Disturbance Sensitivity of Distillation Control Structures

Kurt V. Waller, Kurt E. Häggblom,
Peter M. Sandelin, Dan H. Finnerman
Process Control Laboratory
Department of Chemical Engineering
Åbo Akademi
20500 Åbo, Finland

A good deal of the research on distillation control during the last decade has concerned the structure of the control system, i.e., which inputs should be connected to which outputs (Waller, 1986). One example is the discussion of the relative merits of so-called material balance control and conventional control in dual-composition control of distillation (McAvoy and Weischedel, 1981).

Not only the manipulator part of the control scheme has been discussed. Various choices of the controlled variables have been suggested, such as sums of and differences between compositions or temperatures (Waller and Finnerman, 1987).

The various structures have usually been discussed and compared on the basis of interaction, i.e., the structure giving the smallest (steady state) interaction between the control loops has been considered the preferable one.

This note treats the importance of disturbance sensitivity, especially in relation to interaction, when choosing a control system structure. An important difference between control structures is that the compositions (or temperatures) to be controlled may be affected very differently by the same disturbances. This can be shown theoretically by control structure transformations (Häggblom, 1986; Häggblom and Waller, 1987a), and it has been verified experimentally (Finnerman and Sandelin, 1986; Waller et al., 1987). The subject has also been discussed by Shinskey (1985).

In the following, four different control structures are used to illustrate how the control qualities obtained correlate with the degree of interaction and the disturbance sensitivity implied by the open-loop schemes.

Control Structures Studied

A pilot-plant distillation column was used in the experimental part of this study. The column has 15 bubble cap plates, is 30 cm in diameter, and separates a mixture of ethanol and water. The nominal steady state at which the column was operated is given in Table 1. The control objective is to keep the temperatures on plates 4 and 14 as constant as possible in spite of disturbances in feed composition.

The four control configurations studied are denoted by the manipulators used to control the two temperatures, the first manipulator controlling the top temperature, T_4 (or y, for short), and the second the bottom temperature, T_{14} (or x). The control structures are (see Table 1 for notation):

1. The conventional or energy balance scheme (L, V), with inventory controlled by D and B
2. The material balance scheme (D, V), with inventory controlled by L and B
3. Ryskamp's (1980) scheme $[D/(L + D), V]$, with inventory controlled by $L + D$ and B
4. The two-ratio scheme $[D/(L + D), V/B]$ (Takamatsu et al., 1982, 1984; Shinskey, 1984), with inventory controlled by $L + D$ and B

Transfer functions (first order plus dead time) for the four control structures were obtained experimentally, with only inventory control loops (accumulator level and column base level) closed. Due to nonlinear process behavior and inexact measurements, the experimentally obtained process gains did not satisfy certain consistency relationships that must hold for both physical and mathematical reasons (Häggblom 1986; Häggblom and Waller 1986, 1987a,b). The process gains were therefore reconciled by optimization subject to these consistency relationships (Häggblom, 1987). The reconciled transfer functions, which are used in the simulations below, are given in Table 2. The gains between the inputs and the manipulator used *for accumulator* level control in the (L, V) scheme are given in Table 3. These process gains essentially explain the differences in disturbance sensitivity between the control structures, as shown further in the paper.

Table 1. Nominal Steady State Data

Feed flow rate, F	200 kg/h
Distillate flow rate, D	60 kg/h
Bottoms flow rate, B	140 kg/h
Feed composition, z	30 wt.%
Distillate composition	87 wt.%
Bottoms composition	5 wt.%
Reflux flow rate, L	60 kg/h
Steam flow to reboiler, V	72 kg/h
Feed temp.	65°C
Reflux temp.	62°C

Results

The disturbance sensitivity of the control structures defined above has been investigated both by experimentation with a pilot-plant distillation column and by simulations based on experimentally obtained transfer functions. Disturbances in both feed flow rate and feed composition were considered. The results of the simulations agree qualitatively, and for the most part also quantitatively, with the experimental results (Finnerman and Sandelin, 1986; Waller et al., 1987). For the sake of clarity and brevity, only results simulated for feed composition disturbances will be presented in this paper.

Open-loop responses

Figure 1 shows the open-loop responses (only level control loops closed) of the four strategies to the same disturbance in feed composition. Clearly, the (L, V) structure is the structure least sensitive to feed composition disturbances, while the (D, V) structure is the one most sensitive. The differences are more pronounced in the top temperature than in the bottom temperature. The differences in disturbance sensitivity of different control structures are qualitatively explained in the next section. That, in turn, is followed by an analytical treatment.

Table 3. Process Gains Due to Inventory Control

K_{DL}^{LV}	K_{DV}^{LV}	K_{DF}^{LV}	K_{Dz}^{LV}
−0.61	1.35	0.056	1.08

Units: flow rate, kg/h; composition, wt.%.

Single-point control

Figure 2 shows a comparison between the four control structures under single temperature control, i.e., when only one temperature control loop is closed at a time. PI controllers, with settings according to Ziegler's and Nichols' recommendations, were used as temperature controllers. The controller settings are given in Table 4.

These settings result in the same set-point responses for the corresponding control loops in the four control structures. This also means that the responses to disturbances of corresponding control loops would be the same, if the transfer functions from the disturbances to the outputs were the same in the control structures. The differences in Figure 2 between the control structures are therefore entirely due to differences in the way feed composition disturbances affect the outputs in the control structures.

The top loop responses in Figure 2 clearly correlate with the steady state gains between the disturbance and the top temperature in the respective control structures—a larger gain results in less good control. However, the response of the (L, V) structure is not quite as good as expected from steady state considerations. This is explained by the relatively small time constant in the transfer function from the disturbance to the top temperature. For the bottom loop, where the relative differences between the process gains of the control structures are small, the dynamics of the effects of the disturbance seems to be the determining factor.

The control qualities, measured for example by integrated

Table 2. Transfer Functions of Four Distillation Control Structures

Structure (u, v)	G_{yu}^{uv} / G_{xu}^{uv}	G_{yv}^{uv} / G_{xv}^{uv}	G_{yF}^{uv} / G_{xF}^{uv}	G_{yz}^{uv} / G_{xz}^{uv}
(L, V)	$\frac{-0.045e^{-0.5s}}{8.1s+1}$	$\frac{0.048e^{-0.5s}}{11s+1}$	$\frac{-0.001e^{-1.0s}}{10s+1}$	$\frac{0.004e^{-1.0s}}{8.5s+1}$
	$\frac{-0.23e^{-1.5s}}{8.1s+1}$	$\frac{0.55e^{-0.5s}}{10s+1}$	$\frac{-0.16e^{-1.0s}}{5.5s+1}$	$\frac{-0.65e^{-1.0s}}{9.2s+1}$
(D, V)	$\frac{0.074e^{-0.5s}}{14s+1}$	$\frac{-0.052e^{-1.5s}}{15s+1}$	$\frac{-0.005e^{-1.0s}}{23s+1}$	$\frac{-0.076e^{-3.0s}}{20s+1}$
	$\frac{0.38e^{-1.5s}}{15s+1}$	$\frac{0.03e^{-0.5s}}{10s+1}$	$\frac{-0.18e^{-1.0s}}{7.5s+1}$	$\frac{-1.06e^{-1.0s}}{15s+1}$
$\left(\frac{D}{L+D}, V\right)$	$\frac{6.7e^{-0.5s}}{11s+1}$	$\frac{0.010e^{-0.5s}}{13s+1}$	$\frac{-0.003e^{-1.0s}}{28s+1}$	$\frac{-0.026e^{-2.5s}}{23s+1}$
	$\frac{34e^{-1.3s}}{12s+1}$	$\frac{0.35e^{-0.5s}}{10s+1}$	$\frac{-0.17e^{-1.0s}}{4.5s+1}$	$\frac{-0.81e^{-1.0s}}{13s+1}$
$\left(\frac{D}{L+D}, \frac{V}{B}\right)$	$\frac{6.4e^{-0.5s}}{15s+1}$	$\frac{1.0e^{-0.5s}}{25s+1}$	$\frac{0.001e^{-1.0s}}{6.0s+1}$	$\frac{-0.029e^{-7.5s}}{18s+1}$
	$\frac{23e^{-1.3s}}{26s+1}$	$\frac{34e^{-0.5s}}{15s+1}$	$\frac{-0.05e^{-1.0s}}{7.5s+1}$	$\frac{-0.89e^{-1.0s}}{7.5s+1}$

Units: flow rate, kg/h; composition, wt.%; temperature, °C; time, min.

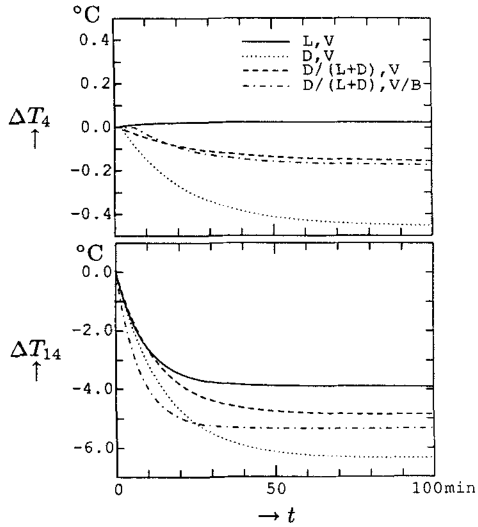

Figure 1. Open-loop responses (only level control loops closed).

Responses of four control strategies to a step change in feed composition from 27 to 33 wt. % ethanol at time $t = 0$

absolute error, correlate very well with a parameter defined as the gain divided by the time constant of the transfer function from the disturbance to the respective output variable. This parameter can thus be taken as a measure of the disturbance sensitivity of the outputs in a control structure. The disturbance sensitivity parameters for the four control structures are given in Table 5.

Two-point control

Figure 3 shows the closed-loop, two-point control responses of the four control strategies to the same disturbance in feed composition. The controller gains used for single-point control were

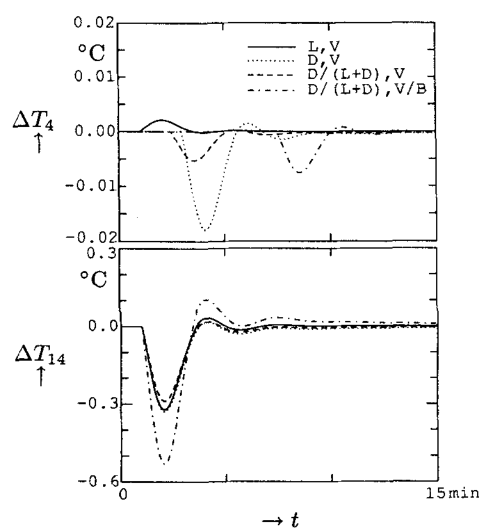

Figure 2. Single-point control.

Responses of four control strategies to a step change in feed composition from 27 to 33 wt. % ethanol at time $t = 0$
Top, control of top temperature
Bottom, control of bottom temperature

Table 4. PI Controller Settings for Single-Point Control

Structure	Top Loop K_c	Top Loop T_i	Bottom Loop K_c	Bottom Loop T_i
(L, V)	−260	1.6	26	1.6
(D, V)	270	1.6	480	1.6
$\left(\frac{D}{L+D}, V\right)$	2.4	1.6	41	1.6
$\left(\frac{D}{L+D}, \frac{V}{B}\right)$	3.4	1.6	0.63	1.6

Units consistent with units of transfer functions, Table 2.

adjusted so that reasonable responses for two-point control were obtained. For the (D, V) structure, the integration times also were adjusted. The controller settings are given in Table 6.

The closed-loop responses in Figure 3 are clearly affected by the interaction between the control loops. The control quality of the (D, V) structure is strongly degraded, entirely in accordance with the very unfavorable relative gain of the pairing used in the control structure, Table 7. Also, the control quality of the (L, V) structure is degraded, as expected from the relative gain.

The poor control quality obtained with the two-ratio scheme might seem surprising at first, considering that the relative gain is close to unity. However, due to the large difference between the dead times in the transfer functions from the disturbance to the outputs in combination with the magnitudes and signs of the process gains in the structure, Table 2, an inverse response in the top temperature will ensue when the bottom control loop is closed. For the same reason, there will also be an overshoot in the bottom temperature when the top control loop is closed. It is obvious that this behavior, due to dynamic interaction, will make a good two-point control with the two-ratio structure difficult.

An Explanation for the Difference in Disturbance Sensitivity

As shown by the process gain K_{Dz}^{LV} in Table 3, a change in the feed composition z produces a change in the distillate flow D, when L and V are constant. This happens because the latent heats of vaporization, expressed on mass basis, are not the same for the components separated, i.e., ethanol and water. Examples of realistic simulations found in the literature also show that

Table 5. Disturbance Sensitivity Parameters

Structure	Top Loop $(K_{yz}/T_{yz}) \times 10^3$	Bottom Loop $(K_{xz}/T_{xz}) \times 10^3$
(L, V)	0.47	−70.7
(D, V)	−3.80	−70.7
$\left(\frac{D}{L+D}, V\right)$	−1.13	−62.3
$\left(\frac{D}{L+D}, \frac{V}{B}\right)$	−1.61	−119

Units consistent with units of transfer functions, Table 2.

Table 6. PI Controller Settings for Two-Point Control

Structure	Top Loop K_c	Top Loop T_i	Bottom Loop K_c	Bottom Loop T_i
(L, V)	-130	1.6	13	1.6
(D, V)	130	3.3	9.0	4.5
$\left(\frac{D}{L+D}, V\right)$	0.76	1.6	13	1.6
$\left(\frac{D}{L+D}, \frac{V}{B}\right)$	1.0	1.6	0.19	1.6

Units consistent with units of transfer functions, Table 2.

assumptions of constant mass or molar overflow (i.e., essentially equal latent heats of vaporization expressed on mass or molar basis, respectively) are seldom justified, regardless of the components distilled (Häggblom and Waller 1986, 1987a,b; Häggblom, 1987). The assumption of constant molar overflow is, however, frequently used in the literature. Even if the assumption were theoretically justified, it would seldom be useful in practice, because such an assumption implies that the flows in the column should be manipulated on a molar basis (which they rarely can be).

A change in the feed flow rate F also causes a change in D when L and V are constant, as shown by the process gain K_{DF}^{LV}. Because the feed was a subcooled liquid in this study, it seems reasonable to expect D to decrease with increasing F. However, here D increased with increasing F, which is possible only because the components separated have different latent heats of vaporization expressed on mass basis.

Also the reflux flow L was subcooled, as it usually is when a total condenser is used. If the latent heats of vaporization were equal for the two components, then a change in L would induce an even greater change, but in the opposite direction, in D. However, when an assumption of constant overflow is used, the process gain K_{DL}^{LV} is usually taken to be -1. In this study, the process gain K_{DL}^{LV} was only -0.61, not -1.

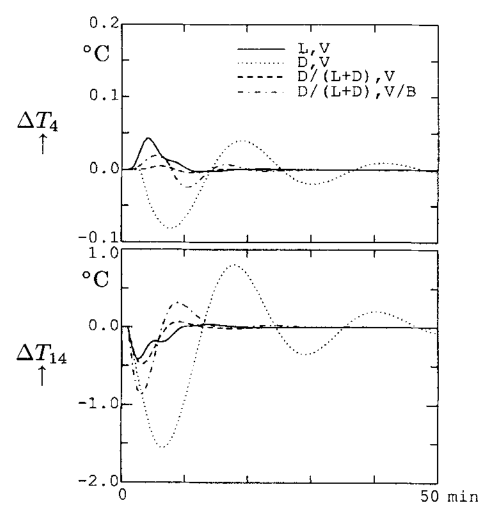

Figure 3. Two-point control.
Responses of four control strategies to a step change in feed composition from 27 to 33 wt. % ethanol at time $t = 0$

Table 7. Relative Gains

λ_{yL}^{LV}	λ_{yD}^{DV}	λ_{yR}^{RV}	λ_{yR}^{RS}
1.81	0.10	1.17	1.12

$R = D/(L + D), S = V/B$.

The differences in disturbance sensitivity between the four schemes can now be explained in the following way.

In the (L, V) scheme, L and V are constant in open-loop operation and D and B control the column inventory. A disturbance in feed composition or feed flow rate then induces a change in D and B.

In the (D, V) scheme, D and V are constant in open-loop operation and L and B control column inventory. A disturbance will then cause a change in L and B. The change in L will clearly make the (D, V) scheme behave differently from the (L, V) scheme.

In Ryskamp's scheme, $L + D$ controls the accumulator level while the ratio $D/(L + D)$ is kept constant in open-loop operation and $L + D$ controls the accumulator level. In the upper part of the column, a disturbance will therefore cause a change both in L and D. This makes Ryskamp's scheme behave as a hybrid between the (L, V) scheme and the (D, V) scheme. The process gains given in Table 2 confirm this interpretation.

The two-ratio scheme $[D/(L + D), V/B]$ works like Ryskamp's scheme, except that the ratio V/B is used as manipulator instead of B. In open-loop operation, any disturbance affecting B will therefore also affect V. For a disturbance in feed flow rate, the two-ratio scheme can thus be expected to have a disturbance sensitivity different from that of Ryskamp's scheme, but for a disturbance in feed composition, the difference should be minor. This is confirmed by the process gains given in Table 2 as far as steady state sensitivity is concerned.

The reason for the large difference in the two-ratio scheme between the dead times in the transfer functions from a feed composition disturbance to the outputs can also be explained qualitatively. A change in the feed composition will not only alter the composition profile in the column. Due to the difference in the latent heats of vaporization of the two components, the liquid flow down to the bottom of the column will also change. This will cause a change in both V and B because the ratio V/B is kept constant in open-loop operation. The change in V, in turn, induces a change in L because the ratio L/D is kept constant at the top of the column. The changes in V and L counteract each other in a way that results in a long apparent dead time for the top temperature response.

In an analogous way, the behavior of other control schemes can be explained, and even predicted. For example, the (L, B) scheme will in general behave differently from both the (L, V) scheme and the (D, V) scheme. In the (L, B) scheme, any disturbance affecting the bottom level will immediately cause a change in V, which does not happen in the (L, V) and (D, V) schemes. Hence, the schemes do not have the *same* sensitivity to a disturbance affecting the bottom level. This difference between the (L, V) structure and the (L, B) structure was recently noticed also by Retzbach (1986) in a study of extractive distillation.

Analytical Treatment

The results of the open-loop experiments, i.e., that the uncontrolled outputs of different control structures generally respond differently to the same disturbances, can be analytically explained by control structure transformations. The dynamic interaction effects observed in the closed-loop responses can also to a certain degree be explained by simple analytical expressions.

Control structure transformations

If the top temperature and the bottom temperature are denoted by y and x, respectively, then a steady state model for the (L, V) structure, including the disturbances F and z, can be written

$$\begin{bmatrix}\Delta y\\ \Delta x\end{bmatrix}=\begin{bmatrix}K_{yL}^{LV} & K_{yV}^{LV}\\ K_{xL}^{LV} & K_{xV}^{LV}\end{bmatrix}\begin{bmatrix}\Delta L\\ \Delta V\end{bmatrix}+\begin{bmatrix}K_{yF}^{LV} & K_{yz}^{LV}\\ K_{xF}^{LV} & K_{xz}^{LV}\end{bmatrix}\begin{bmatrix}\Delta F\\ \Delta z\end{bmatrix} \quad (1)$$

Here Δ denotes a deviation from the nominal steady state. The superscripts of the process gains K denote the manipulators in the control structure. By this notation, the process gain between the two variables specified by the subscripts can be distinguished from the process gain between the same two variables in another control structure.

Two other relationships, which exist due to the control of the column inventory, are

$$\Delta D = K_{DL}^{LV}\,\Delta L + K_{DV}^{LV}\,\Delta V + K_{DF}^{LV}\,\Delta F + K_{Dz}^{LV}\,\Delta z \quad (2a)$$

$$\Delta B = K_{BL}^{LV}\,\Delta L + K_{BV}^{LV}\,\Delta V + K_{BF}^{LV}\,\Delta F + K_{Bz}^{LV}\,\Delta z \quad (2b)$$

These relationships hold exactly as long as the column inventory is properly, but not necessarily perfectly, controlled.

Suppose one wants to control the top inventory by L instead of by D. Relationship 2a must still hold. To stress that L is a dependent variable when it is used for inventory control, the relationship can be rewritten as

$$\Delta L = \frac{1}{K_{DL}^{LV}}\Delta D - \frac{K_{DV}^{LV}}{K_{DL}^{LV}}\Delta V - \frac{K_{DF}^{LV}}{K_{DL}^{LV}}\Delta F - \frac{K_{Dz}^{LV}}{K_{DL}^{LV}}\Delta z \quad (3)$$

As L is no longer a free variable that can be manipulated (or kept constant), it should be eliminated from the set of input variables in the process model, Eq. 1. This gives a model

$$\begin{bmatrix}\Delta y\\ \Delta x\end{bmatrix}=\begin{bmatrix}K_{yD}^{DV} & K_{yV}^{DV}\\ K_{xD}^{DV} & K_{xV}^{DV}\end{bmatrix}\begin{bmatrix}\Delta D\\ \Delta V\end{bmatrix}+\begin{bmatrix}K_{yF}^{DV} & K_{yz}^{DV}\\ K_{xF}^{DV} & K_{xz}^{DV}\end{bmatrix}\begin{bmatrix}\Delta F\\ \Delta z\end{bmatrix} \quad (4)$$

where

$$\begin{bmatrix}K_{yD}^{DV} & K_{yV}^{DV}\\ K_{xD}^{DV} & K_{xV}^{DV}\end{bmatrix}=\begin{bmatrix}\dfrac{K_{yL}^{LV}}{K_{DL}^{LV}} & K_{yV}^{LV}-K_{yL}^{LV}\dfrac{K_{DV}^{LV}}{K_{DL}^{LV}}\\ \dfrac{K_{xL}^{LV}}{K_{DL}^{LV}} & K_{xV}^{LV}-K_{xL}^{LV}\dfrac{K_{DV}^{LV}}{K_{DL}^{LV}}\end{bmatrix} \quad (5a)$$

and

$$\begin{bmatrix}K_{yF}^{DV} & K_{yz}^{DV}\\ K_{xF}^{DV} & K_{xz}^{DV}\end{bmatrix}=\begin{bmatrix}K_{yF}^{LV}-K_{yL}^{LV}\dfrac{K_{DF}^{LV}}{K_{DL}^{LV}} & K_{yz}^{LV}-K_{yL}^{LV}\dfrac{K_{Dz}^{LV}}{K_{DL}^{LV}}\\ K_{xF}^{LV}-K_{xL}^{LV}\dfrac{K_{DF}^{LV}}{K_{DL}^{LV}} & K_{xz}^{LV}-K_{xL}^{LV}\dfrac{K_{Dz}^{LV}}{K_{DL}^{LV}}\end{bmatrix} \quad (5b)$$

Equation 4 is now a model for the (D, V) structure that includes the same disturbances F and z as the model for the (L, V) structure.

This shows, in particular, that generally the process gains between the outputs and the disturbances are not equal in the two control structures. For example, K_{yz}^{DV} being equal to K_{yz}^{LV} would demand that K_{yL}^{LV} or K_{Dz}^{LV} be equal to zero. Clearly, K_{yL}^{LV} is not zero, and in practice K_{Dz}^{LV} is hardly ever zero either, as pointed out in the previous section.

In a similar manner the process gains can be derived for the other two control structures considered. This involves a linearization of the two ratios $D/(L + D)$ and V/B. When Eqs. 2a and 2b are used, it must be noted that the process gains in the equations satisfy certain consistency relationships imposed by the steady state material balance $F = D + B$, i.e., the process gains are not independent. However, these system transformations are more elegantly brought about by using a compact matrix notation, as illustrated in Häggblom (1986) and Häggblom and Waller (1987a,b).

Dynamic interaction

The poor control quality obtained with the two-ratio scheme under two-point control can also be explained analytically. With the notation $R = D/(L + D)$ and $S = V/B$, a model for this control structure can be written

$$\begin{bmatrix}\Delta y(s)\\ \Delta x(s)\end{bmatrix}=\begin{bmatrix}G_{yR}^{RS} & G_{yS}^{RS}\\ G_{xR}^{RS} & G_{xS}^{RS}\end{bmatrix}\begin{bmatrix}\Delta R(s)\\ \Delta S(s)\end{bmatrix}+\begin{bmatrix}G_{yF}^{RS} & G_{yz}^{RS}\\ G_{xF}^{RS} & G_{xz}^{RS}\end{bmatrix}\begin{bmatrix}\Delta F(s)\\ \Delta z(s)\end{bmatrix} \quad (6)$$

In order to investigate the main interactive effects of one closed loop on the other, let us consider the theoretical case where the bottom temperature x is perfectly controlled by S. Assuming, for simplicity, F and the set point for x to be constant gives the relation

$$\Delta S(s) = -\frac{G_{xR}^{RS}}{G_{xS}^{RS}}\Delta R(s) - \frac{G_{xz}^{RS}}{G_{xS}^{RS}}\Delta z(s) \quad (7)$$

Inserting this equation into the expression for y in Eq. 6 gives, with F constant,

$$\Delta y(s) = \left(G_{yR}^{RS} - G_{yS}^{RS}\frac{G_{xR}^{RS}}{G_{xS}^{RS}}\right)\Delta R(s) + \left(G_{yz}^{RS} - G_{yS}^{RS}\frac{G_{xz}^{RS}}{G_{xS}^{RS}}\right)\Delta z(s) \quad (8)$$

As shown by Eq. 8, closing the bottom loop not only affects the way the manipulator R affects the output y, but also the way the disturbances (here the feed composition z) affect y. By inserting the transfer functions given in Table 2 it is confirmed that a step change in z, for example, will result in an inverse response in y.

Of course, in practice perfect control cannot be obtained by pure feedback control. However, the manipulator S tends to follow Eq. 7, which can be taken as a justification for this short-cut analysis.

Conclusions

The disturbance sensitivity of four distillation control structures is investigated by simulations based on transfer functions experimentally obtained from a pilot-plant distillation column. The four control structures studied are the conventional structure (L, V), the material balance structure (D, V), Ryskamp's structure $[D/(L + D), V]$, and the two-ratio structure $[D/(L + D), V/B]$. The results reported are for disturbances in feed composition, but the phenomena observed are relevant to any kind of disturbance.

It is found that the open-loop responses (only level control loops closed) to a disturbance, and especially the gains, may be quite different for different control structures. In one-point control the closed-loop responses depend on both the gains and the dynamics of these open-loop responses. The control qualities obtained correlate well with a parameter defined as the gain divided by the effective time constant of the corresponding open-loop responses. Hence, this parameter can be taken as a measure of the disturbance sensitivity of the outputs in a control structure.

It may seem peculiar that the open-loop responses of different control structures can be very different for the same disturbance. The explanation is twofold. First, the open-loop responses shown are actually responses of partially controlled systems: there are two level control loops closed in the systems. This makes two manipulators, often different in different control structures, change dependently on the way a disturbance affects the levels. Second, it does not imply that two physical manipulators are constant, although the free manipulators of a control structure are constant when the outputs considered are not controlled by these manipulators. For example, keeping the ratio $D/(L + D)$ constant will change both L and D if at least one of the two variables is used for level control.

The open-loop responses are different for different disturbances. This means, for example, that a scheme may be strongly affected by one disturbance and only little by another, while for the same disturbances the opposite may be true for another scheme.

In two-point control, the interaction between the control loops will also affect the control qualities. The more the relative gains of the chosen pairings differ from unity, the more the control qualities tend to be degraded. However, even if the relative gains are close to unity, the closed-loop responses obtained by two-point control may be much worse than those obtained by one-point control. The reason for this is the dynamic interaction, which may be especially strong when a disturbance affects the controlled variables through different dead times.

As this study has illustrated, properties of partially controlled systems may be very interesting and important, as may also transformations between various control structures in such systems. This subject is treated more extensively in Häggblom (1986) and Häggblom and Waller (1986, 1987a,b).

Acknowledgment

Financial support from the Academy of Finland, Research Council for Technical Sciences, is gratefully acknowledged.

Literature Cited

Finnerman, D., and P. Sandelin, "Four Control Configurations for Two-Product Control of a Distillation Column," M.Sc. Theses. Åbo Akademi (1986; in Swedish).

Häggblom, K. E., "Consistent Modeling of Partially Controlled Linear Systems with Application to Continuous Distillation," Lic. Tech. Thesis, Åbo Akademi (1986; in Swedish).

———, "Estimation of Consistent Process Gains for Distillation Control Structures," *Rept. 87-8*, Process Control Lab., Åbo Akademi (1987).

Häggblom, K. E., and K. V. Waller, "Relations between Steady State Properties of Distillation Control System Structures," *Prep. IFAC Symp. Dynamics and Control of Chemical Reactors and Distillation Columns*, Bournemouth, U.K., 243 (Dec., 1986).

———, "Transformations and Consistency Relations of Distillation Control Structures. I: Theory," *Rept. 87-6*, Process Control Lab., Åbo Akademi (1987a).

———, "Transformations and Consistency Relations of Distillation Control Structures. II: Applications," *Rept. 87-7*, Process Control Lab., Åbo Akademi (1987b).

McAvoy, T. J., and K. Weischedel, "A Dynamic Comparison of Material Balance vs. Conventional Control of Distillation Columns," *Proc. 8th IFAC World Cong.*, Kyoto, 2773 (1981).

Retzbach, B., "Control of an Extractive Distillation Plant," *Prepr. IFAC Symp. Dynamics and Control of Chemical Reactors and Distillation Columns*, Bournemouth, U.K., 225 (Dec., 1986).

Ryskamp, C. J., "New Strategy Improves Dual Composition Column Control," *Hydrocarb. Proc.*, **59**(6), 51 (1980).

Shinskey, F. G., *Distillation control*, 2nd ed., McGraw-Hill, New York (1984).

———, "Disturbance-Rejection Capabilities of Distillation Control Systems," *Proc. ACC*, Boston, 1072 (June, 1985).

Takamatsu, T., I. Hashimoto, and Y. Hashimoto, "Multivariable Control System Design of Distillation Columns System," *Proc. PSE 82*, Kyoto, 243 (1982).

———, "Dynamic Decoupler Sensitivity Analysis and Its Application in Distillation Control," *Prepr. 9th IFAC World Cong.*, **3**, 98, Budapest (1984).

Waller, K. V., "Distillation Column Control Structures," *Prepr. IFAC Symp. Dynamics and Control of Chemical Reactors and Distillation Columns*, Bournemouth, U.K., 1 (Dec., 1986).

Waller, K. V., and D. H. Finnerman, "On Using Sums and Differences to Control Distillation," *Chem. Eng. Commun.*, **56**, 253 (1987).

Waller, K. V., D. H. Finnerman, P. M. Sandelin, K. E. Häggblom, and S. E. Gustafsson, "An Experimental Comparison of Four Control Structures for Two-Point Control of Distillation," Accepted, *Ind. Eng. Chem. Res.* (1987).

Manuscript received Feb. 2, 1987, and revision received Dec. 30, 1987.