from the levels above simple.

The higher levels in the control system may include a supervisory and optimizing control system or simply the operator. In any case, the issue of control structure selection is usually most important for regulatory control. This is because the main control objective at this level is to facilitate good operation, that is, to implement a simple control system that makes it easy for the operators to operate the plant. Thus, the control objectives are not clearly defined at this level, and since the control system should be simple we generally want to implement decentralized SISO controllers.

Specifically, one often has extra measurements which are not particularly important to the control of the plant from an overall point of view. However, at the regulatory control level one often uses these variables as *secondary control objectives* by closing local loops. Typically, such variables may include selected temperatures and pressures. The setpoints for these loops may be adjusted from the higher levels giving rise to a *cascaded control system*. Effectively, by closing secondary control loops, we replace the original independent variables (typically, flows and valve positions in process control applications) by some new independent variables (the setpoints for the secondary control variables). The idea is then that the control problem in terms of these new independent variables is simpler, and at least that they need not be adjusted so frequently, that is, the "fast control" is taken care of by the secondary control loops implemented at the regulatory level.

A block diagram is shown in Fig.1. Here u_2 represents the original independent variables which are used to control the secondary ("extra") outputs, y_x. The setpoints for the secondary loops y_{xs} then become the new control variables, $u_x = y_{xs}$.

In most cases it is desirable to have the secondary loops as fast as possible. Thus, when the operator or higher levels in the control system change u_x, this results in an almost immediate change in y_x, i.e., $y_x \approx u_x$. Also, in this case the tuning of the secondary loops does not matter for the overall system (provided they are sufficiently fast).

In this paper we use distillation column control as an application. The main control problem here is the strong coupling between the two loops as indicated, for example, by the large RGA-values. In the paper, we study how the use of temperature cascades, in addition to improving the operation, may help reduce this interaction.

2 Distillation control

Control of distillation columns is a challenging problem due to strong interactions, ill-conditionedness, nonlinearity and the large number of possible control structures. A simple distillation column (Fig. 2) may, from a control point of view, be considered a 5×5 problem with L, V, D, B and V_T as the manipulated inputs (actuators) and x_D, x_B (product quality), M_D, M_B (levels) and p (pressure) as the controlled outputs (control objectives).

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Figure 1: Block diagram with secondary loop closed.

Typical disturbances (d) include feed composition (z_F), feed flowrate (F) and feed enthalpy (q_F).

In practice, distillation columns are usually controlled in a hierarchical manner with the three loops for level and pressure control implemented at the regulatory control level. The "conventional" distillation control configuration selection problem, which addresses which of the five inputs should be used for control in these three loops, has been discussed by a number of authors (e.g., Shinskey, 1984, Skogestad et al., 1990). By convention, the resulting configuration is named by the two independent variables which are left for composition (quality) control, for example, the LV-configuration uses reflux L and boilup V for composition control. In this paper only the LV-configuration is considered.

The quality control is often implemented at some higher level or left for manual control by the operators. However, this approach has several problems:

- Unless very fast control is used, the use of $u_1 = L$ and $u_2 = V$ to control the $y_1 = x_D$ and $y_2 = x_B$ yields a very difficult control problem with strong interactions and large RGA-values.

- There is often a long delay associated with measuring the product compositions which makes fast control impossible.

- There is a need to close at least one loop with relatively fast control in order to "stabilize" the compositions in the distillation column, which otherwise behave almost as a pure "integrator".

Figure 2: Typical distillation column using LV-configuration.

To deal with at least the last problem one often implements a secondary temperature loop at the regulatory control level (e.g., Kister, 1991). This loop makes it possible for the operators to operate the column when the composition loops are not closed.

Remark. An alternative approach is to use multiple temperature measurements along the column to estimate the compositions (e.g., Mejdell and Skogestad, 1991ab). This avoids the measurement delay and makes it easier to have fast control. However, even in this case one may for operational reasons want to close one temperature loop at the regulatory control level as described above.

3 Closing secondary loops

General results. From Fig.1 we have with the secondary loops open ($C_x = 0$)

$$y = G_1 u_1 + G_2 u_2 + G_d d \tag{1}$$

(Note that we may have assumed that some regulatory loops, e.g., the pressure and level loops for distillation columns, have been closed). Similarly, with $C_x = 0$, the model for the secondary output is

$$y_x = G_{x1}u_1 + G_{x2}u_2 + G_{xd}d \tag{2}$$

Closing the secondary loops, effectively means that we replace the inputs u_2 by the setpoints $u_x = y_{xs}$, and for the cascaded system, we get,

$$y = \tilde{G}_1 u_1 + \tilde{G}_2 u_x + \tilde{G}_d d \tag{3}$$

where

$$\tilde{G}_1 = G_1 - G_2 C_x (I + G_{x2}C_x)^{-1} G_{x1} \tag{4}$$
$$\tilde{G}_2 = G_2 C_x (I + G_{x2}C_x)^{-1} \tag{5}$$
$$\tilde{G}_d = G_d - G_2 C_x (I + G_{x2}C_x)^{-1} G_{xd} \tag{6}$$

In most cases we use decentralized control for the cascade loops and C_x is a diagonal matrix. The use of the cascade clearly changes the "effective" plant as seen from the disturbances and inputs. Specifically, if the cascade loops are slow ($C_x \to 0$) we have

$$\tilde{G}_1 = G_1, \quad \tilde{G}_2 = G_2 C_x, \quad \tilde{G}_d = G_d \tag{7}$$

and as expected the system behaves as without the cascade except that the inputs u_2 are scaled by C_x. At the other extreme, tight control of the secondary variables ($C_x \to \infty$) yields $G_2 C_x (I + G_{x2}C_x)^{-1} \approx G_2 G_{x2}^{-1}$ and

$$\begin{aligned} \tilde{G}_1 &= G_1 - G_2 G_{x2}^{-1} G_{x1}, \\ \tilde{G}_2 &= G_2 G_{x2}^{-1}, \\ \tilde{G}_d &= G_d - G_2 G_{x2}^{-1} G_{xd} \end{aligned} \tag{8}$$

The changes in control properties resulting from implementing the secondary loops may be analyzed by use of a number of standard measures for linear controllability evaluation, such as RHP-zeros, RGA-analysis for interactions, disturbance sensitivity and sensitivity to model uncertainty.

In this paper we mainly use the relative gain array (RGA or Λ) to look at interaction in the distillation column with an added temperature control loop. The properties of the RGA are well known (e.g., Grosdidier et al., 1985). The most important for our purpose are: 1) No two-way interaction is present when $\Lambda = I$, 2) The RGA is independent of scaling in inputs or outputs, and 3) The rows and columns both sum up to 1. For 2×2 systems the RGA is especially easy to compute; because of the third property mentioned, we only have to compute the $(1,1)$ element of the RGA which is given by $\lambda_{11} = 1/(1-Y)$, $Y = \frac{g_{12}g_{21}}{g_{11}g_{22}}$. To evaluate the disturbance sensitivity, we consider the closed-loop disturbance gain (CLDG) which is the appropriate measure when we use decentralized control (Hovd and Skogestad, 1992). The CLDG is defined as $G_{diag}G^{-1}G_d$, where G_{diag} consists of the diagonal elements of G.

Although the main part of the analysis is based on the RGA, we also provide detailed controller designs and simulations to confirm the predictions.

Temperature cascade for distillation column. We here consider composition control by manipulating the reflux L and boilup V (LV-configuration), but the following development also applies to other control configurations as well.

We want to use an internal tray temperature measurement for cascade control. For a binary mixture with constant pressure there is a direct relationship between tray temperature (T) and composition (x). In terms of deviation variables we then have $T_i = K_{Tx}x_i$, where for ideal mixtures K_{Tx} is approximately equal to the difference in pure component boiling points. The open-loop model for the LV-configuration may be written,

$$\begin{pmatrix} x_D \\ x_B \\ x_i \end{pmatrix} = \begin{pmatrix} g_{11} & g_{12} \\ g_{21} & g_{22} \\ g_{x1} & g_{x2} \end{pmatrix} \begin{pmatrix} L \\ V \end{pmatrix} \tag{9}$$

where x_D, x_B and x_i are the compositions in the top, bottom and i^{th} tray of the distillation column.

We now implement a SISO controller from the temperature T_i to the reflux L: $L = c_x(T_s - T_i)$ (we could have used boilup instead). Here T_s is the setpoint for the temperature loop which becomes the new manipulated variable instead of L. In terms of the general problem discussed above this corresponds to selecting $u_1 = V$, $u_2 = L$, $u_x = T_s$, $y_x = T_i$ and $y = [x_B \; x_D]^T$. We can now write the linear equations relating the top and bottom compositions to the new set of manipulated variables as

$$\begin{pmatrix} x_D \\ x_B \end{pmatrix} = \tilde{G} \begin{pmatrix} T_s \\ V \end{pmatrix}, \tag{10}$$

$$\tilde{G} = \begin{pmatrix} \frac{g_{11}c_x}{1+g_{x1}c_xK_{Tx}} & g_{12} - \frac{g_{11}g_{x2}c_xK_{Tx}}{1+g_{x1}c_xK_{Tx}} \\ \frac{g_{21}c_x}{1+g_{x1}c_xK_{Tx}} & g_{22} - \frac{g_{21}g_{x2}c_xK_{Tx}}{1+g_{x1}c_xK_{Tx}} \end{pmatrix}$$

The RGA for $\tilde{G}$ can now be computed to study the interaction properties of the column for different temperature loop gains c_x,

$$\lambda_{11}(\tilde{G}) = \left(1 - \frac{g_{21}g_{12} - \frac{g_{21}g_{11}g_{x2}c_xK_{Tx}}{1+g_{x1}c_xK_{Tx}}}{g_{11}g_{22} - \frac{g_{21}g_{11}g_{x2}c_xK_{Tx}}{1+g_{x1}c_xK_{Tx}}}\right)^{-1} \tag{11}$$

We have the two limiting cases,

$$c_x = 0: \quad \lambda_{11}(\tilde{G}) = \left(1 - \frac{g_{12}g_{21}}{g_{11}g_{22}}\right)^{-1} = \lambda_{11}(G) \tag{12}$$

$$c_x = \infty: \quad \lambda_{11}(\tilde{G}) = \left(1 - \frac{g_{21}(g_{12}g_{x1} - g_{11}g_{x2})}{g_{11}(g_{22}g_{x1} - g_{21}g_{x2})}\right)^{-1} \tag{13}$$

As expected, with sufficiently slow temperature cascade controllers the RGA is unchanged.

Ideally, we would like no two-way interaction. Setting $\lambda_{11} = 1.0$ and solving Eq. 11 for c_x yields the following "optimal" feedback controller,

$$c_x^* = K_{Tx}^{-1} \left(\frac{g_{12}}{g_{11}g_{x2} - g_{12}g_{x1}} \right) \tag{14}$$

The "optimal" loop transfer function for the temperature loop is then given by,

$$L^* = c_x^* K_{Tx} g_{x1} = -(1 - \frac{g_{11} g_{x2}}{g_{12} g_{x1}})^{-1} = \lambda_{11}(G^s) - 1 \quad (15)$$

where

$$G^s = \begin{pmatrix} g_{11} & g_{12} \\ g_{x1} & g_{x2} \end{pmatrix} \quad (16)$$

Thus the optimal loop gain is essentially equal to the RGA involving x_D and x_i as outputs. We note that when x_D and x_i are strongly coupled (in terms of the RGA) then the loop gain should be large. Also, the bandwidth of the cascade loop should be approximately equal to the frequency where this RGA approaches 1. Since for the LV-configuration the shapes of the open-loop gains (e.g. g_{x1}) and the RGA as a function of frequency are similar (they break off at the dominant time constant, e.g., see Skogestad et al. (1990)), it seems that a simple proportional controller should be close to the optimal choice. Thus, in the example below we will only consider the steady-state value of $\lambda_{11}(\tilde{G})$ and assume that c_x is a P controller.

4 Distillation example

We consider as an example the high-purity binary distillation column studied by Skogestad and Mejdell (1991a). The basic data are given below:

#Trays	x_D	$1 - x_B$	z_F	L/F	$M_i/F[min]$
41	.99	.99	.5	2.71	0.5

We use a 82nd order model which includes liquid flow dynamics, and the resulting liquid lag from the top to the bottom of the column is about $\theta_L = 1.5$ min. The steady-state RGA-value of the model is $\lambda_{11}(G) = 35.5$ and approaches 1 at frequency $1/\theta_L$ (also see fig. 4 with $K_c = 0$).

Selection of tray for temperature sensor. There are several effects that must be taken into account when choosing where to install the temperature sensor: 1) When using a secondary loop involving reflux as the input, it should be placed in the top part of the column to minimize the process delay due to the liquid flow dynamics. 2) The interaction properties as expressed by the optimal loop gain in eq.15 will depend on the location. When reflux is used for the secondary temperature loop, placing the temperature close to the top will lead to an infinite loop gain and it will drop down to a value of $\lambda_{11}(G) - 1 = 34.5$ with the measurement located at the bottom of the column. 3) The temperature measurement should be sensitive such that it may be distinguished from noise (this consideration is probably the most important).

Figure 3 depicts different column temperature profiles as a function of operating conditions. To get high sensitivity (point 3 above) we have chosen to control the temperature at tray 34 (tray 8 counted from the top) for the remaining analysis.

Figure 3: Steady-state column temperature profile at different operating points. (Bottom is Tray 1)

The model in Eq.9 then becomes, at steady-state,

$$\begin{pmatrix} x_D \\ x_B \\ x_8 \end{pmatrix} = \begin{pmatrix} .8754 & -.8618 \\ 1.0846 & -1.0982 \\ 6.3912 & -6.3051 \end{pmatrix} \begin{pmatrix} L \\ V \end{pmatrix} \quad (17)$$

We get $\lambda_{11}(\tilde{G}^s(0)) = 477.9$, and since for our example $K_{Tx} = -13.5$, (15) tells us we will obtain no two-way interaction at steady-state ($\lambda_{11}(\tilde{G}(0)) = 1$) with a P-controller with gain $K_c = -5.53$. Frequency-dependent RGA-plots for the column, $\lambda(\tilde{G}(j\omega))$, with various gains for the temperature cascade are shown in figure 4. We note that with $K_c = -5.53$ the RGA is close to 1.0 at most frequencies (and not only at steady-state), confirming that a simple P-controller is close to the optimal.

The loop gain $L = K_c K_{Tx} g_{x1}$ for the cascade loop with the "optimal" controller gain $K_c = 5.53$ is shown in Fig.5. The loop gain crosses 1 in magnitude at frequency $w_c = 3.0 rad/min$, which is the approximate bandwidth of that loop. Due to valve dynamics, measurement dynamics, a liquid lag of about 0.3 min from the top to tray 8, etc. it seems that the closed-loop bandwidth must be about 1 rad/min or less. Thus in practice, the controller gain

Figure 4: Effect on the frequency-dependent RGA, $\lambda_{11}(\tilde{G}(j\omega))$, of varying gain in secondary loop, $c_x = K_c$.

Figure 5: Loop gain for secondary temperature loop with $K_c = 5.53$.

should be reduced by a factor of about 3.0, and we will use a controller gain $K_c = 1.84$ in the following.[1] This will not have much effect on the "decoupling effect" of the secondary loop, as we note from Fig. 4 that the RGA-plot is rather insensitive to the value of K_c.

No RHP-zeros are obtained for the resulting "open-loop" system $\tilde{G}(s)$ for any value of K_c.

The "open-loop" disturbance-rejection properties are also improved through use of the temperature cascade. This is seen from Figure 6 which shows the Closed-Loop Disturbance Gain, CLDG $= G_{diag} G^{-1} G_d$, as a function the secondary controller gain, K_c, for the most difficult disturbance (effect of F on x_B). Note that in this plot the outputs have been scaled such that an output of magnitude 1 corresponds to 0.01 mole fractions units. Similar results are obtained from Fig. 7 which shows a simulation of a step change in the same disturbance. We thus find that closing the secondary loop strongly reduces the sensitivity of the bottom composition to disturbances. The reason is of course that the compositions inside the column are strongly coupled, and fixing the composition at one point[2] results in small changes also at other locations. This is important because there is then less need to use fast control in the primary composition loops, and fast control in the primary loops are often impossible because of long measurement delays.

Closed loop simulations with also the two primary composition loops closed are shown in figure 8. A measurement delay of 6 minutes for the compositions is used in both loops. For the original plant without the secondary temperature loop (dotted lines) we use PID tunings from Skogestad and Lundstrøm (1990):

Loop	K	τ_I	τ_D
x_D	0.14	16.6	3.17
x_B	0.12	14.3	3.54

(18)

For the case with a secondary temperature loop (solid lines) we use PID tunings based on the Ziegler-Nichols tuning rule but with the proportional gain reduced by a factor two:

Loop	K	τ_I	τ_D
x_D	3.71	6.28	1.57
x_B	0.56	6.98	1.75

(19)

[1] Alternatively, we might have introduced dynamics into c_x to avoid instability. For example, since with infinite gain the RGA at steady-state is 0.925, which is very close to 1, we might have used a PI controller.

[2] For a binary separation, temperature is a direct measure of composition.

Figure 6: Improved disturbance rejection with temperature cascade. Plot shows CLDG for effect of disturbance in feed flow on x_B.

Figure 7: Improved disturbance rejection with temperature cascade. Plot shows response in x_B to disturbance in F.

We see from these simulations that the secondary temperature loop provides for much better control of the top composition, x_D, with somewhat less improvement for the bottom composition, x_B. This is as expected since the temperature sensor is located towards the top (stage 8 from the top) and its setpoint is determined by the x_D-controller. In effect we have achieved a one-way decoupling: With $u_1 = V$ and $u_{2x} = T_s$ as the new inputs, we find that u_1 has an effect on $y_1 = x_B$, but very little effect on $y_2 = y_D$, whereas u_{2x} has an effect on y_1 and somewhat less effect on y_2.

Including the secondary loop gives larger input signals (changes in reflux and boilup) than for the original plant, but without being close to violating the constraints ($L = 0$, $V = 0$).

5 Discussion

1. We note from the simulations that the feed flow disturbance has a rather large effect on x_B even with the secondary loop closed. The reason is that it takes some time before the temperature sensor near the top sees this disturbance. To improve this response the temperature sensor should be located in the bottom part of the column. However, in this case the top composition would become more sensitive to disturbances. The obvious conclusion is to place the temperature sensor in the top part (and close this loop using L) if the top composition is most critical, or place it in the bottom (and close this loop with V) if the bottom composition is most critical.

2. If large variations in the operating point of the column is expected one may choose to use the weighted average of several tray temperatures for the temperature measurement. This will avoid the problem of an insensitive measurement if the temperature profile becomes flat at the selected tray location. The outer cascade which contains integral action will in any case reset the setpoint of the average temperature to its correct value.

3. The use of "weighted average temperature" is indeed very similar to the static composition estimator of Mejdell. But, as noted before, even with such an estimator, it may be a good idea to implement an independent inner temperature cascade.

4. The reason why the temperature cascade reduces interaction is essentially as follows. The distillation column is actually quite decoupled at high frequencies due to the flow dynamics. Therefore, if one can close one loop

Figure 8: Response to a setpoint change in x_D.

Figure 9: Response to a step change in feed rate.

with sufficiently high gain, one can at least make the system one-way interactive and reduce the RGA. It is then possible to implement advanced controllers on top of this without regard for the robustness problems which follow when the RGA is large.

References

[1] Grosdidier, P., Morari, M. and Holt, B.R., 1985, "Closed-Loop Properties from Steady-State Gain Information", *Ind. Eng. Chem. Fundam.*, **24**, 221-235.

[2] Kister, H.Z., 1990, "Distillation operation". McGraw-Hill, New York.

[3] Mejdell, T. and Skogestad, S., 1991a, "Estimation of distillation composition from multiple temperature measurements using PLS regression", *Ind. Eng. Chem. Res.*, **30**, 12, 2543-2555.

[4] Mejdell, T. and Skogestad, S., 1991b, "Composition estimator in a pilot plant distillation column using multiple temperatures", *Ind. Eng. Chem. Res.*, **30**, 12, 2555-2564.

[5] Skogestad, S. and M. Hovd, 1992, "Simple Frequency-Dependent Tools for Control System Analysis, Structure Selection and Design". *Automatica*, **28**, 989-996.

[6] Shinskey, F.G., 1984, *Distillation Control*, 2nd Edition, McGraw Hill, New York.

[7] S. Skogestad and Morari, M., 1988, "LV-control of a High-Purity Distillation Column", *Chem. Eng. Sci.*, **43**, 1, 33-48.

[8] S. Skogestad, Lundström, P. and Jacobsen, E.W., 1990, "Selecting the best distillation control configuration", *AIChE Journal*, **36**, 5, 753-764.

[9] Stephanopoulos, G., 1984, "Chemical Process Control", Prentice-Hall, London.