Ratio and bidirectional control applied to distillation columns

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Abstract: Ratio and bidirectional control are simple and powerful data-based strategies for feedforward control and coordination, respectively. In terms of ratio control, distillation columns, with given feed and pressure, have two steady-state degrees of freedom, so two ratios need to be set. However, ratio control is sometimes applied incorrectly by keeping one ratio and one flow constant, and this may result in very poor performance. Bidirectional control has so far been not applied to distillation columns in a systematic way, and the paper shows that it is simple and powerful. The results are confirmed using dynamic simulation in Aspen Dynamics of a methanol-water distillation column with 40 theoretical stages.

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 designing an automatic process control system (Foss, 1973). In this paper, we focus on two important data-based schemes. The first is ratio control, which is a special case of feedforward or decoupling control, but it does *not* require a model for how the disturbance d and input u affects the controlled variable y. Rather, it is based on the *process insight*, that for certain processes, namely the ones that satisfy the scaling principle, the controlled property variable y will be constant provided the ratio(s) u/d (both assumed to be extensive variables) is kept constant. The scaling principle applies to distillation columns with constant stage efficiency. However, physical insight and intuition has its limitations, and because of the lacking theoretical basis for ratio control, it is sometimes used wrong. As an example, it has been common to recommend using a fixed reflux ratio L/F in a distillation column (e.g., theeYoung55 (p. 321) Luyben (2022)), But, as discussed in this paper, this is not a good solution if the heat input (or equivalently, the boilup V) is constant, for example, because of saturation. This is because, according to the theory or ratio control (Skogestad, 2025), *all extensive variables* (including the boilup V) must be increased proportionally when the feed rate F increases. A structure with constant ratios V/F and L/F is shown Figure 1, but there are also other acceptable feedforward architectures as discussed in Section 3.

The other control strategy studied in this paper is bidirectional inventory control, 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 scheme" with two inventory (level) controllers with different setpoints (H and L) with a MIN-selector to do the override. The scheme was first proposed by (Shinskey, 1981), who also suggested to make

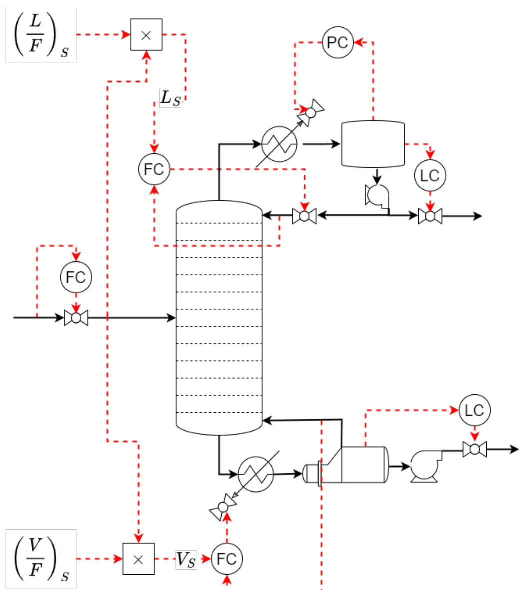

Fig. 1. Ratio control of distillation column with fixed L/F and V/F (that is, with feedrate F as the basis (wild flow). (scheme B4 in Section 3)

use of the flexibility offered by letting the inventory vary between a high (H) and low (L) setpoint to maximize the production rate. Bidirectional inventory control of a single unit is shown in Figure 2 and of units in series in Figure 3 (Shinskey, 1981; Zotică et al., 2022; Skogestad, 2023). Here, IC represents inventory control, which is level control (LC) for liquids and pressure control (PC) for gases. To simplify the block diagram, the two inventory controllers (with H and L setpoints) are in this paper combined into one, as shown in Figure 2b.

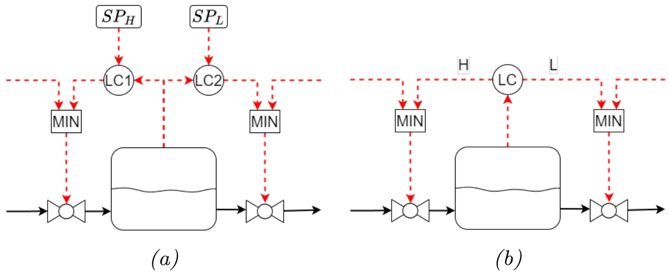

Fig. 2. Bidirectional inventory control of a single unit. (a) "Correct" flowsheet with two split parallel inventory controllers (LC1 and LC2) with different setpoints (SP_H and SP_L). (b) Simplified representation (used in this paper) where one block (LC) represents the two parallel controllers and where the setpoints (SP_H and SP_L) are shown indirectly as just H and L above the signals.

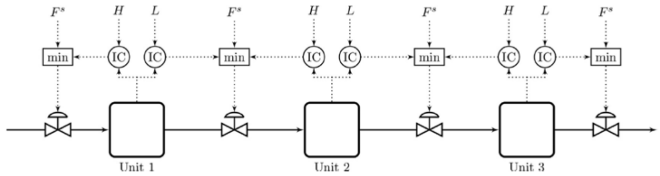

Fig. 3. Bidirectional inventory control of units in series (Shinskey, 1981). The operator or another controller can set the desired throughput F_k^s at any given location k (between units).

For a series process, a new production constraint (bottleneck) must result in giving up the original throughput manipulator (TPM). Intuitively, one may think that this requires some plantwide coordination. However, in the bidirectional scheme in Figure 3, the coordination is solved using local controllers, where the information about where the location of the constraint is discovered indirectly. For example, the inability stay at the low inventory setpoint (L), means that some downstream unit has encountered a constraint and has become the new bottleneck. One additional "magic" feature of the bidirectional scheme in Figure 3 is that it delays the moving of the throughput manipulator and thus indirectly maximizes production (Shinskey, 1981; Zotică et al., 2022).

2. THEORETICAL BASIS OF RATIO CONTROL: THE SCALING PROPERTY

Ratio control is based on the scaling property which for a steady-state process may be formulated as follows (Skogestad, 2025): *For a process that satisfies the* **scaling property***, we have that scaling (changing) all the independent extensive variables F_i by the same factor k (with all independent intensive variables x_i constant) scales (changes) all the dependent extensive variables Y by the same factor k and keeps all the dependent intensive variables y constant.*

Mathematically, the scaling property implies that:

$$y \text{ intensive}: \quad y = f_y(x_1, x_2, kF_1, kF_2) = f_y(x_1, x_2, F_1, F_2) \tag{1a}$$

$$Y \text{ extensive}: \quad Y = f_y(x_1, x_2, kF_1, kF_2) = kf_y(x_1, x_2, F_1, F_2) \tag{1b}$$

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

Note that a ratio (for example, F_1/F_2 or F/F_1) is itself an intensive variable. Then, from Eq.(1a) we arrive at the following important general conclusions for the *use of ratio control*:

(1) The system must satisfy the scaling property.
(2) The controlled variable(s) y is implicitly assumed to be an intensive variable, for example, a property variable like composition, density or viscosity, but it could also be a temperature or pressure.
(3) If the system has n independent extensive variables at steady state (including disturbances), then we need to use $n-1$ of these extensive variables to keep $n-1$ independent intensive variables constant (for example, $n-1$ ratios) in order to keep all (other) independent intensive variable y constant (at steady state).
(4) The remaining single independent extensive variable (sometimes called the throughput manipulator or basis) sets the throughput of the system, and changing it by a given factor k, automatically changes all extensive variables by the same factor k.

This has some important implications for the practical use of ratio control:

(5) If there are more than one extensive disturbance variable (which we cannot manipulate) then the assumption behind ratio control does not hold.
(6) Even for the case with a single extensive disturbance variable (for example, the feedrate), we must require that *all* the extensive manipulated variables are actively used for keeping constant intensive variables (including ratios). (For distillation, this means that we need to control two intensive variables).

Note that we must assume that there are no changes in the independent intensive variables x_i (disturbances, like feed composition), but in practice this is usually not a real limitation because these disturbances can be handled by an outer feedback controller that adjusts the intensive variable setpoint(s).

When does the scaling property hold? Similar to the use in thermodynamics, it holds for all equilibrium systems. Thus, the scaling property (and thus the use of ratio control) applies to many process units, including

- Mixers,
- equilibrium reactors,
- equilibrium distillation

3. RATIO CONTROL FOR DISTILLATION

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

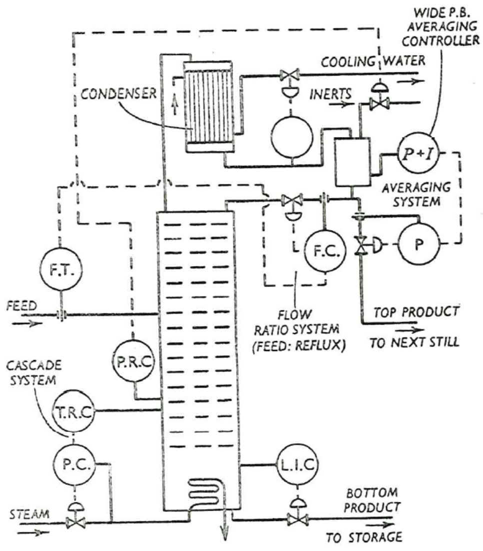

Fig. 18.11. Examples of the use of complex systems.

Fig. 4. Recommended ratio control of distillation column with fixed L/F and temperature (Young, 1955, page 323). (scheme B3)

For a typical two-product distillation column with a single feed and two products, there are with constant pressure, three 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 the scaling property, we get that with constant feed composition, all intensive variables in the column (including the product compositions) will remain constant if we keep two 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 ratio (L/D) and temperature (T somewhere in the column), see Figure 4.

As mentioned in the introduction, it has been common since the 1950s to recommend to use reflux to feed (L/F) ratio control for distillation. However, one needs to be careful about this. Indeed, already Young (1955) (footnote, p. 321) warns that it may not be advisable to keep L/F constant unless the heat to the reboiler (V) is also changed (as in Figures 1 and 4). The problem is that the heat input may reach its maximum value (and becomes constant) and the scaling property does not hold; recall point 6 above.

In fact, it turns out that with constant boilup (V), keeping the ratio L/F constant (see Figure 5) is worse than the simpler strategy of keeping reflux L constant. This is easy to explain by noting that for a distillation column, the most important operating parameter is the split ratio D/F, which from the material balance should be equal to amount 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 a simple structure V and

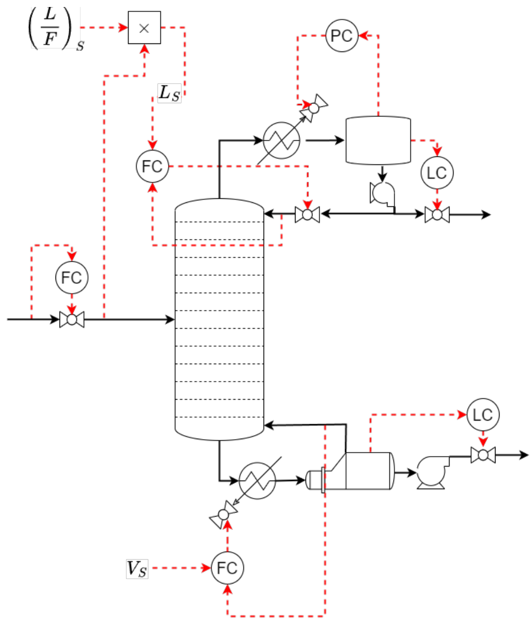

Fig. 5. Not recommended ratio control of distillation column with fixed L/F and V. (scheme B2)

L constant, all the increase in F will come out in the bottom (B), so D remains unchanged. This is obviously not good, because we want D to increase to keep D/F constant. However, with a control structure with constant V and reflux ratio L/F, things are even worse (especially for columns with a large L/F) because here also L increases which means that D will actually decrease, which is the opposite of what we wanted. A more mathematical proof is given in the Appendix.

For the steady-state and dynamic distillation simulations, we used Aspen Plus and Aspen Dynamics (files available at thesis home page of Bang (2024)). We consider the separation of a mixture of methanol and water. Column data are given in Table 1. The McCabe-Thiele diagram for the column is shown in Figure 6.

Parameter	Value
Number of 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
Reboiler type	Kettle
Reflux ratio L/D	1.013
Top product, x_D (water)	0.001
Bottom product, x_B (methanol)	0.001

Table 1. Typical steady-state operating parameters for the methanol-water distillation column

Dynamic simulations for a 10% increase in feedrate (F) are shown for the following four control structures

- 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)

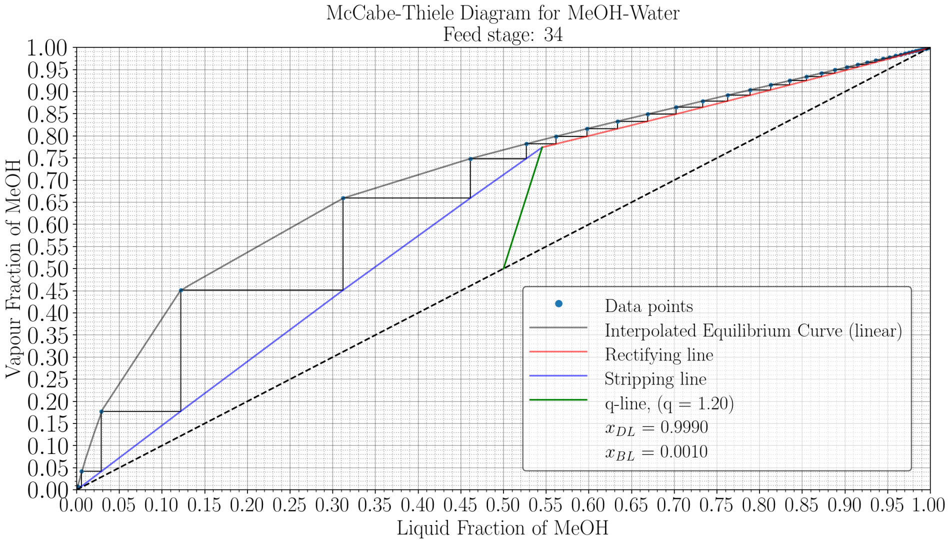

Fig. 6. McCabe diagram for methanol-water distillation column

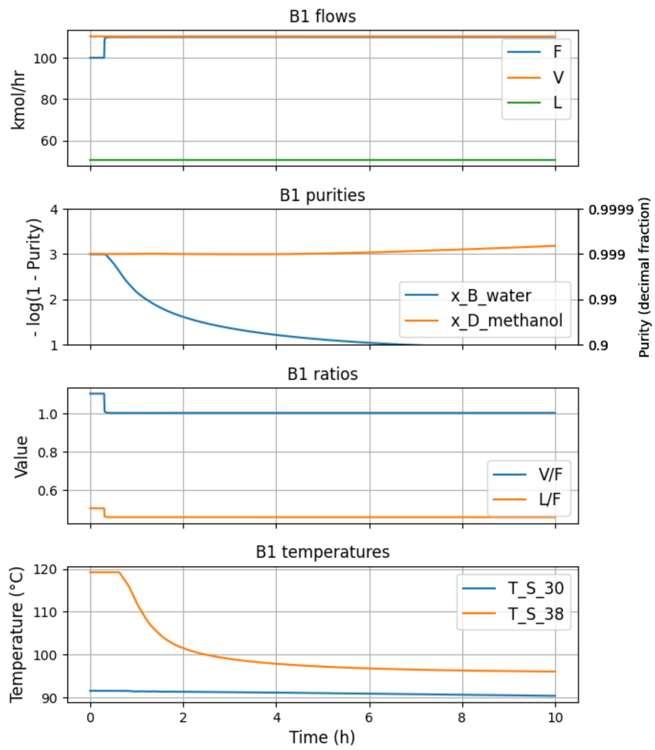

Fig. 7. Scheme B1: 10% feed flow disturbance with constant reflux (L) and boilup (V).

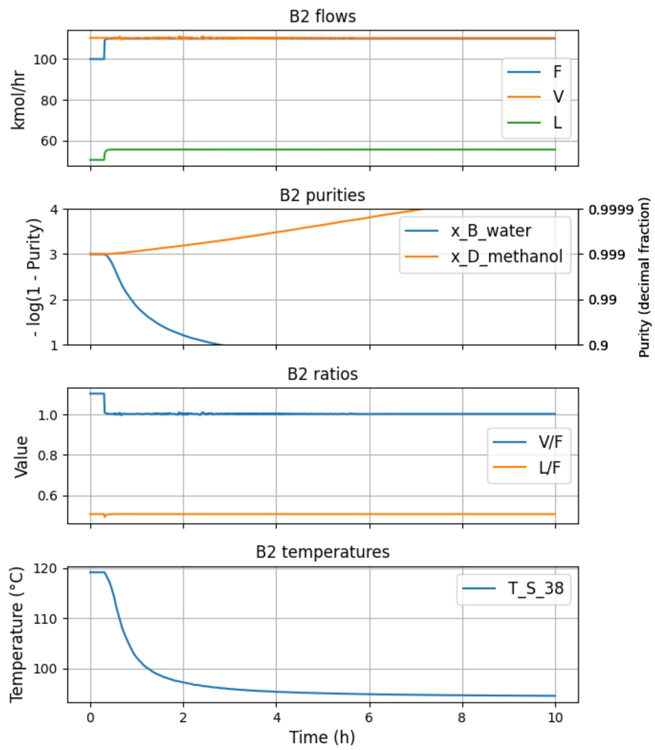

Fig. 8. Scheme B2: 10% feed flow disturbance with constant L/F and V.

• Scheme B4. Constant L/F and V/F (good)

The comments in parenthesis summarize the observed control performance in the corresponding dynamic simulations in Figures 7-10. The main control objective is to keep both product impurities at mole fraction 10^{-3} (corresponding to the value 3 in the "purities" plot).

As expected from the earlier discussion, scheme B2 with constant L/F and V is the worst. Schemes B3 and B4 both give constant properties (intensive variables, including compositions) but scheme B3 is a little better dynamically.

The 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 for the case when the feed is vapor.

Note that both schemes B3 and B4 become scheme B2 (the worst scheme, with L/F and V constant) 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 may be better because L rarely saturates and may be used to control temperature.

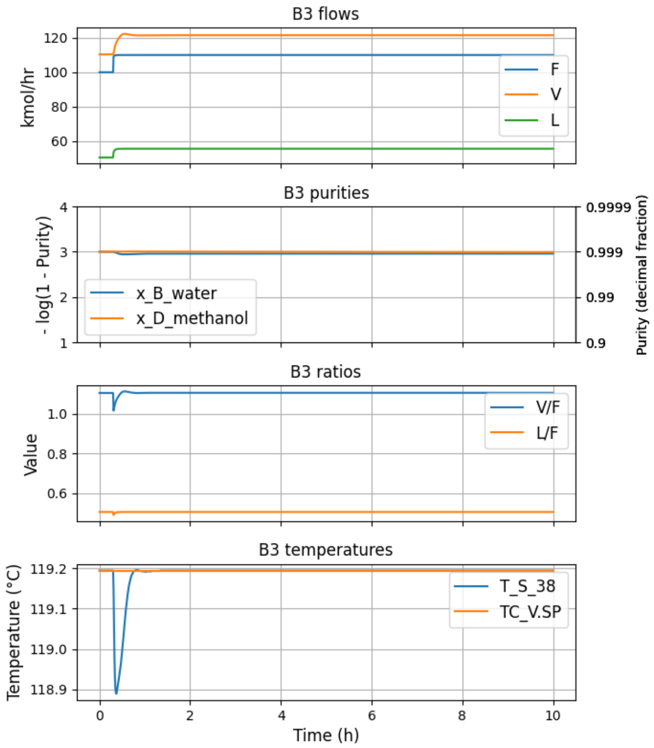

Fig. 9. Scheme B3: 10% feed flow disturbance with constant L/F and constant temperature on stage 38 (using V).

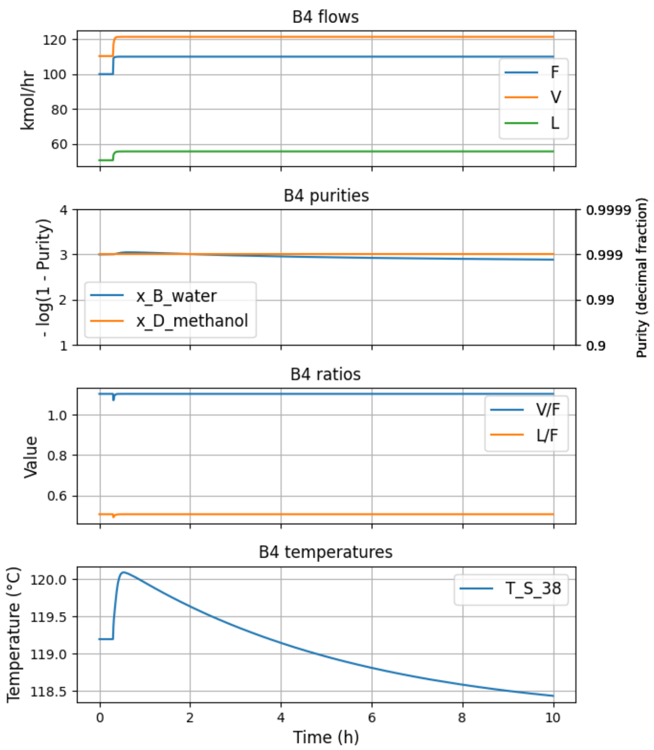

Fig. 10. Scheme B4: 10% feed flow disturbance with constant L/F and V/F.

Thus a structure with constant V/F and constant column temperature T will work well even if V saturates. The V/F-T structure 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 control of a gas-liquid separator with overrides for level and pressure (to handle saturation for bottom and top product rates) is shown in Figure 11. Note that both overrides use split parallel control (with setpoints H and L) go back to the feed F. The two MIN-selectors are needed to perform the overrides and also the transitions back again when the constraints are no longer active.

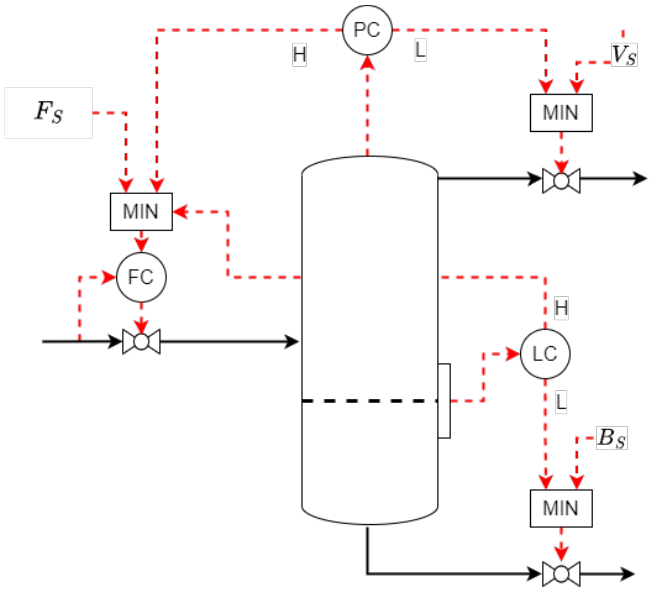

Fig. 11. Bidirectional control of a gas-liquid separator with overrides for pressure and level. The H-override for pressure is to handle cases where the vapor product rate saturates at maximum or is set by another controller (V_s).The H-override for level is to handle cases where the liquid bottom rate saturates at maximum or is set by another controller (B_s).

A corresponding simple bidirectional control scheme for a distillation column with two level control overrides (to handle saturation for D and B) is shown in Figure 12. Figure 13 shows an advanced bidirectional distillation scheme with four overrides. The overrides are for bottom level, top level, pressure and bottom composition and handle saturation (max) for B, D, cooling and heating V, respectively. The control in Figure 13 is based on the V/F-T control structure (mentioned at the end of Section 3), and the setpoints for the V/F ratio and the column temperature (T) are set by the bottom and top composition controllers, respectively. The override for bottom composition (with an H-setpoint) to the feedrate is to handle the case when the heat input (V) saturates at maximum.

As argued in Section 3 on ratio control, the inner fast temperature loop, which aims at stabilizing the column profile (and keeping the product split D/F constant on a

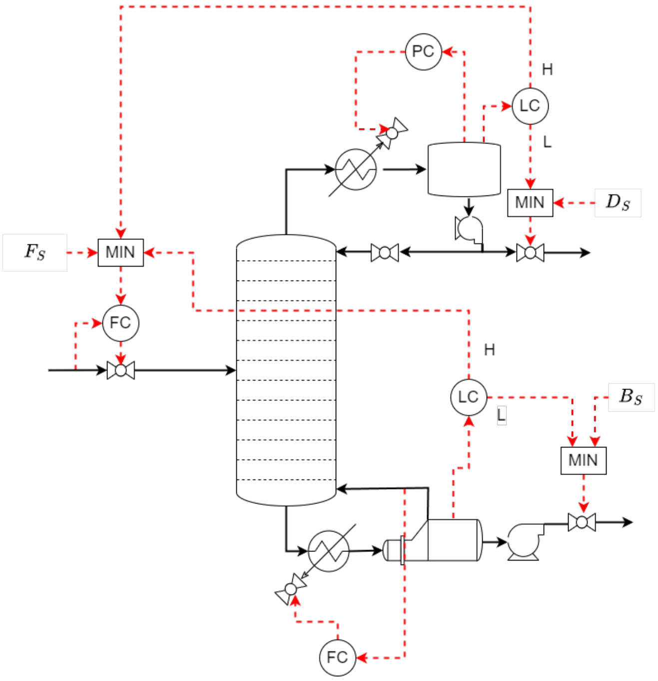

Fig. 12. Simple bidirectional control of top and bottom levels in distillation column (with reflux and boilup constant.)

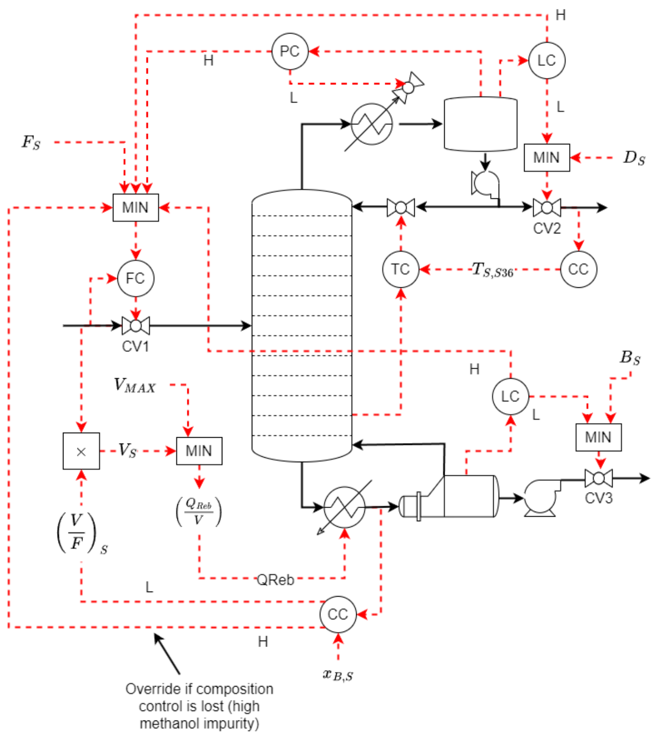

Fig. 13. Advanced bidirectional control for distillation column with four H-overrides for bottom (sump) level, top (condenser) level, pressure and bottom composition, which reduce the feedrate in case of saturation in bottoms flow (B), distillate (D), cooling and heating (V), respectively.

fast time scale), uses reflux (L) because the boilup (V) may saturate. The justification for using the V/F ratio (and not just V) as the manipulated variable for controlling bottom composition is that 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 on top (condenser) level.

For this particular mixture, an additional problem arises, because the thermodynamics of methanol-water are somewhat unusual with a pinch-like behavior in the entire top section, as seen from the McCabe-Thiele diagram in Figure 6. For this reason the temperature sensor for reflux (L) is not placed in the top part of the column, as would be expected, but rather in the bottom part (TC_{S36}, only 4 stages from the reboiler) which has a high gain from reflux to temperature. As expected, with the temperature sensor for reflux located in the bottom, there will be severe interactions with the bottom composition control loop, which forces the bottom composition loop to be very slow. However, attempts to place the temperature sensor in the top section were not successful because of the small gain. In any case, the main point of this paper is to show how to use bidirectional control for distillation, and not to find a non-interacting structure for composition control.

Time (h)	**Throughput manipulator (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/hr]	-0.963	-0.867 (-10%)
40	QReb [MCal/hr]	1.107	0.997 (-10%)

Table 2. Summary of constraint limit changes that result in activating H-overrides that result in moving the TPM in Figure 13.

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	2.0 m
$LC_{H,B}$	*	10	7200	2.0 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 3. Tuning parameters for the distillation column in Figure 13. 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.

The simulations in Figures 14 and 15 show clearly the effectiveness of the advanced bidirectional (override) control strategy, and that it handles changes in the TPM location, where different MVs may saturate (see details in Table 2). At t=0.5 h, the feed rate is increased, and this

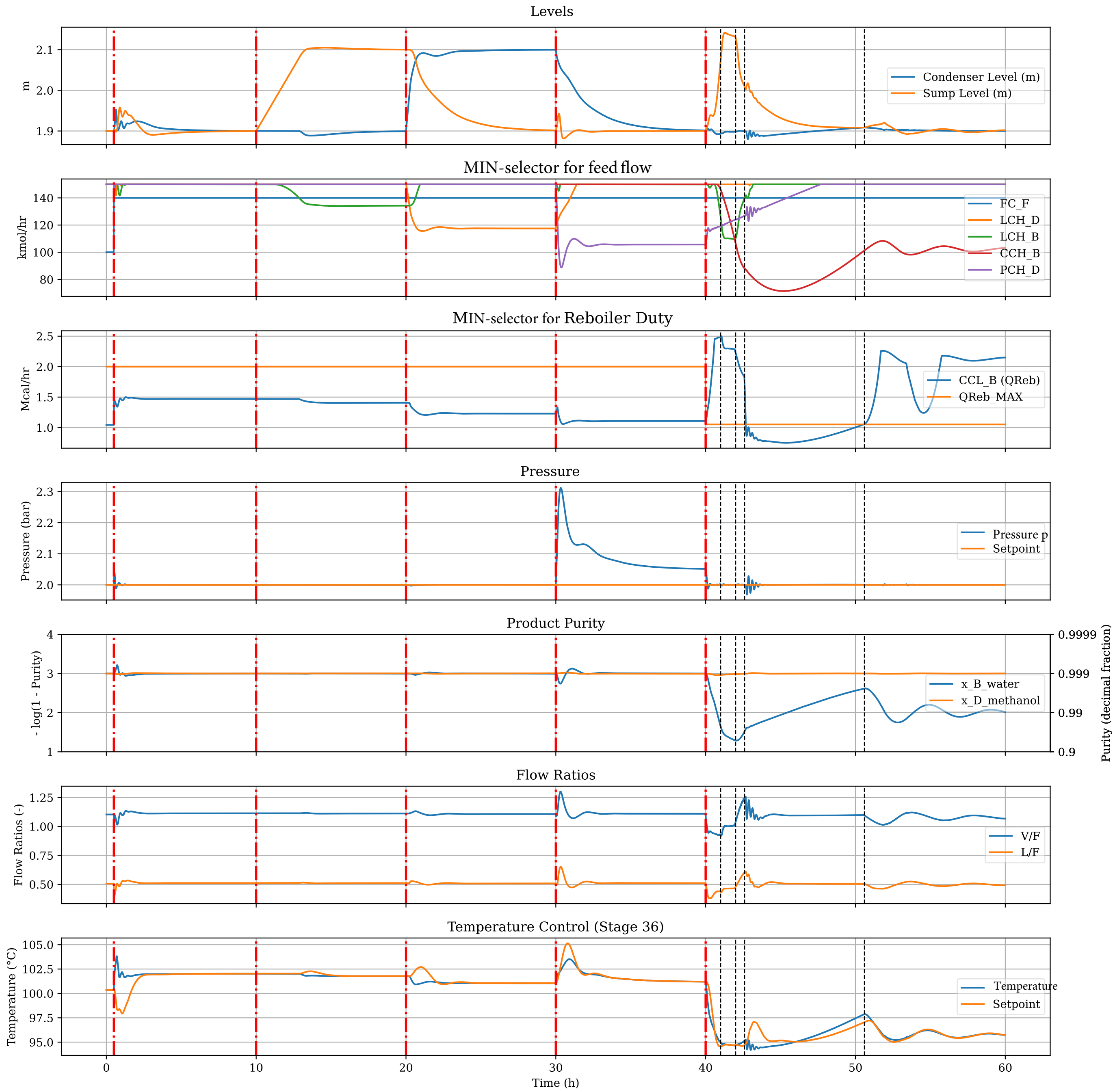

Fig. 14. Simulation results for the advanced bidirectional distillation column control scheme in Figure 13. Each vertical red line signifies one of the five events described in Table 2. The vertical black lines after t=40 h signifies the activation of an override controller.

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 TPM is seen from the subplot "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 (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 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).

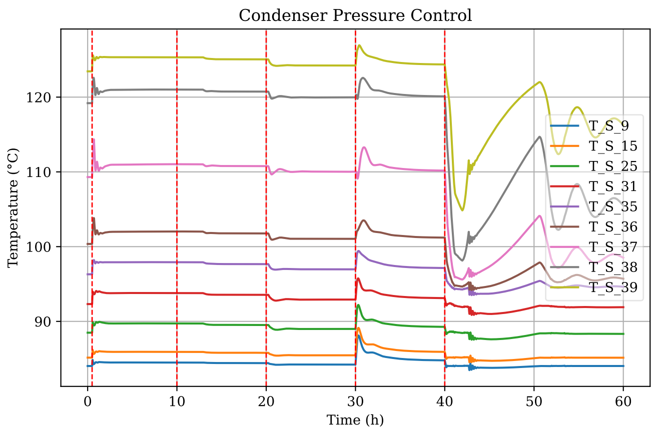

Fig. 15. Details of temperature evolution for the simulations in Figure 13.

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) 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 (V_{MAX}), 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 value (105 kmol/h). All of these measures reduces the time for the H-override composition controller ($CC_{H,B}$) to activate and reduce the feedrate. 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 14 and 15 demonstrate the robustness of the proposed control architecture.

The tuning of the PI controllers are given in Table 3 and were performed sequentially. The four level controllers are essentially P-controllers (with an integral time of 7200 s = 2 h). In addition to the tuning, there are several decision variables for the H-override controllers, for example, the threshold value (logical gate) and minimum input. However, the tunings could certainly be improved. In particular, the H-override controllers were not tuned systematically.

Discussion. Note that the flowsheets in Figures 12 and 13 are for the “LV-configuration” where reflux L and boilup V are used for composition control. In addition, note that Figure 13 is for the case where composition control of the top product is important, for example, because it is the valuable product so we want to avoid product give-away (e.g., Jacobsen and Skogestad (2011)). In this case, it may be optimal to overpurify the bottom product to avoid loss of valuable top product in the bottom. For example, if energy is free (or at least very cheap) we may put an unachievable setpoint ($x_{B,S} = 0$ for the light product), which will make V go to V_{MAX}. On the other hand, if tight composition control in the bottom is more important, then the temperature loop should be in the bottom using boilup (V). In this case, an additional override for reflux (L) to take over composition control in the bottom may be needed.

5. CONCLUSION

The paper demonstrates the power of the simple control architectures of ratio control and bidirectional control when applied to distillation columns. Intuitively, one may expect that ratio control (a special case of feedforward control) and bidirectional 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.

APPENDIX

Mathematical proof of desired changes in flowrates for distillation

To keep a constant split ratio D/F, we want for a change in feedrate F to have $d(D/F) = 0$, that is, $1/FdD–(D/F^2)dF = 0$ or

- Ideally: $dD = (D/F)dF$

If we assume constant molar flows and liquid feed fraction q_F ($q_f = 0$ is vapor, $q_F = 1$ is saturated liquid and $q_F > 1$ is subcooled liquid) then we have (noting that $dF = dD + dB$)

- Constant L and V: Get $dB = qFdF$ so: $dD = (1-q_F)dF$.
- Constant L/F and V: Get $dB = qFdF + dL$ where $dL = (L/F)dF$ so: $dD = (1 - q_F - L/F)dF$.

For the common case with saturated liquid feed ($q_F = 1$) we get

- Constant L and V: $dD = 0dF$
- Constant L/F and V: $dD = -(L/F)dF$

Ideally, we want to increase D, i.e., we want $dD = (D/F)dF$. In the first case (constant L) we get no change in D ($dD = 0$) which is not correct, but it is even worse in the second case (constant L/F) where because here the change in D is negative, which is the opposite of what we want; and especially for a column with a large L/F we get a very large decrease in D so the composition response with constant L/F and V will be very bad.

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