Ratio and bidirectional control applied to distillation columns

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Abstract: Ratio control and bidirectional inventory control are simple and powerful data-based strategies for feedforward control and coordination, respectively. By "data-based" it is meant that no explicit process models is needed, which simplifies implementation. The paper demonstrates the power of these simple architectures when applied to distillation columns.

Keywords: Distillation control, advanced regulatory control, decentralized control, inventory control, selectors, throughput manipulator.

1. INTRODUCTION

The choice of control structure or architecture is the most important part of the design of a control system (Foss, 1973; Skogestad, 2023a). With the right choice of architecture, it may be possible to apply data-based control, based on simple PID controllers, which eliminates the time-consuming and costly step of obtaining detailed dynamic models, which are needed for conventional advanced control solutions, like model predictive control (MPC).

In this paper, we focus on two important data-based schemes. The first is ratio control, which may be viewed as a special case of feedforward control, but which does *not* require a model for how the disturbance d and input u affect the controlled variable y and does not even require that we measure the disturbance. Rather, it is based on the *process insight*, that for certain processes, namely the ones that satisfy the scaling assumption, the controlled property variable y will remain constant provided the ratio(s) u/d (both assumed to be extensive variables) is kept constant. For example, the scaling assumption applies to distillation columns with constant stage efficiency.

The other control strategy studied in this paper is bidirectional inventory control (Shinskey, 1981; Zotică et al., 2022; Skogestad, 2023a), which may be viewed as an override scheme for keeping the production going when constraints in the process are encountered. The scheme combines a "split-parallel" control scheme (two parallel inventory controllers with different setpoints) with a MIN-selector to do the override.

2. THEORETICAL BASIS OF RATIO CONTROL: THE SCALING ASSUMPTION

Ratio control is based on the **scaling assumption** which for a process at steady state may be formulated as follows (Skogestad, 2023b): *Scaling all the independent extensive variables X_i by a factor k, with the independent intensive variables x_i constant, scales all the dependent extensive variables Y by the same factor k and keeps all the dependent intensive variables y constant.*

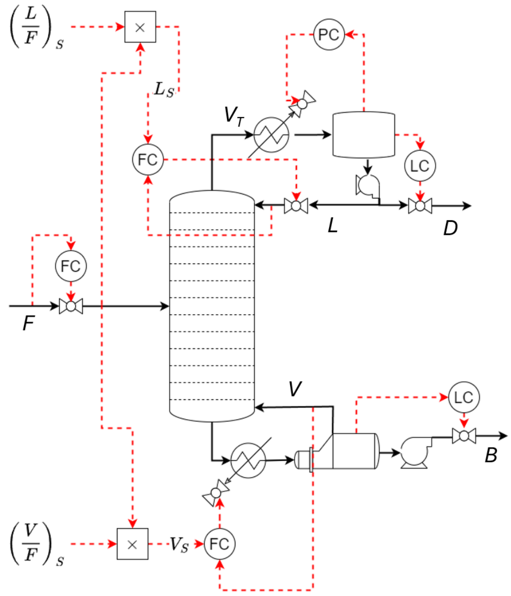

Fig. 1. Ratio control of distillation column with fixed L/F and V/F (this is scheme B4 in Section 3).

Mathematically, the scaling assumption implies for the intensive (y) and extensive (Y) dependent variables:

$$y = f_y(x_1, x_2, kX_1, kX_2, kX_3) = f_y(x_1, x_2, X_1, X_2, X_3) \quad (1a)$$
$$Y = f_Y(x_1, x_2, kX_1, kX_2, kX_3) = kf_Y(x_1, x_2, X_1, X_2, X_3) \quad (1b)$$

A simple example is a mixing process (food recipe), where we know that if all feed flows (extensive variables) are increased proportionally (with fixed ratios), then the production rate increases proportionally, and, most importantly, all dependent intensive variables (including product compositions) remain constant (at steady state).

From the scaling assumption and (1) we arrive at the following important conditions for the use of ratio control:

- (R1) The systems must satisfy the scaling assumption.
- (R2) The controlled variable y is implicitly assumed to be an intensive variable, for example, composition, density, viscosity or temperature.
- (R3) Since all extensive variables must be scaled by the same factor k, there can only be one independent extensive disturbance variable. This variable is sometimes called the “basis”, “wild variable” or “throughput manipulator” (TPM).
- (R4) If the system has n independent extensive variables, then we need to keep $n-1$ ratios (or other intensive variables) constant in order to keep the controlled variable y constant (at steady state).

Note that the theory of ratio control assumes constant independent intensive variables x_i (such as feed composition), but in practice such variations may be handled by outer feedback controllers which adjust the ratio setpoints.

3. RATIO CONTROL FOR DISTILLATION

For the scaling assumption to hold for distillation, we need to assume (a) constant pressure, (b) vapor-liquid equilibrium, (d) constant feed composition, and (d) constant stage efficiency (or number of theoretical stages) in each section.

For a typical two-product distillation column with a single feed and two products, there are with constant pressure, $n = 3$ independent extensive variables at steady state, for example, feedrate F, reflux L and boilup V (but this set is not unique, for example, any of these three variables may be replaced by one of the product rates D and B). Then, from (R4), we get that with constant feed composition, all intensive variables y in the column will remain constant if we keep $n-1 = 2$ dependent intensive variables constant, for example, two ratios, like V/F and L/F (Figure 1). However, note that any other two specifications of two intensive variables will give constant property variables y, for example, constant reflux-to-feed ratio (L/F) and constant temperature T somewhere in the column (scheme B3 in Figure 2).

It has been common since the 1950s to recommend using a constant reflux-to-feed ratio (L/F) and Luyben (2022) says that it is the most frequently applied distillation column configuration. However, one needs to be careful about this policy. First, there is a potential dynamic problem in that the change in reflux L may come too soon, because with a liquid feed it will take time for the composition in the top to change; so from this point of view the L/D ratio may be better. However, a more serious problem is that the heat input (V) may saturate. Indeed, already Young (1955) (footnote, p. 321) warns that it may not be advisable to keep L/F constant because it requires that V must change. Furthermore, if the heat input (V) reaches its maximum value (and becomes constant) then from (R3) and (R4) the scaling assumption does not hold. In fact, it turns out that with constant heat input (V), keeping the ratio L/F constant (scheme B2; see Figure 3) is worse than the simpler strategy of keeping reflux L constant (scheme B1, not shown in any figure). This is easy to explain, because for a distillation column, the most important operating parameter is the product-to-feed split ratio, D/F

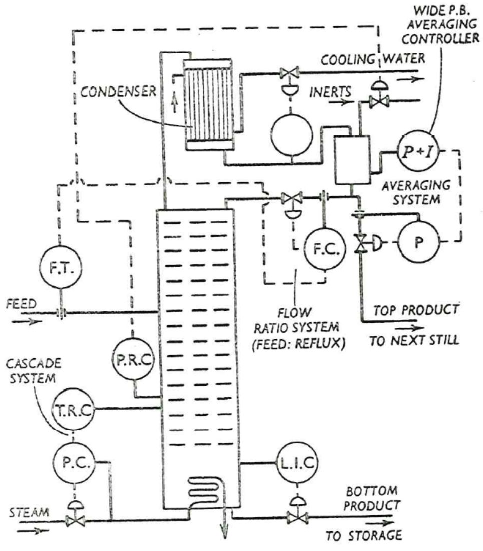

FIG. 18.11. Examples of the use of complex systems.

Fig. 2. Ratio control of distillation column with fixed L/F and temperature (Young, 1955, page 323) (scheme B3).

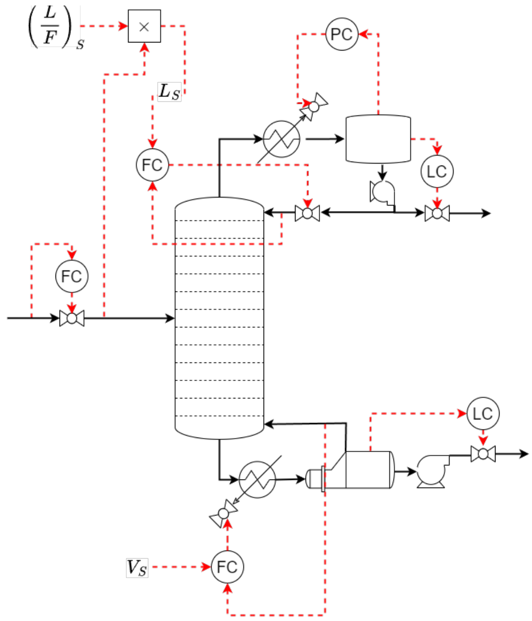

Fig. 3. Not recommended ratio control of distillation column with fixed L/F and V (scheme B2).

or B/F=1-D/F. From the material balance, D/F should be approximately equal to the fraction of light components in the feed in order to obtain pure products. So, for a change in feedrate F with constant feed composition, we want to keep D/F (approximately) constant, that is, D should increase when F increases. If we have a liquid feed, then with V and L constant (scheme B1), all the increase in F will come out in the bottom (B), so D remains unchanged, whereas we want it to increase. However, with a control structure with constant V and constant reflux ratio L/F (scheme B2), things are even worse (especially for columns with a large L/F) because here L increases which means that D decreases, which is the opposite of what we want.

Parameter	Value
Number of theoretical stages	40
Feed stage (numbered from top)	34
Feed flow F	100 kmol/h
Feed mole fraction (methanol)	0.50
Feed state	Liquid
Column pressure	2 bar
Reflux ratio L/D	1.013
Top product, x_D (water)	0.001
Bottom product, x_B (methanol)	0.001
Reboiler type	Kettle
Vapor-liquid equilibrium model	NRTL

Table 1. Nominal operating data for the methanol-water distillation column

To confirm this, we performed dynamic simulations with Aspen Plus Dynamics (for files, see Bang (2024)). We consider the separation of a mixture of methanol and water, with column data given in Table 1. The responses to a 10% increase in feedrate (F) are shown in Figure 4 for the following four control schemes:

- Scheme B1. Constant L and V (bad)
- Scheme B2. Constant L/F and V (even worse)
- Scheme B3. Constant L/F and temperature in bottom section (best, but V must not saturate)
- Scheme B4. Constant L/F and V/F (good, but V must not saturate)

The comments in parenthesis summarize the observed control performance. The main control objective is to keep the two product impurities at mole fraction 10^{-3} (corresponding to the value 3 in the (-log) "purities" plot).

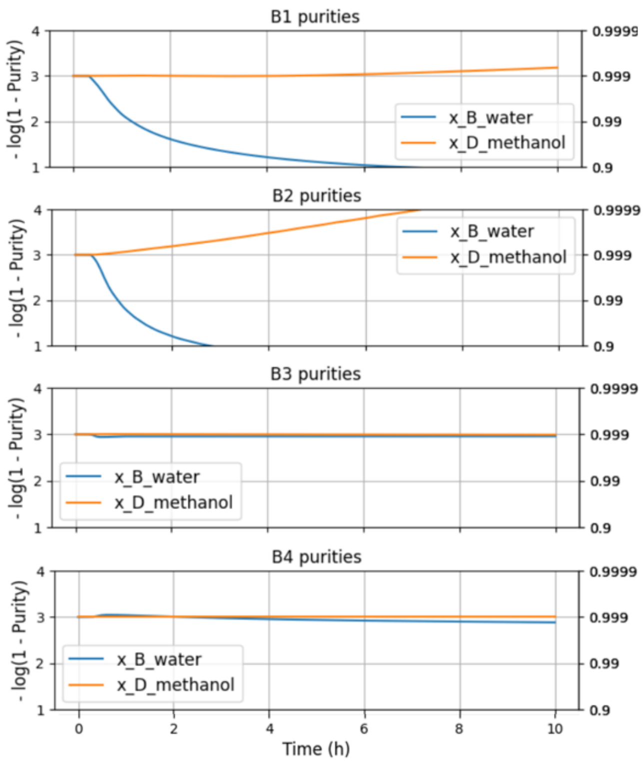

Fig. 4. Ratio control: Product composition responses for 10% feed flow disturbance with control structures B1, B2, B3 and B4.

As expected from the above discussion, scheme B2 with constant L/F and V is the worst, and scheme B2 (constant L and V) is also poor. Schemes B3 and B4 are both good and give constant intensive variables (including compositions) at steady state. Scheme B3 is a little better dynamically. A scheme with V/F and L constant (not shown in the simulations) has similar problems as the L/F scheme with V constant (B2), in particular if the feed is vapor.

Note that both schemes B3 and B4 become scheme B2 (the worst scheme) if V saturates. This implies that one should be careful about applying L/F ratio control if it is likely that V saturates. On the other hand, V/F ratio control, with L used to control an intensive variable, may be better because L rarely saturates. For example, a structure with constant V/F and constant column temperature T will work well even if V saturates. This V/F-T structure (Figure 8) is therefore used in the next section on bidirectional control where the main topic is how to handle saturation.

4. BIDIRECTIONAL CONTROL FOR DISTILLATION

Bidirectional inventory control (Figure 5) was first proposed by Shinskey (1981), who also suggested to make use of the flexibility offered by varying the inventory between a high (H) and low (L) setpoint to maximize the production rate. The two MIN-selectors are needed to perform the overrides and the transitions back again when the constraints are no longer active. An application to a gas-liquid separator is shown in Figure 6. The "H-overrides" for level and pressure, both decrease the feedrate F in case of saturation (maximum constraint) in the bottom liquid product (B) and top vapor product (V), respectively. The maximum constraints are represented here by the signals B_s and V_s. A simple corresponding bidirectional control scheme for a distillation column is shown in Figure 7. It has H-overrides on the level controllers to handle maximum constraints (B_s, D_s) on bottom (B) and top (D) products.

Figure 8 shows an advanced bidirectional distillation scheme with four overrides which reduce the feedrate when constraints are encountered. The overrides are for bottom level, top level, pressure and bottom composition and handle constraints (max) for B, D, cooling, and heating, respectively. The override for bottom composition (which activates when the methanol fraction has increased from 1.e-3 to the H-setpoint of 1.e-2) is to handle the case when the heat input (which generates the boilup V) saturates at maximum. The inner fast temperature loop, which aims at stabilizing the column profile (and keeping the product split D/F constant on a fast time scale), uses reflux (L) because the boilup (V) may saturate. An important reason for using the V/F ratio (and not just V) as the manipulated variable to control the bottom composition is that otherwise the top level and pressure H-overrides to the feed F will be slow, especially if the feed is liquid. The dynamics are much faster with the V/F ratio scheme because the boilup V has a much more direct effect on pressure and top (condenser) level. The setpoints for the V/F ratio and the column temperature (T) are set by the bottom and top composition controllers (CC), respectively.

The dynamic simulations in Figure 9 (with PI tuning parameters in Table 2) show clearly the effectiveness of

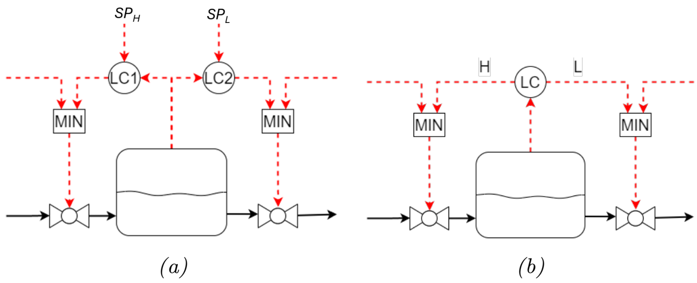

Fig. 5. Bidirectional inventory control of a single unit (Shinskey, 1981). (a) "Correct" flowsheet. (b) Simplified representation (used in this paper)

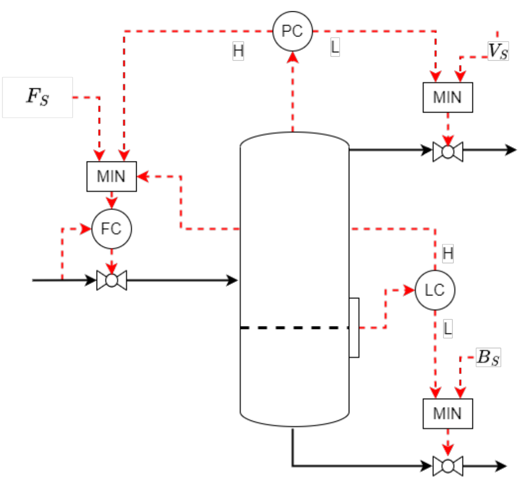

Fig. 6. Bidirectional inventory control of a gas-liquid separator with H-overrides for pressure and level.

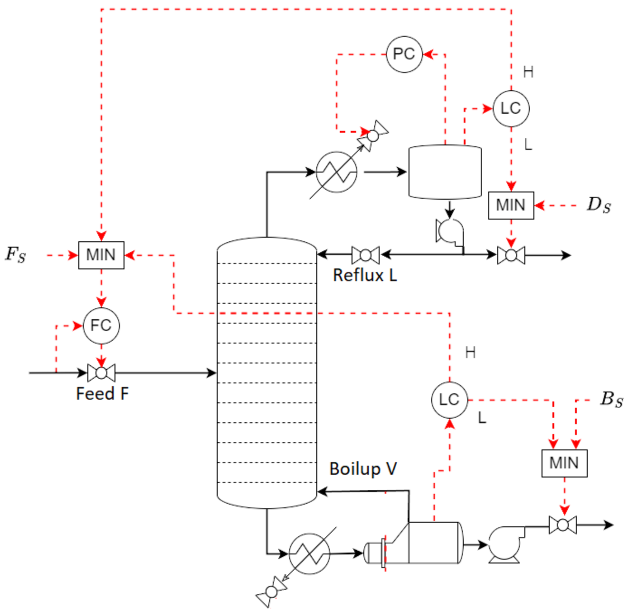

Fig. 7. Simple bidirectional inventory control for distillation with H-overrides for top and bottom levels.

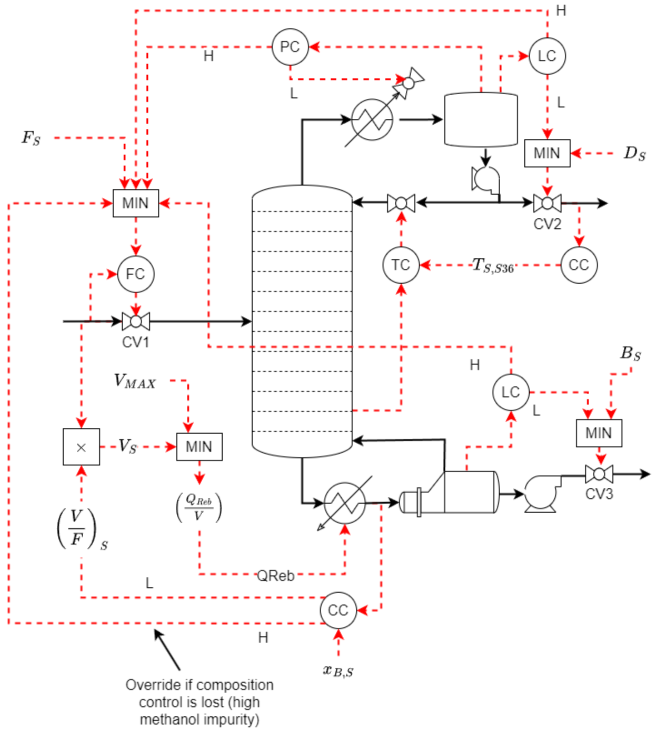

Fig. 8. Advanced bidirectional control for distillation column with four H-overrides which reduce the feedrate when constraints are encountered. It is based on the V/F-T configuration with two outer composition controllers (CC).

Controller	τ_C	K_C	τ_I [s]	**Setpoint**
$LC_{L,D}$	*	-50	7200	1.9 m
$LC_{L,B}$	*	-50	7200	1.9 m
PC_L	*	3	28	2.0 bar
TC_{S36}	60s	5.3	2336	set by CC
CC_D	600s	-528	3600	1.e-3
$LC_{H,D}$	*	10	7200	2.1 m
$LC_{H,B}$	*	10	7200	2.1 m
PC_H	*	1	3600	2.05 bar
$CC_{L,B}$	*	7.95	7500	1.e-3
$CC_{H,B}$	*	1	600	1.e-2

Table 2. Tuning parameters for the distillation column in Figure 8. The controllers were tuned sequentially in the order given in the table. From the desired τ_C, we obtain K_C and τ_I from the SIMC rules and open-loop experiments performed in Aspen Plus. The controllers marked (*) were tuned manually based on qualitative process dynamics. Simple anti-windup schemes with bounds on the controller output are used for most PI controllers.

the advanced bidirectional (override) control strategy. It handles changes in the active constraint location, where different MVs may saturate (see details in Table 3). At t=0.5h, the feed rate is increased, and this is handled nicely without any overrides being activated (as no constraints are encountered). Thus, the location of the throughput manipulator (TPM) remains at the feed. The location of the active constraint (and thus the location of the TPM) is given by the lowest curve in the subplot

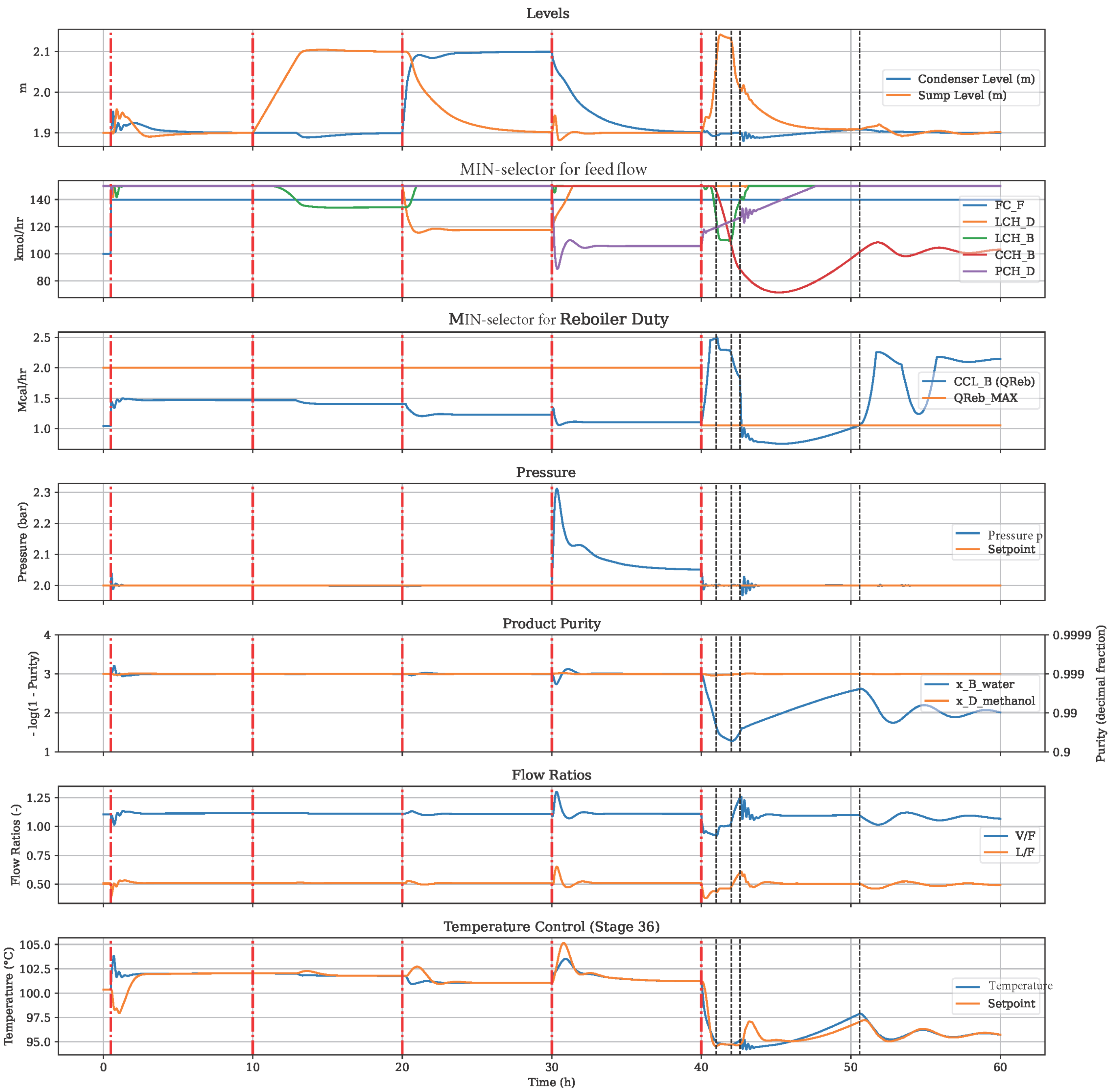

Fig. 9. Simulation results for the advanced bidirectional distillation column control scheme in Figure 8. Each vertical red line signifies one of the five events described in Table 3. Each vertical black line after t=40 h signifies the activation or deactivation of an override controller.

"MIN-selector for feed flow" which shows the five inputs to the feed MIN-selector. Initially, the blue line FC_F is the minimum, which means that the TPM is at the feed. At t=10 h, a constraint on the bottom flow (valve CV3) is introduced (so that this becomes the TPM). Initially, we lose control of bottom (sump) level, but as the bottom level approaches the H-setpoint (2.1 m), the H-override bottom level controller (LCH_B) is activated and reduces the desired feedrate, which at about t=13 h becomes the actual feedrate. This happens at the time where the green line is the smallest in the subplot "MIN-selector for feed flow".

Next, at t=20 h, a constraint on the distillate flow (D) is introduced, so that we lose control of top (condenser) level. The H-override condenser level controller ($LC_{H,D}$) reduces the desired feedrate, which shortly after becomes the actual feedrate (at the time where the orange line is the smallest in the subplot "MIN-selector for feed flow"). Through the V/F ratio control, this reduces the boilup V and condenser level is stabilized at the H-value (2m).

Next, at t=30 h, a constraint on the cooling duty (Qcond) is introduced, so that we lose control of pressure. The H-override pressure controller (PC_H) (which is a bit slow)

Time (h)	Constraint (TPM)	Initial value	New Limit
0.5	F_s [kmol/h]	100	140 (+40%)
10	CV3 (bottoms valve)	31.98%	22.39% (-30%)
20	CV2 (distillate valve)	44.90%	31.45% (-30%)
30	QCond [Mcal/h]		-0.963
40	QReb [Mcal/h]		1.107

Table 3. Summary of constraint limit changes that result in activating H-overrides and movement of the TPM in Figure 8.

reduces the desired feedrate, which shortly after becomes the actual feedrate (at the time where the purple line is the smallest in the subplot "MIN-selector for feed flow"). Through the V/F ratio control, this reduces the boilup V and pressure is eventually stabilized at the H-value (2.05 bar).

Finally, at t=40 h, a constraint on the heating duty (QReb) is introduced, which is equivalent to a max-constraint on boilup V, so that we lose control of bottom composition and the H-override composition controller ($CC_{H,B}$) should further reduces the feedrate. However, initially, before $CC_{H,B}$ is activated, there is a loss of control of bottoms (sump) level, which by controller $LC_{H,B}$ results in an additional reduction in feedrate (see green line at t=41h) Then, at t=42 h, $LC_{H,B}$ becomes inactive as $CC_{H,B}$ becomes active. At t=42.6 h, also the "normal" composition controller $CC_{L,B}$ temporary becomes active (see "MIN-selector reboiler duty" subplot) because QReb < QReb_MAX (due to the decrease in the feedrate by $CC_{H,B}$), and it tries to bring the composition back to its "normal" L-setpoint of 1.e-3. The result of these interactions between the H-overrides for bottom composition and bottom level is that it takes about 20 hours before the bottom impurity (methanol) is reduced from its "normal" setpoint of SPL=1.e-3 and stabilizes at the H-setpoint of SPH=1.e-2.

Several measures may be taken to avoid this problem and have been tested in simulations (Bang, 2024). One is to change the H-setpoint to a value closer to the L-setpoint (e.g., to SPH=2.e-3), a second is to use higher controller gains in the H composition controller ($CC_{H,B}$), and a third is to introduce tracking for anti-windup ($CC_{H,B}$) so that the controller output (red line in "Select minimum feed flow" subplot) does not start so far away (150 kmol/h) from the present feed value (105 kmol/h). All of these measures reduce the time for the H-override composition controller ($CC_{H,B}$) to activate. A fourth possibility is to increase the sump level H-setpoint (e.g. from SPH=2 m to SPH=2.5 m) to delay the time before this controller ($CC_{H,B}$) is activated. In any case, the simulations in Figures 9 demonstrate the robustness of the proposed control architecture.

Discussion. The control scheme in Figure 8 is for the case where composition control of the top product is important. If instead tight composition control in the bottom is important, then the inner temperature loop should be in the bottom using boilup (V/F). In this case, an additional override for reflux (L) to take over composition control in the bottom may be needed.

The thermodynamics of methanol-water are somewhat unusual with a pinch-like behavior in the entire top section (see McCabe-Thiele diagram in Bang (2024)). For this reason the temperature sensor for reflux (L) in Figure 8 is not placed in the top section (above the feed), as would be expected, but rather in the bottom part (TC_{S36}, only 4 stages above the reboiler), because it has a high gain from reflux to temperature. As expected, and seen in the simulations, with the temperature sensor for reflux located in the bottom, there are severe interactions with the bottom composition control loop, which forces the bottom composition loop to be very slow. In any case, the main point of this section is to show how to use bidirectional control for distillation, and not to find a non-interacting structure for composition control.

5. CONCLUSION

Intuitively, one may expect that ratio control (often viewed as a special case of feedforward control) and bidirectional inventory control (a plantwide coordination scheme) require a detailed process model. However, ratio control is based on process insight rather than process models and bidirectional control is based on feedback rather than model-based coordination. For distillation columns, ratio control is sometimes applied incorrectly by keeping one ratio and one flow constant, and this may result in very poor performance (see scheme B2). Bidirectional control has so far been not applied to distillation columns in a systematic way, and the paper shows that it is relatively simple and powerful. The proposed control scheme in Figure 8 maximizes the throughput when constraints are encountered without the need for model-based coordination using MPC or real-time optimization (RTO).

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