

## **MODULE III**

### **ISSUES IN PLANTWIDE CONTROL SYSTEM DESIGN**

The control structures for common unit operations as presented in the previous section may give the impression that developing effective control systems for a complete plant should be a piece of cake in that we simply put in the control structures for each of the individual unit-operations. As we will see in this module, there are unique challenges presented by material / energy recycle that make the plantwide control structure design problem much more challenging than simply putting in structures for each of the individual unit operations. In fact, there are many-many reasonable structures that will work to provide safe and stable operation on a given process. The economic performance of these different structures can however be significantly different. Industrial examples with prudent altering of the control structure resulting in the maximum achievable throughput for the same plant increasing by as high as 20-30% are part of industry folklore. What are the specific plantwide issues that must be considered and addressed in the design of such effective (including economics) plantwide control systems is the focus of this module.

For a firm grip on the plantwide control problem, we start from scratch covering degrees of freedom (control and steady state) and the tremendous flexibility that exists in the choice of the controlled variables (CVs) corresponding to these dofs as well the combinatorial complexity in the manipulated variables (MVs) used to regulate these CVs. We also discuss the snowball effect due to non-linearity caused by material recycle and the integrating nature of the component inventories in a recycle loop. We then discuss the design of the plantwide regulatory control system using the conventional CV-MV pairing approach and the more recent, Luyben pairing approach, along with an illustration on two toy-problems. Finally we bring in economic considerations and show how these considerations may require operating the plant at or close to equipment capacity constraints. We also discuss different ways of handling these constraints and their pros and cons in the plantwide context including illustrations on the two toy examples.

## Chapter 9: Control and Steady State Degrees of Freedom

### 9.1. Control Degrees of Freedom

The plantwide control system design problem can be considered as devising the “best” strategy for managing the available degrees of freedom (dof) in a process. From the operations perspective, a degree of freedom may quite simply be interpreted as having the freedom to make an adjustment, usually to a process / utility flow (a control valve opening). With no control system on a process, the operator is free to adjust the opening of the available independent control valves. These are referred to as the control degrees of freedom. By independent control valves, we imply respecting hydraulic fluid flow laws so that eg on a fluid flow pipe, only a

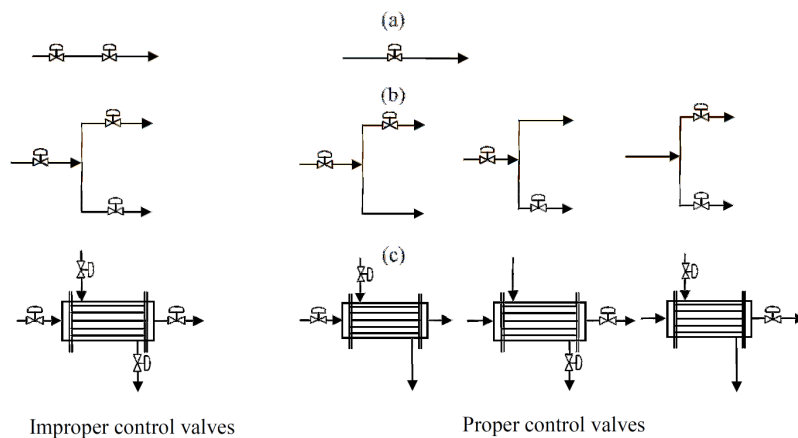


Figure 9.1. Examples of properly and improperly installed control valves

- (a) Flow through a pipe
- (b) Flow splitter
- (c) Process to process heat exchanger

single control valve is adjusted. Figure 9.1 provides illustrative examples of proper and improper installation of independent control valves.

How should adjustments be made to the independent control degrees of freedom (control valves). First and foremost, these adjustments must ensure safe and stable process operation. This requires using a control system for stabilization of potential instabilities and avoiding undesirable drifts in process variables. Reactor thermal runaway is an example potential instability. Process inventories such as liquid levels or gas pressure are examples of process variables that drift in the absence of proper regulation leading to potentially unsafe situations such as a tank running dry / overflowing or a rupture disc breaking open to release pressure. The control system for safe and stable process operation is referred to as the basic regulatory plantwide control system.

Given basic regulatory control that ensures safety, stability and acceptably small drifts, further adjustments may be made to any remaining valves or to the setpoints in the regulatory control system for ensuring the process is operated in the most profitable manner. This may

correspond to operating condition adjustments (valve positions or regulatory loop setpoints) to e.g. minimize steam consumption per kg product, maximize yield to the desired product, on-aim product quality with no product give-away, proper effluent discharge management etc.

## 9.2. Steady State Degrees of Freedom

For continuous chemical processes, it is the steady state at (around) which the process is being operated that determines the operating profit. Of all the control degrees of freedom, not all affect the steady state. This is illustrated for a very simple 'three-tanks-in-series' process in Figure 9.2. There are four control valves. Since liquid level in a tank is non-self regulatory (i.e. unless the inflow and outflow are exactly balanced, the level is either rising or receding), all three tank levels must be controlled to avoid large drifts in the levels. This would take away three control valves leaving one valve free. Let us say this free valve is at the process feed. We may then flow control the feed stream using this valve to set the fresh feed flow at the desired value. The level controllers then adjust the respective tank outlet valves as shown in Figure 9.2. The operator can adjust 4 setpoints (one fresh feed flow setpoint and three level setpoints). Of these the final steady state is determined only by the fresh feed flow setpoint and not by the choice of the level setpoints, which only has a dynamic effect. We therefore distinguish between the steady state operating degrees of freedom and the control degrees of freedom. The steady state operating degrees of freedom is the number of independent adjustments (to valve positions or regulatory setpoints) that affect the process steady state. For the simple example process, the

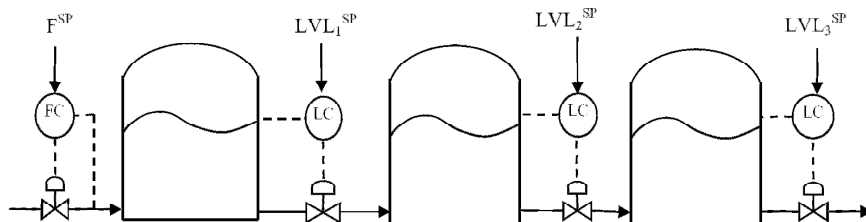


Figure 9.2. Three-tank-in-series process

steady state operating dof is 1, corresponding to the steady flow through the process, while the control dof is 4 corresponding to the number of independent control valves. Notice that the number of setpoints that the operator must input to the control system is 4, the same as the number of independent valves. Of these, the level setpoints have no steady state effect. Only the feed flow setpoint affects the steady state.

This then leads to a very simple procedure for calculating the steady state degrees of freedom for a process. We count the number of independent control valves and subtract the number of non-reactive surge levels as they have no effect on the steady state solution. If the inventory however is reactive, eg level in a liquid phase CSTR, it must not be subtracted (discounted) as the inventory (reactor holdup) affects the reaction extent (conversion) and hence the steady state solution. We also subtract any other variables (e.g. column pressures) that must be kept fixed at a given value for operational reasons to obtain the steady state operating degrees of freedom.

As an illustration, consider a simple distillation column. It has six valves (including feed). Two valves will get used for reflux drum and bottom sump level control. One valve would get used to control the column pressure. Usually the column pressure must be maintained at the

design value so that temperature inferential control can be applied. Also the column feed is not in our hands and is specified by an upstream process. Thus for a given feed and column pressure, the steady state operating dof of a simple distillation column is  $6 - 2 \text{ levels} - 1 \text{ column pressure} - 1 \text{ column feed} = 2$ . The operator is free to make 2 independent adjustments. These 2 independent adjustments may be made for maintaining 2 variables such the light key impurity in the bottoms and the heavy key impurity in the distillate.

In Figure 9.3, we show typical steady state dofs for simple unit operations with the implicit assumption that the feed to the unit is given (eg set by an upstream process). Figure 9.4 shows the steady state dof calculation for two example chemical processes. Notice the ease with which dofs can be calculated without having to worry about number of independent variables and number of independent constraints, counting which can befuddle even experienced engineers.

### 9.3. Degrees of Freedom, Controller Variables (CVs) and Control Structures

The steady state operating dofs are the number of independent adjustments an operator can make to a process that would affect the steady state solution of the process. Consider a simple distillation column. Given the column pressure and feed rate, the operator may choose to keep two appropriately chosen variables constant, corresponding to the two steady state dofs. The simplest option is to fix the reflux rate ( $L$ ) and the boilup ( $V$ ). This is equivalent to choosing  $L$  and  $V$  as the two column specifications. For changes in the feed rate / composition, the light key and heavy key impurity in respectively, the bottoms and the distillate, would show unacceptably large variation. To prevent excessive heavy key leakage down the bottoms, the operator may choose to adjust the boilup to maintain a sensitive stripping tray temperature ( $T_S$ ). To ensure that the light key leakage up the top is regulated, at least for changes in the feed flow, the operator may choose to maintain  $L$  in ratio with the column feed  $F$ . This is equivalent to  $T_S$  and  $L/F$  as the two column dof specifications. We may similarly have the operator maintaining  $T_S$  and  $T_R$ , a sensitive rectifying tray temperature, or alternatively the distillate heavy key mol fraction ( $x_{hk}^D$ ) and the bottoms light key mol fraction ( $x_{lk}^B$ ). Many other choices can be made for the 2 specification variables for simple distillation column. This example shows that there are several options for choosing the specification variable corresponding to steady state dofs.

From the discussion above, it is apparent that holding a particular variable constant implicitly assumes a control loop that manipulates an appropriate valve (or setpoint) in order to maintain the variable. Figure 9.5 shows example control structures corresponding to  $L$ - $V$ ,  $L/F$ - $T_S$ ,  $T_R$ - $T_S$  and  $x_{hk}^D$ - $x_{lk}^B$  as the specification (controlled) variables on a simple distillation column. In these structures a basic regulatory control structure is assumed where feed flow is controlled by the feed valve, column pressure is controlled by the condenser duty and the reflux drum and bottoms levels are controlled using respectively the distillate and bottoms.

Implicit in the pairings implemented in the structures shown in the Figure are some common sense principles. For fast level and pressure control, the manipulated variables are chosen 'local' to the concerned unit. Similarly, reflux is used to control a variable related to the rectifying section ( $T_R$  or  $x_{hk}^D$ ) and boilup is used to control a variable related to the stripping section ( $T_S$  or  $x_{lk}^B$ ). This pairing philosophy reflects the heuristic:

"Choose close by manipulated variables for controlling a process variable for a fast dynamic pairing".

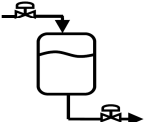
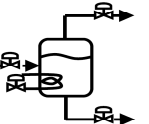
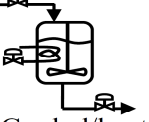
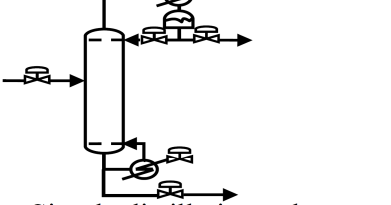
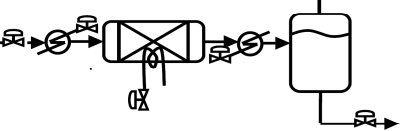
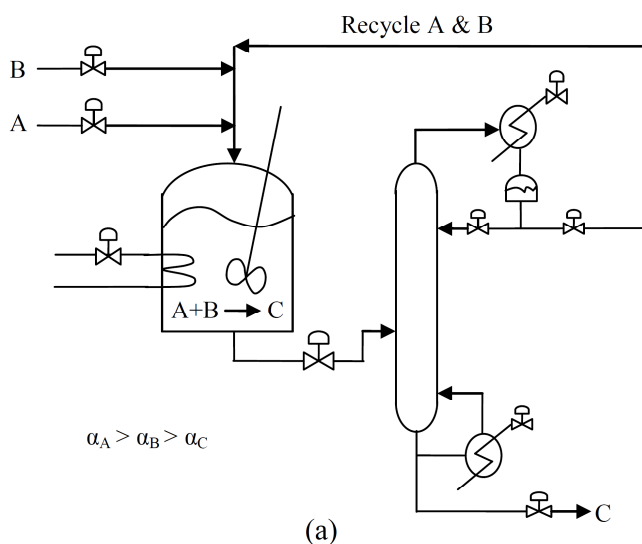
	Unit Operation	Independent control valves (Control dof)	Non-reactive surge levels	Other specification	Steady state dof	Typical steady state dof specification
(1)	 Liquid surge drum	2	1	1 (Throughput)	0	---
(2)	 Flash drum	4	1	2 (Throughput, Drum pressure)	1	Drum temperature
(3)	 Cooled/heated liquid CSTR	3	0	1	2	Reactor temperature Reactor hold up
(4)	 Simple distillation column	6	2	2 (Throughput, Column pressure)	2	Reflux & boil up or distillate heavy key impurity & bottoms light key impurity
(5)	 Gas phase packed bed reaction section	6	1	2 (Throughput, Flash drum pressure)	3	PBR inlet temperature, PBR hot spot/outlet temperature, Cooler outlet temperature

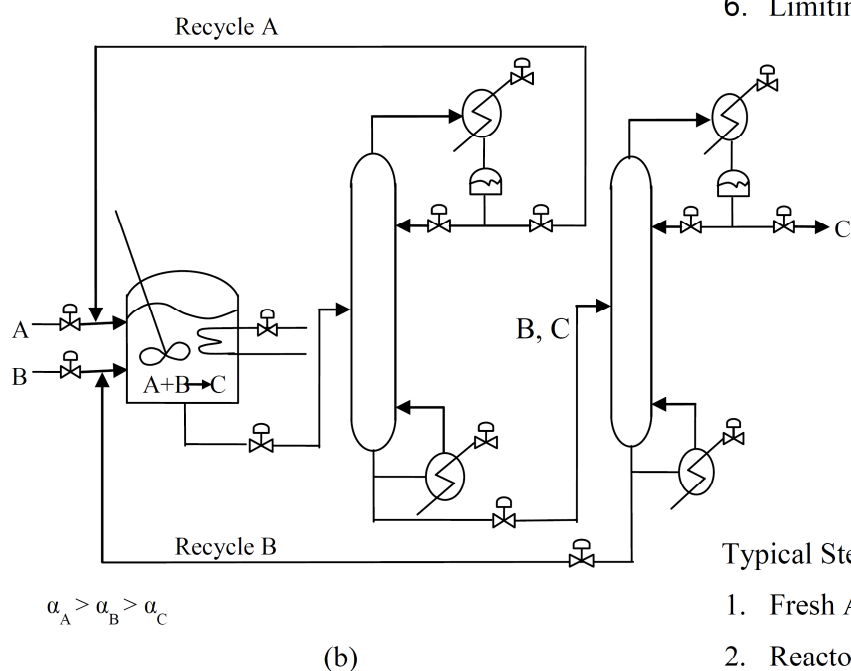
Figure 9.3. Illustration of control and steady state dofs for some typical unit operation



Control dofs:	9
Non-reactive surge levels:	2
Given column pressure:	1
Steady state dofs:	<u>6</u>

#### Typical Steady State Specifications

1. Fresh A feed rate (throughput)
2. Reactor temperature
3. Reactor level
4. C impurity in recycle stream
5. B impurity in product C
6. Limiting reactant conc. in reactor



#### Typical Steady State Specifications

1. Fresh A (or B) feed rate
2. Reactor temperature
3. Reactor level
4. C impurity in recycle A
5. A impurity in column 1 bottom
6. B impurity in product C
7. C impurity in recycle B
8. Total (recycle + fresh) B (or A)

Control dofs:	14
Non-reactive surge levels:	4
Given column pressures:	2
Steady state dofs:	<u>6</u>

Figure 9.4. Illustration of dof analysis for complete chemical plants

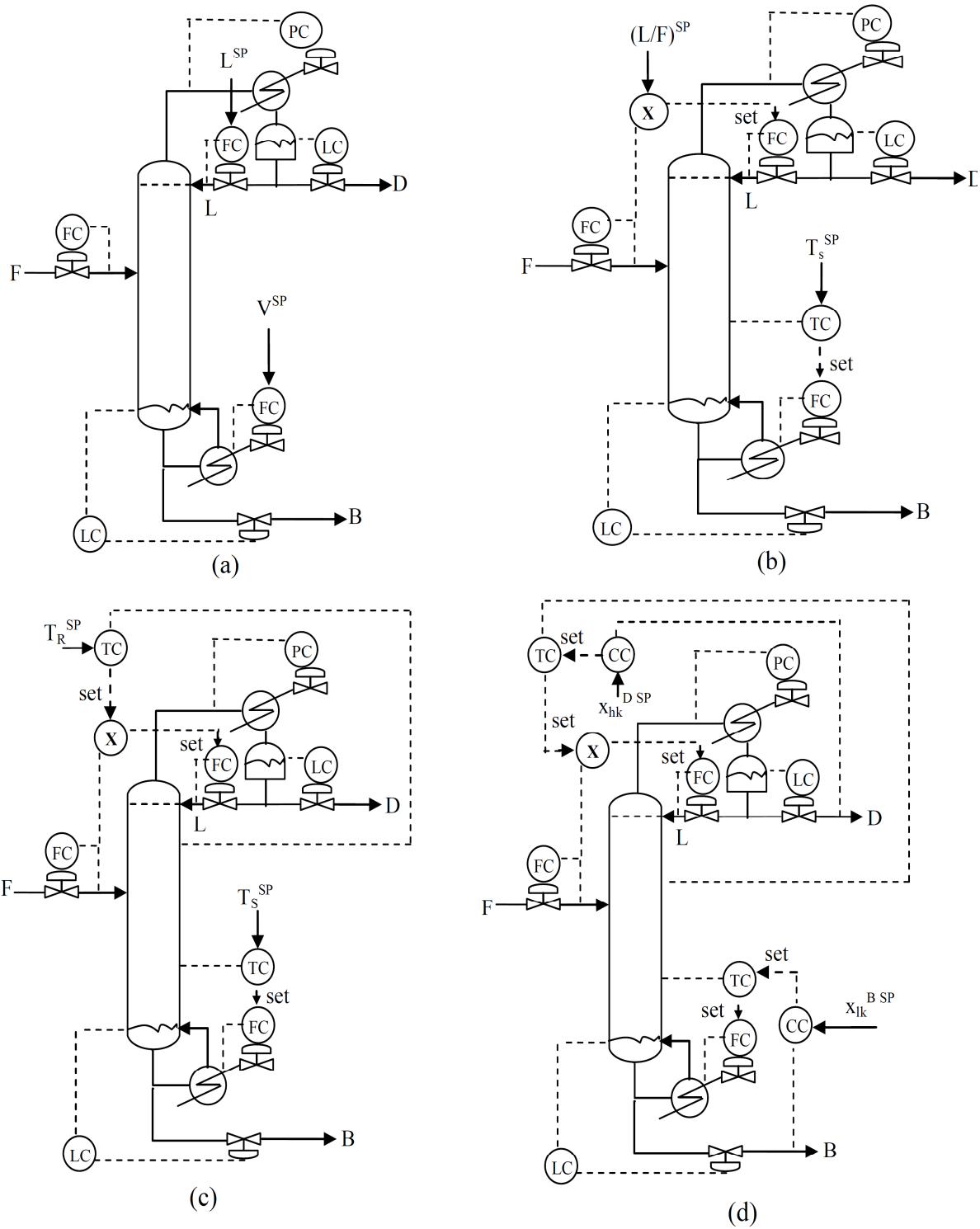


Figure 9.5. Alternative CV<sub>s</sub> corresponding to steady state dof<sub>s</sub> on a simple distillation column.  
 (a) L-V (b) L/F-T<sub>S</sub> (c) T<sub>R</sub>-T<sub>S</sub> (d) x<sub>hk</sub><sup>D</sup> - x<sub>lk</sub><sup>B</sup>

If we let go of the "close-by" pairing philosophy, for the same set of CVs, several alternative pairings can be proposed. Giving up close-by pairing on a unit would usually be due to plantwide control considerations that require tighter control of a particular plant subsection. For example, let us say the distillate from the column feeds the reaction section of a plant, where a highly exothermic reaction occurs. We would like to hold the flow to the reactor section constant to prevent propagation of transients to this section as it is hard to stabilize and variability in the reaction section upsets the downstream product separation section. So now, the distillate must be flow controlled to eliminate flow transients to the reaction section. This flow setpoint then sets the flow through the column, instead of the column feed. Since distillate is fixed, reflux drum level gets controlled using the reflux. The bottoms level is controlled as before using the bottoms. Since it is important to have tight impurity control in the distillate (which feeds a reactor), we use boil-up to control a rectifying tray temperature, as a change in boilup has an almost immediate effect on tray temperature, unlike reflux which has a slower effect particularly if the control tray is further down from the top. This pairing would give tighter distillate impurity regulation. The stripping tray temperature then gets controlled using the column feed. Figure 9.6 shows four alternative pairings for  $T_R$ - $T_S$  as the CVs on a column. These structures differ particularly in the location where the flow through the column, also referred to as the throughput, is set. Which structure should get implemented would depend on the specific plantwide context. Even as we have not said much about plantwide control considerations, the point of the whole exercise is to show that even for a simple distillation column with 2 steady state dofs, there exists tremendous flexibility in the control structure that can be implemented on it due to the choice of the specification variable corresponding to the steady state dofs as well as the pairings for the CVs (including regulatory level and pressure loops).

How do we go about systematically choosing the CVs and the corresponding pairings is like piecing a puzzle together. In what follows, we look at different ways of piecing together this puzzle. The first step, as evident in what has already been discussed previously, is to count the number of control and steady state degrees of freedom. The next step is to tabulate the different control objectives and appropriate controlled variables (CVs) for those objectives. All control objectives regulate some process inventory, inventory being interpreted in its most general sense to include total material, phase, component and energy contained in a process unit and the overall process. The regulatory control system is required to ensure (In – Out + Generation) of the inventories in a unit and the overall process is zero so that accumulation is forced to zero to ensure unit specific / plantwide drifts are avoided / mitigated.

The number of CVs are the same as the number of control degrees of freedom and would encompass all inventory regulation objectives. Of these, pure surge capacities have no steady state impact and are therefore economically not relevant. The level of component inventories in recycle loops and product / purge streams on the other hand usually impact the steady state plant economics significantly. The reactor operating conditions (temperature and composition) also are usually important as the single-pass conversion and selectivity determine the cost for recycling unreacted reactants and side-product processing cost.



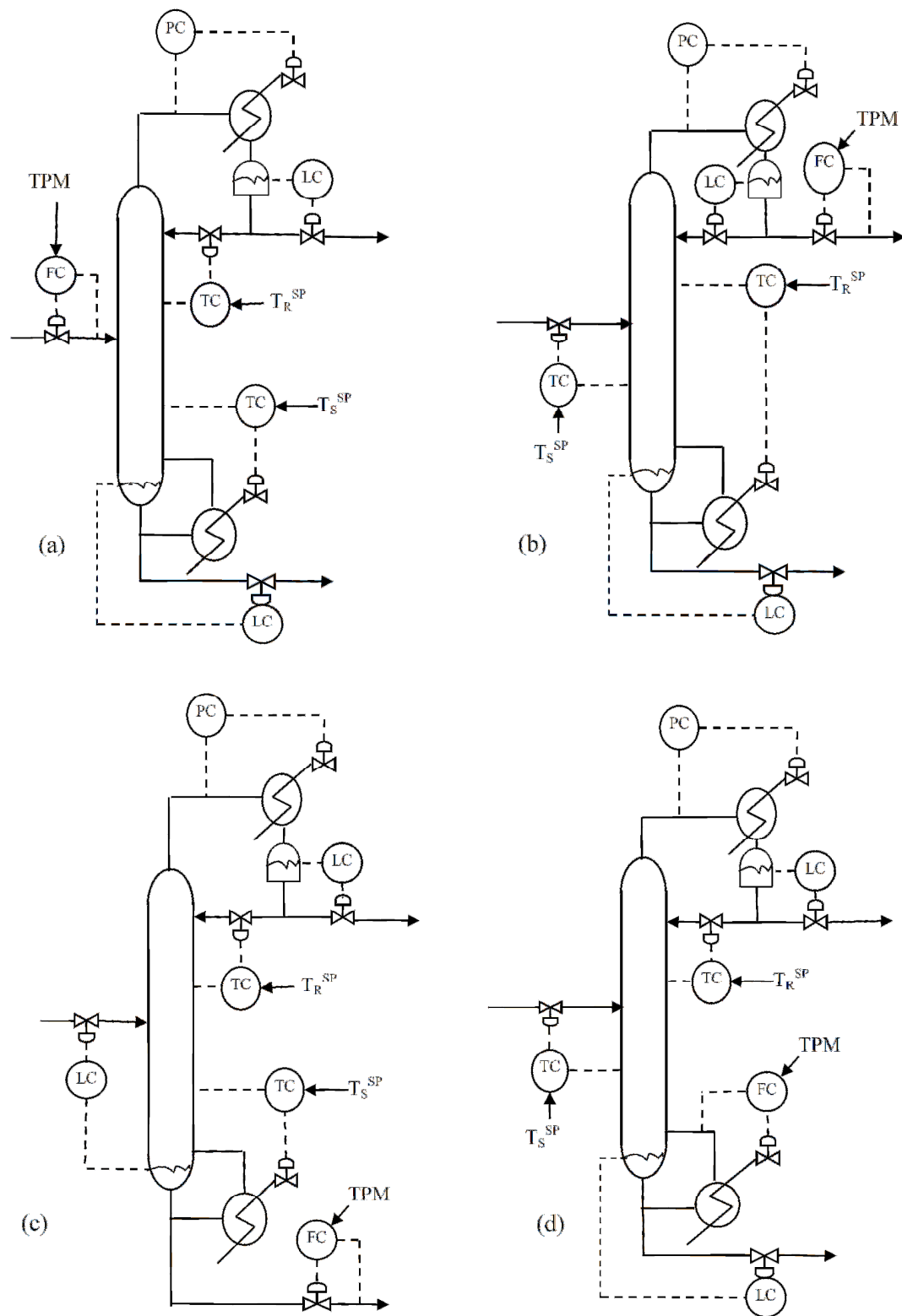


Figure 9.6. Alternative pairings (structure) for holding  $T_R$ - $T_S$  as the two steady state dof CVs as a simple distillation column

## 9.4. Control Objectives and Choice of CVs

Given a set of control objectives and corresponding CVs plus the prioritization of the CVs, it is relatively straightforward to devise the control loop pairings. How does one go about systematically determining the control objectives and corresponding CVs. To the experienced engineer, control objectives and corresponding CVs for a process are usually evident. To the novice however, this is usually not very clear. In the following we attempt to provide a basic framework to help figure out the control objectives and appropriate CVs.

The control system on a continuous chemical process with material and energy integration may be viewed as an automatic mechanism for ensuring that all process inventories are regulated at safe / optimal levels and not allowed to drift, regardless of process disturbances such as changes in the process throughput, ambient conditions, equipment characteristics etc. All the CVs directly/indirectly reflect process inventories; e.g. level reflects liquid inventory, pressure reflects gas/vapor inventory, temperature reflects energy inventory and composition reflects component inventory (inferential measurements such as column tray temperature or a recycle flow or an appropriate separator level also indirectly reflect component inventory). Since inventories are prone to large drifts (accumulation/depletion) unless regulated, the plantwide control system attempts to maintain them at desired values for economic reasons or at the very least, within an acceptable band (e.g. surge drum levels) to avoid unsafe operating conditions. From the economic standpoint, typically component inventory levels in recycle and product/discharge streams have a large impact on the steady state operating profit so that these should be controlled tightly. On the other hand, surge drum levels that are part of the material balance control system have no effect on the process steady state.

As a starting point, let us take a liquid tank with a liquid stream in and a liquid stream out as a very simple example. If both the inlet and outlet control valves are flow controlled as shown in Figure 9.7(a), the control structure is fundamentally flawed as it violates the overall material balance constraint. Two flows are being independently set and any mismatch in the setpoints would necessarily imply the liquid inventory in the tank (indicated by a level sensor) either builds up (*inflow* > *outflow*) or depletes (*inflow* < *outflow*). The tank is then guaranteed to run dry or over flow. In other words the implemented control system is guaranteed to fail.

The novice may argue that to satisfy the material balance constraint, both the setpoints can be set equal. That still does not solve the basic problem as a mismatch in the two tank flows would any way occur since sensors are never 100% accurate, the slightest of biases implying a slow build-up / depletion in the tank level. The basic issue is that the liquid inventory in the tank is non-self regulatory and must therefore be regulated. We need to measure (or estimate) the liquid inventory and adjust one of the flows to ensure that the inventory is maintained within an acceptable band. The other flow is set independently by the operator or an upstream / downstream process. A direct measure of the liquid inventory inside the tank is its level. Figure 9.7(b-c) shows two workable control configurations that respect the material balance constraint by controlling the tank level.

Even as the above is a very trivial example, treating a complex process with several units and recycles as a tank and questioning if the implemented control system ensures all process inventories (material, phase, component or energy) on each of the individual units as well as the overall process are regulated and do not drift would reveal if the control system is workable or

not. We note that routine level, pressure, temperature and flow measurements that indicate appropriate inventory levels are usually self evident.

The control structures on individual unit operations that have already been discussed in previous chapters may be interpreted as regulating inventories. For example, in dual ended temperature inferential LV control structure of a simple distillation column, the condenser duty regulates the column pressure (total vapor inventory), the distillate flow regulates the reflux drum level (reflux drum liquid inventory), the bottoms flow regulates the sump level (sump liquid inventory), the reflux rate is adjusted to maintain a sensitive rectification section temperature to regulate the heavy key leakage in the distillate (component inventory) and the boilup is adjusted to maintain a sensitive stripping tray temperature to regulate the light key leakage down the bottoms (component inventory). Each control loop on the column fixes (regulates) a process inventory. Of these, while the two levels have no economic significance, the light key and heavy impurity leakage levels significantly affect the column energy consumption and are therefore economically important. The interpretation can be easily extended to control structures on other unit operations studied earlier.

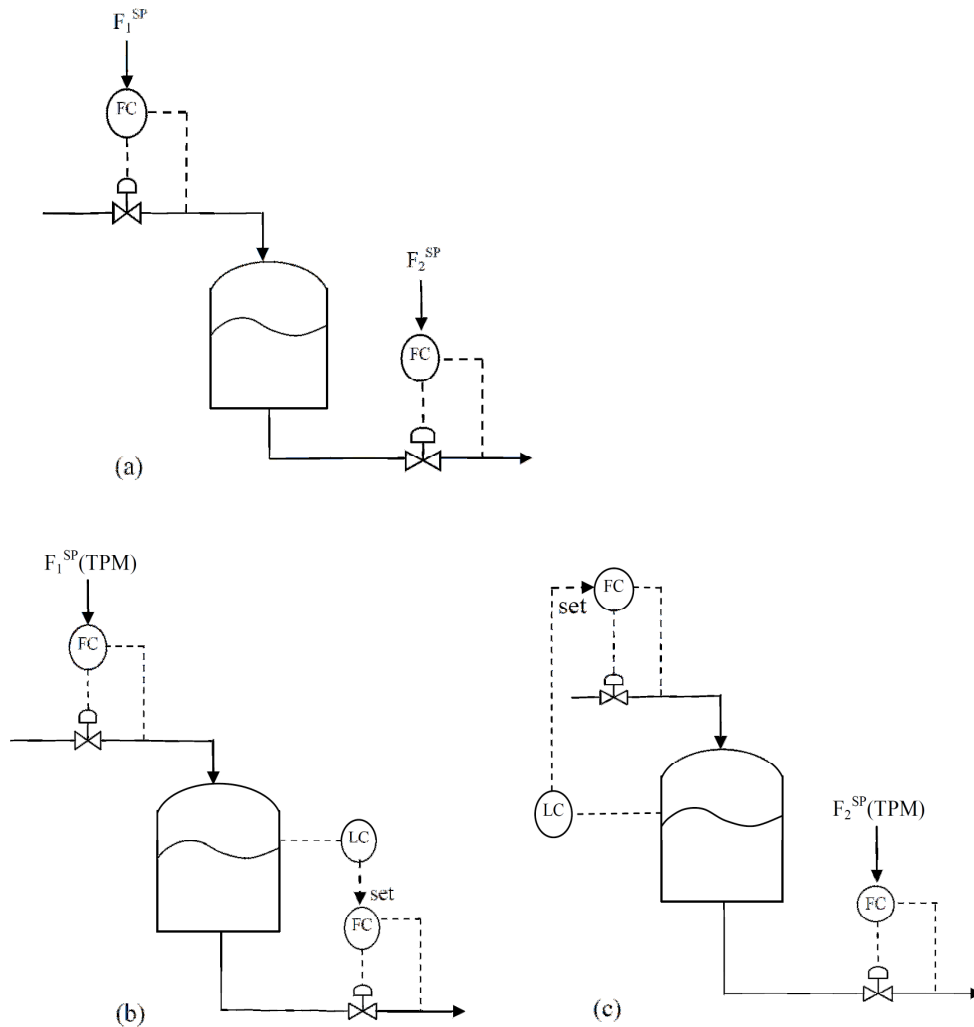


Figure 9.7. Material balance control on a liquid surge drum  
 (a) Unacceptable control structure  
 (b) & (c) Acceptable control structure

## 9.5. Illustration of Control Objectives and CVs for Example Processes

We are now ready to illustrate control objectives and corresponding CVs for a complete plant. Let us consider the process flowsheet in Figure 9.4(a). It has 9 control dofs and these valves can be used for regulating 9 objectives. On the reactor, the total material hold-up and energy hold-up must be regulated. The reactor level and temperature are appropriate CVs for the same (1<sup>st</sup> – 2<sup>nd</sup> CVs). On the distillation column, the liquid holdup in the reflux drum and bottom sump must be regulated. Also, the vapor hold-up in the column must be regulated. The reflux drum and sump levels along with the column pressure are appropriate CVs for these inventories (3<sup>rd</sup>-5<sup>th</sup> CVs). We also need to regulate the product C leakage up the top and the B impurity leakage down the bottoms. A sensitive stripping tray temperature is a good inferential measure of the latter (6<sup>th</sup> CV). Holding the reflux in ratio with the column feed would provide loose but adequate regulation of the C leakage in the recycle stream (7<sup>th</sup> CV).

The remaining 2 control objectives are more subtle. By the design of the process, the recycle stream would contain significant amounts of both the reactants, A and B, with small amounts of C. If we look at the overall material balance across the entire plant, 1 mol A would react with exactly 1 mol of B. The slightest excess of fresh A (or fresh B) is not allowed to leak in the product stream due to a stringent product purity constraint and must necessarily accumulate in the recycle loop. Unless the fresh feeds are balanced exactly as dictated by the reaction stoichiometry, the recycle loop would slowly but surely get filled up with the excess reactant (A or B). The recycle rate and its excess reactant composition would then increase. This slow drift of component inventories inside the recycle loop is referred to as the snowball effect. We need to regulate the component inventory of both the reactants in the recycle loop to ensure stoichiometric feed balancing. This would ensure the recycle rate and its composition does not drift. Since the reactor is inside the recycle loop, one may hold composition of a reactant (usually the limiting reactant) to regulate its inventory (8<sup>th</sup> CV) and the total flow to the reactor to regulate the inventory of the other component (9<sup>th</sup> CV). Note that the reactor temperature and composition indirectly sets the production rate inside the reactor through the kinetics. We may change either of these to bring about a change in process production rate.

As another illustration of control objectives, consider the process in Figure 9.4(b). The process control dof is 14. The reactor material and energy inventories are reflected by reactor level and temperature (1<sup>st</sup> – 2<sup>nd</sup> CVs). On the first column, the liquid and vapor inventories are reflected by the reflux drum and sump levels and column pressure (3<sup>rd</sup> - 5<sup>th</sup> CVs). The column prevents C (heavy key) leakage up the top and A (light key) leakage down the bottoms. Any A that leaks down the bottoms would necessarily end up in the product C stream. It must therefore be tightly regulated and a sensitive stripping section tray temperature is a good inferential measure of the same (6<sup>th</sup> CV). Since the first column distillate is a recycle stream, loose regulation of the C impurity in it is acceptable. Holding the column reflux to feed ratio ( $L_1/F_1$ ) constant should suffice (7<sup>th</sup> CV). On the second column, we again have the reflux drum / bottom sump levels and pressure as measures of liquid and vapor inventories (8<sup>th</sup> – 10<sup>th</sup> CVs). The column prevents B (heavy key) leakage up the top and C (light key) leakage down the bottoms. Tight regulation of the B impurity in the product stream (component inventory) is desirable and a sensitive rectifying tray temperature is a good inferential measure of the same (11<sup>th</sup> CV). Since the bottoms is a recycle stream, loose regulation of the C impurity in it is acceptable. Assuming

that boilup is paired for tight control of rectifying tray temperature for tight product quality control, we may hold the reflux-to-feed ratio ( $L_2/B_1$ ) to indirectly achieve the same (12<sup>th</sup> CV).

We now consider the stoichiometric balancing of the two fresh feeds to the process. By the design of the process, if an excess of fresh A (fresh B) is being fed, it would accumulate in the A (B) recycle stream. The total (fresh + recycle) A (B) rate would then increase. This total rate to the reactor then indirectly reflects the A (B) inventory in the process. We may then choose the total (recycle + fresh) A to the reactor and total (recycle + fresh) B to the reactor as very convenient measures of the component inventories in the recycle loops (13<sup>th</sup> – 14<sup>th</sup> CVs). As in the previous example, the total rate of either reactant to the reactor or the reactor temperature may be adjusted to bring about a change in the process production rate.

Table 9.1 summarizes the regulatory control objectives and corresponding CVs for the two example processes. The relationship of the control objectives with ensuring unit specific and plantwide material and energy balances are evident in the objectives. Comments are also provided to highlight their economic / regulatory significance.

## 9.6. Snowball Effect

From the discussion above, it is evident that while the inventories that require regulation on a specific unit are quite self-evident, figuring out recycle component inventories that require regulation is subtler and requires some thought with respect to guaranteeing that the overall material balance around the plant for all the components is satisfied. Material recycle introduces high non-linearity into the process with the recycle rates being highly sensitive to small changes in the fresh feed flow(s). This is referred to as the snowball effect.

If we consider the example process in Figure 9.4(a), its steady state dof is 6. The reactor level and temperature and the light key / heavy key leakage in the bottoms / distillate of the column specify four of these dofs. Let us say that we arbitrarily choose the two fresh feed rates as specifications for the remaining 2 steady state dofs. If we try and converge the flowsheet using a commercial simulator, we will find that if the two fresh feeds are specified to be even slightly different, the recycle tear does not converge and keeps on blowing up. This is because the reaction stoichiometry and nearly pure product constraint implies the reactant fed in slight excess has no way out of the process and therefore must necessarily build up in the recycle loop. The sensitivity of the recycle to even the slightest of mismatch between the two fresh feeds is then infinity. If we purge a very small fraction of the recycle stream, the sensitivity of the recycle stream rate to small changes in the fresh feed rates would still be very high, though not infinity. This is the snowball effect.

The choice of the specification variables for the two dofs is not appropriate as the two flows are related by overall process material balance. For robust convergence, a better specification is specifying the total flow rate to the reactor and its A (or B) mol fraction. Both the fresh feeds then get calculated to satisfy these two specifications.

From the operations perspective, if the fresh feed(s) are specified (ie flow controlled), the high sensitivity of the recycle rates to the fresh feeds would cause large swings in the recycle streams and all the equipment in the recycle loop would be subjected to large plantwide transients for small changes in the fresh feed(s). To avoid these large swings, it is better to hold appropriate component inventories in the recycle loop by manipulating the fresh feed(s). The fresh feed(s) are then fed as make-up streams and only as much is fed as gets consumed. Since

the reactor is always inside the material recycle loop, a common industrial practice is to hold the total reactant component feed (fresh + recycle) to the reactor constant by adjusting the corresponding fresh feed. In cases where the recycle stream is nearly pure reactant, the corresponding fresh feed may be adjusted to hold the total (recycle + fresh) flow constant. In cases where the recycle stream is a mixture of reactants, appropriate composition(s) inside the reactor and total flow to the reactor are held constant by manipulating the fresh feeds.

The basic idea of feeding fresh feeds to hold appropriate reactor conditions constant achieves two objectives. It ensures the component inventories in the recycle loops are properly managed. Also, by maintaining the reactor operating conditions (flow and composition) constant, robust stabilization of the most non-linear unit operation in the process is achieved mitigating the transients propagated to the downstream separation section.

Table 9.1. Regulatory objectives and CVs for the two example processes

<i>SNo</i>	<i>Regulatory objective</i>	<i>CV</i>	<i>Significance</i>
Single column recycle process			
1	Reactor liquid inventory	Reactor level	Closes reactor MB*. Affects conversion and separation load.
2	Reactor energy inventory	Reactor temperature	Closes reactor EB**. Affects conversion and separation load.
3	Column reflux drum liquid inventory	Reflux drum level	Closes reflux drum MB.
4	Column sump liquid inventory	Sump level	Closes sump MB.
5	Column vapor inventory	Column pressure	Closes column EB.
6	Distillate hk <sup>&amp;</sup> (C) leakage	Reflux to feed ratio	Closes the lk/hk balance on the column. Affects column steam consumption. Bottoms B leakage fixed by min product quality. Too much distillate C leakage dilutes reactor reducing conversion.
7	Bottoms lk <sup>%</sup> (B) leakage	Stripping tray temp	
8	Component B circulating in recycle	Reactor B mol fraction	Fixes recycle stream conditions and hence affects column steam consumption.
9	Component A circulating in recycle	Total feed to reactor	
Two-column recycle process			
1-2	Reactor liquid and energy inventory	Reactor level and temperature	Closes reactor MB and EB. Affects conversion and downstream separation load.
3-6	Reflux drum/sump liquid inventories	Column reflux drum and sump levels	Closes reflux drum/sump MBs
7-8	Vapor inventory in columns	Column pressures	Closes column EBs
9	Column 1 distillate C (hk) leakage	Reflux to col feed ratio	Closes the lk/hk balance on the columns. Affects reboiler steam consumption. Too much C leakage in recycle streams dilute reactor reducing conversion. Col1 bottoms A leakage and Col2 distillate B leakage set by max product impurity specification
10	Column 2 bottoms C (lk) leakage	Reflux to col feed ratio	
11	Column 1 bottoms A (hk) leakage	Stripping tray temp	
12	Column 2 distillate B (hk) leakage	Rectifying tray temp	
13	Component A circulating in plant	Total <sup>#</sup> A to reactor	Fixes recycle stream conditions and hence affects column steam consumption.
14	Component B circulating in plant	Total <sup>#</sup> B to reactor	

\*: material balance; \*\*: energy balance; &amp;: heavy key; %: light key; #: recycle plus fresh feed

## Chapter 10. The Pairing Issue: Selection of MVs for CVs

Given a set of inventory regulation control objectives and corresponding CVs, the next step is to select the manipulated variable (MV) pairing for each of the CVs. To select pairings for the CVs, they must be prioritized with the pairing for the highest priority CV being selected first followed by the pairing for next one and so on so forth. Different prioritizations would lead to different pairings and hence different control structures.

### 10.1. Conventional Pairing Approach

The conventional approach to designing the loop pairings is to first choose the process variable that is adjusted for setting the throughput. The setpoint corresponding to that process variable control loop is referred to as the throughput manipulator (TPM). Conventionally, the throughput manipulator is chosen at a fresh feed to the process. Other TPM locations are possible and include the product stream flow for on-demand process operation, where the demand from a customer must be immediately met; an intermediate process stream flow for mitigating transients to the connected unit; directly setting reactor temperature or limiting reactant concentration in a process with a reactor etc.

With the TPM in place, local inventory loops on each of the units are then put in place to establish total material balance / energy balance control. By local, we mean that the MV for controlling the inventory is local to the unit containing the inventory. This is illustrated in Figure 10.1 for the 'three tanks in series process', where the throughput may be set at any of the four process streams. The tank level controllers upstream of the TPM (set flow) are then naturally oriented opposite to the process flow while the level controllers downstream of the TPM are oriented in the direction of process flow. The upstream level controllers act to supply the set flow while the downstream level controlled act to process the set flow. The total material balance control structure thus radiates outwards from the TPM. Local loops for energy balance control would usually include temperature control of an exothermic reactor using reactor cooling duty stabilizing the most non-linear unit in the plant.

With the basic material balance / energy balance control pairings in place, the pairings for the remaining CVs are chosen from the remaining valves. These involve loops for regulating component inventories and are usually economically important. In cases where the open loop response of the CV is sluggish, an appropriate cascade arrangement is implemented with a slave controller holding a faster secondary variable and the master controller holding the primary variable by adjusting the slave loop's setpoint.

### 10.2. Luyben's Pairing Approach

In the first significant departure from the conventional pairing approach, Luyben et al.<sup>14</sup> insightfully noted that since non-reactive surge inventories have no steady state economic impact, material balance control loops should have lower prioritization so that the best pairings get implemented for the tightest control of economically important CVs. Their prioritization hierarchy thus first fixes the TPM and energy balance control, then establishes loops for economically important objectives (quality, safety, effluent discharge etc) and finally pairs loops for material balance (material inventory) control.



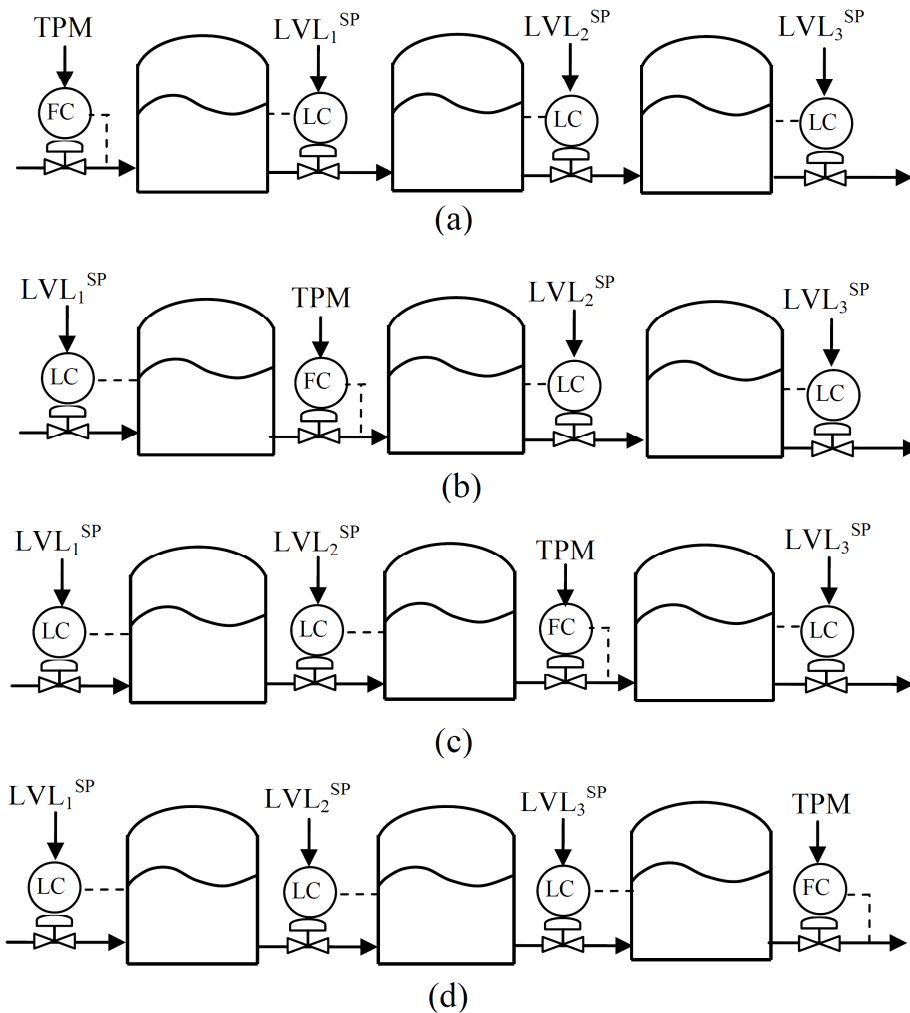


Figure 10.1. Alternative TPM locations and material balance control in three tanks-in-series process

### 10.3. Regulatory Plantwide Control Structure Synthesis Examples: Conventional vs Luyben's Approach

We are now ready to synthesize and contrast plantwide control structures using the conventional approach and Luyben's approach. For continuity, we consider the two example processes in Figure 9.4.

#### 10.3.1. Single Column Recycle Process

In the conventional approach, the TPM is chosen at a process fresh feed. Let us say the fresh B feed ( $F_B$ ) is the TPM (1<sup>st</sup> loop). The reactor temperature ( $T_{rxr}$ ) is then controlled using its cooling duty ( $Q_{rxr}$ ), which would provide tight temperature control to regulate the reactor energy balance (2<sup>nd</sup> loop). Its level ( $LVL_{rxr}$ ) is controlled using the total flow out of the reactor ( $F_1$ ) to



constraint such as on-demand operation or a process feed set by an upstream process. Where to locate the TPM is then left as a decision to be taken later. The next loop to be implemented then is the product purity control loop. For tight regulation of  $x_B^{B1}$ , a cascade arrangement is implemented with the  $x_B^{B1}$  adjusting the setpoint of the  $T_{col}^S$  controller which manipulates the column boilup ( $V_1$ ) (2<sup>nd</sup> loop). In the absence of any other information, the next loops to be implemented are ones for feeding the fresh feeds as make-up streams. The total flow to the reactor ( $F_{rxr}$ ) is maintained by adjusting  $F_B$  (3<sup>rd</sup> loop).  $F_A$  is maintained in ratio with  $F_B$  and its setpoint is adjusted to maintain  $x_B^{rxr}$  (4<sup>th</sup> loop). With these two loops, the recycle rate and composition are not allowed to float or float only within a very narrow band. Snowballing is thus mitigated. We are now ready to put in the material balance control system. The pairings  $LVL_{rxr}$ - $F_{col}$ ,  $LVL_{RD}$ - $D_1$ ,  $LVL_{bot}$ - $B_1$  and  $P_{col}$ - $Q_{cnd}$  are chosen for regulating the liquid and vapor inventories in the process (5<sup>th</sup> – 8<sup>th</sup> loops). Lastly, the  $L_1/F_1$  ratio loop is chosen for managing the column reflux (9<sup>th</sup> loop). The control structure obtained is shown in Figure 10.2(b). Even as it ‘looks’ very similar to the conventional structure (Figure 10.2a), the design philosophy including how fresh feeds are managed and the prioritization of the control objectives is very different. To manipulate the throughput, we may adjust either of the  $T_{rxr}$ ,  $x_B^{rxr}$  or  $F_{rxr}$  setpoints. Usually  $T_{rxr}$  is not adjusted as the catalyst has a very narrow operating temperature range for which the manufacturer guarantees catalyst life. Also, usually the reactor must be operated with one of the reactants being limiting which would fix  $x_B^{rxr}$ .  $F_{rxr}^{SP}$  is then the only option for the TPM.

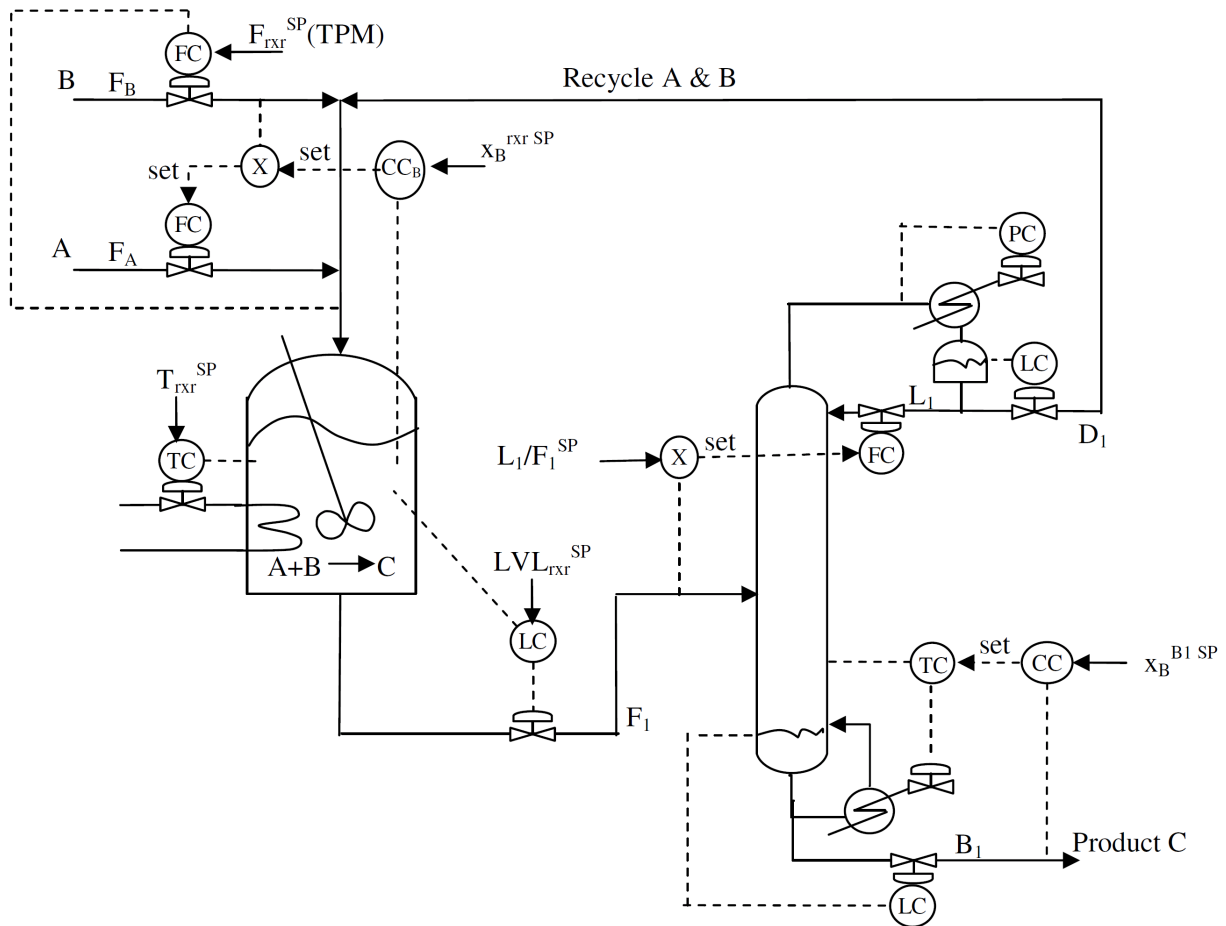


Figure 10.2(b). Luyben's control structure with TPM at reactor inlet

### 10.3.2. Two Column Recycle Process

The conventional plantwide control structure for the two column recycle process (Figure 9.4b) is synthesized as follows. Let us say the fresh B ( $F_B$ ) is the TPM (1<sup>st</sup> loop). The reactor temperature ( $T_{rxr}$ ) is controlled using the reactor cooling duty ( $Q_{rxr}$ ) for tight energy balance regulation on the most non-linear process unit (2<sup>nd</sup> loop). Material balance control consists of controlling reactor level ( $LVL_{rxr}$ ) using reactor outlet flow ( $F_{col1}$ ), the two reflux drum levels ( $LVL_{RD1}$  and  $LVL_{RD2}$ ) using the respective distillate flows ( $D_1$  and  $D_2$ ), the two column sump levels ( $LVL_{bot1}$  and  $LVL_{bot2}$ ) using the respective bottoms flows ( $B_1$  and  $B_2$ ) and the two column pressures ( $P_{col1}$  and  $P_{col2}$ ) using the respective condenser duty valves ( $Q_{cnd1}$  and  $Q_{cnd2}$ ) (3<sup>rd</sup> to 9<sup>th</sup> loops). We now implement component inventory control loops. On the first column, the reflux is maintained in ratio with the feed to provide loose regulation of the C impurity in the A recycle stream (10<sup>th</sup> loop). A sensitive stripping tray temperature ( $T_{col1}^S$ ) is maintained by adjusting the boilup ( $V_1$ ). The temperature setpoint is adjusted by an A impurity in product ( $x_A^{D2}$ ) controller in a cascade arrangement (11<sup>th</sup> loop). On the second column, the reflux is maintained in ratio with the feed and the  $L_2/B_1$  ratio setpoint is adjusted by a B impurity in product ( $x_B^{D2}$ ) controller (12<sup>th</sup> loop). The column boilup ( $V_2$ ) is manipulated to hold a sensitive stripping tray temperature ( $T_{col2}^S$ ) constant to regulate the C leakage down the bottoms (13<sup>th</sup> loop). The last loop must ensure that  $F_A$  exactly balances  $F_B$  (TPM) to satisfy the overall plant material balance through the reaction stoichiometry. The total (fresh + recycle) A rate ( $F_{TotA}$ ) to the reactor is maintained by adjusting  $F_A$  (14<sup>th</sup> loop). The control structure is shown in Figure 10.3(a). Note that in this control scheme, the B recycle can show large swings due to the snowball effect.

We now synthesize the regulatory plantwide control structure using Luyben's pairing approach. The  $T_{rxr}$ - $Q_{rxr}$  pairing is first selected for robust stabilization of the reactor energy balance (1<sup>st</sup> loop). As in the previous example, we assume that the TPM can be chosen anywhere in the plant and leave the decision for later. The next loops to be implemented are for tight product impurity control. The two impurities in the product are A leaking down the first column and B leaking up the second column. For tight regulation of the former, the  $T_{col1}^S$ - $V_1$  pairing is selected with the temperature setpoint cascaded by a  $x_A^{D2}$  controller (2<sup>nd</sup> loop). For tight regulation of  $x_B^{D2}$ , a sensitive rectifying tray temperature in the second column ( $T_{col2}^R$ ) is maintained by manipulating  $V_2$  with its setpoint cascaded by the  $x_B^{D2}$  controller (3<sup>rd</sup> loop). Tray temperature control using boilup achieves the tightest temperature control on a column. Here, this dynamic advantage of the pairing is leveraged for achieving tighter B impurity control than the conventional pairing with reflux rate (or ratio). With the product impurity loops in place, we implement loops for feeding the fresh feeds as make-up streams. The total (fresh + recycle) B ( $F_{TotB}$ ) to the reactor is maintained constant by manipulating  $F_B$  (4<sup>th</sup> loop). The total (fresh + recycle) A ( $F_{TotA}$ ) to the reactor is maintained by adjusting  $F_B$  and its setpoint is maintained in ratio with  $F_{TotB}$  (5<sup>th</sup> loop). Maintaining  $F_{TotA}$  and  $F_{TotB}$  using the fresh feeds ensures the unreacted A and B component inventories in the recycle loops are tightly regulated to mitigate snowballing. Maintaining  $F_{TotA}$  in ratio with  $F_{TotB}$  mitigates the transient variability in the reactor composition. The pairings  $LVL_{rxr}$ - $F_1$ ,  $LVL_{RD1}$ - $D_1$ ,  $LVL_{RD2}$ - $D_2$ ,  $LVL_{bot1}$ - $B_1$ ,  $LVL_{bot2}$ - $B_2$ ,  $P_{col1}$ - $Q_{cnd1}$  and  $P_{col2}$ - $Q_{cnd2}$  are implemented to control the process liquid and vapor inventories (6<sup>th</sup> – 12<sup>th</sup> loops). The last two loops to be implemented are holding the two column reflux rates in ratio with the column feeds ( $L_1/F_1$  and  $L_2/B_1$ ) (13<sup>th</sup> – 14<sup>th</sup> loops). In conjunction with the temperature loops on the two columns, these two loops ensure the impurity leakage in the two recycle streams is loosely regulated. The control structure is shown in Figure 10.3(b). To manipulate the

throughput,  $T_{\text{rxr}}$ ,  $F_{\text{TotA}}$  or  $F_{\text{TotB}}$  may be adjusted. Usually, one is not free to adjust  $T_{\text{rxr}}$ . Also, the reactor must be operated with a minimum excess of one of the reactants (say A). The total limiting reactant (B) flow to the reactor ( $F_{\text{TotB}}$ ) would then be an appropriate TPM. We again highlight that even as the structures in Figure 10.3(a-b) ‘look’ similar, their synthesis philosophies are very different.

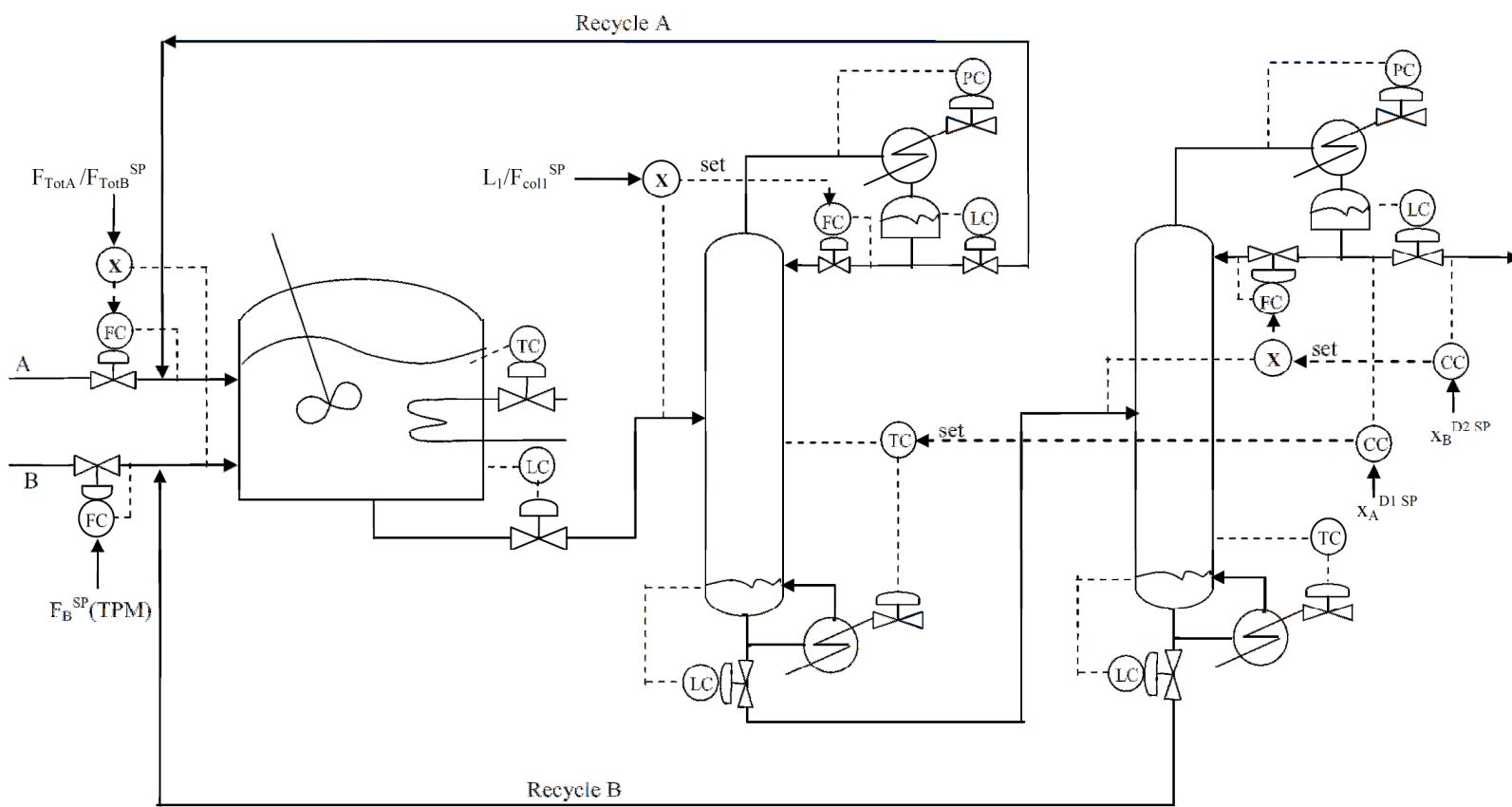


Figure 10.3(a). Conventional control structure for two column recycle process

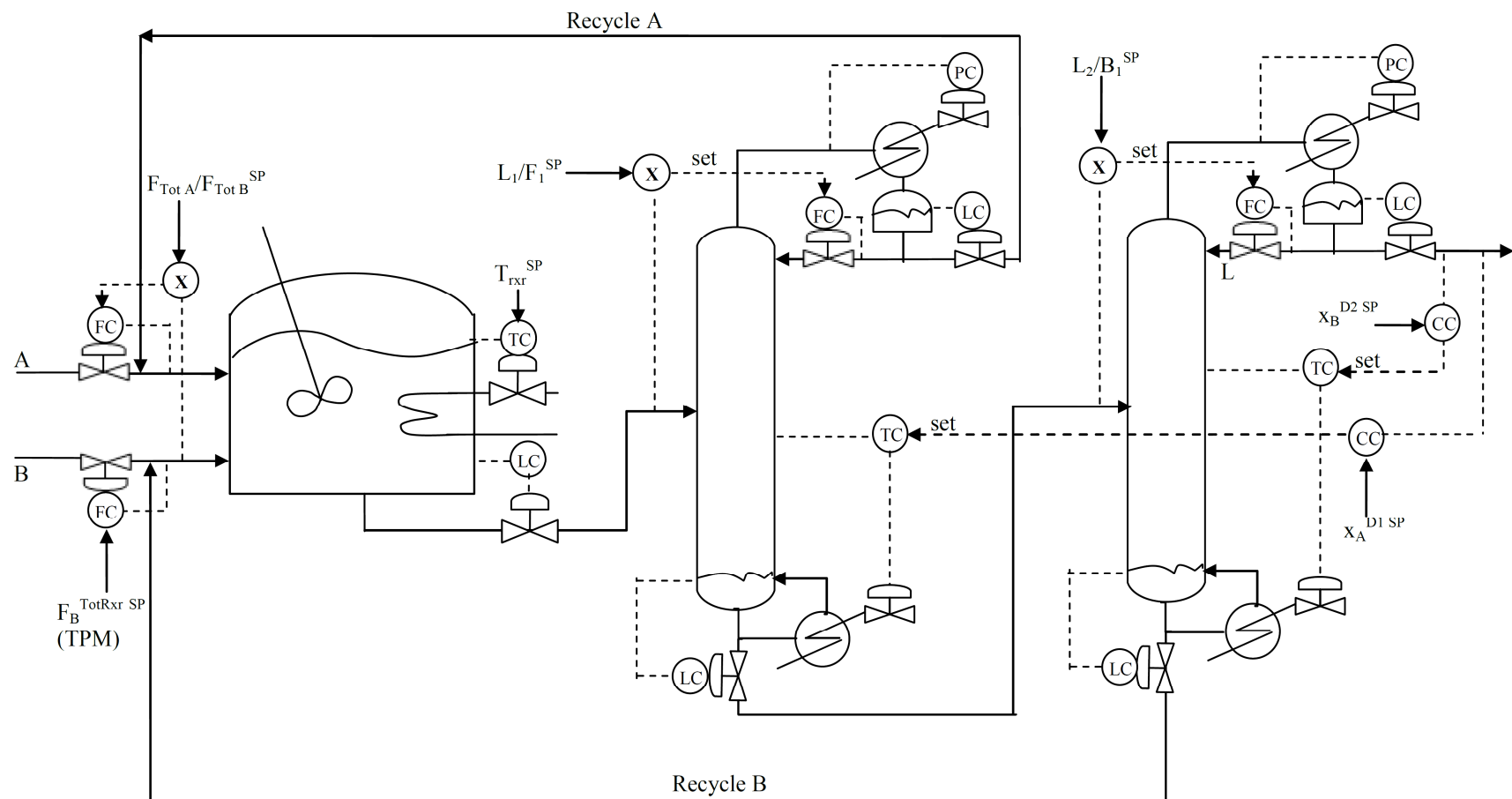


Figure 10.3(b). Luyebn's control structure for two column recycle process

## Chapter 11: Economic Considerations in Plantwide Control

Given a regulatory plantwide control structure that ensures the unit specific and overall material and energy balances are satisfied so that the process inventories do not drift or drift within an acceptably small band, we are ready to bring in economic considerations. The key question is, “What are the process inventories that significantly affect steady operating profit and their optimal levels (values)?” Engineering common sense applied to a process would usually reveal the economically important inventories and we discuss some of the considerations below.

### 11.1. Economic Process Operation Considerations

From the economic point standpoint, on-aim product purity is always desired. The product then contains maximum allowed impurity for zero product give-away or alternatively, for selling maximum allowable cheap impurities for the price of the product (legal adulteration!). Because process raw materials (reactants) are usually quite expensive (much much more than energy), their loss in non-product streams (eg a purge stream or a waste-product stream) discharged from the process must be regulated tightly at an acceptably small value. This includes minimizing the loss of expensive reactants as undesired by-products that are discharged from the plant, since the waste product consumes expensive reactants with no sales revenue.

In reactors, there usually exists a single-pass conversion versus selectivity (yield to desired product) trade-off. Side reactions always occur in any reactor and these are often suppressed by designing the reactor to operate in large excess of a reactant. One would like to maximize the single-pass reactor conversion to reduce the amount of unreacted reactants to be recycled and hence the associated recycle cost. For irreversible reactions, this would correspond to operating the reactor at the maximum allowed temperature. However, because the activation energy of the side reaction(s) is higher than the main reaction with the catalyst significantly reducing the activation barrier for the main reaction, the %age increase in reaction rate per unit temperature increase is higher for the side reaction. Thus for irreversible catalytic reactions, any increase in conversion via an increase in temperature comes at the expense of reduced yield to desired product. The reactor temperature is then likely to have an optimum conversion-yield trade-off with higher single pass conversion reducing the recycle cost (lower unreacted reactants to be recycled) at the expense of lower yield to desired product. If the process is such that the by-product is simply discharged from the process, the loss in yield dominates since energy is significantly cheaper than raw materials and the reactor operating conditions must be chosen to maximize yield. This would usually correspond to maximizing the excess reactant composition in the reactor, usually limited by a recycle equipment capacity constraint, along with an optimal temperature for high yield (say >95%) and not-too-low a conversion. In cases where the by-product is further processed back to the desired product, there is an associated processing cost which goes up as the by-product formation rate goes up (with increase in temperature). The reactor temperature would then still have an optimum; however since both reactant recycle cost and side-product processing cost primarily correspond to energy consumption (which is cheap), it would usually be optimal to have lower than maximum achievable excess reactant in the reactor and a higher operating temperature (as no by-product is discharged).

Unlike the reactor temperature, the reactor hold-up (level for liquid phase reactors and pressure for gas phase reactors) affects all the reaction rates equally with a eg 10% increase in



hold-up causing a 10% increase in all reaction rates. For kinetically limited reactors (ie all irreversible reactions and reversible reactions where the reactor is not large enough for equilibrium to be attained), it is then always optimal to operate at maximum reactor hold up (maximum level for liquid phase CSTRs and maximum pressure for gas phase reactors) as we get an increase in conversion with no yield penalty.

For optimal operation, the total energy consumption per kg product should generally be as small as possible. Heuristics for energy efficient operation of common unit operations are well-known and should be liberally applied. This includes preventing over-refluxing in distillation columns by dual-ended control, efficient operation of furnaces by adjusting the fuel to air ratio to maintain stack-gas composition, floating pressure control of a superfractionator, using valve position control on a variable speed pump feeding parallel process trains etc. These heuristics have been discussed earlier.

### **11.2. Process Operation Modes**

Continuous chemical processes are usually operated in 2 modes. In Mode I, the process throughput (production rate) is specified based on market demand-supply considerations and economic operation is equivalent to maximizing process efficiency (eg minimum steam consumption per kg product or maximum yield to desired product etc). In Mode II, the market conditions are such that it is optimal to operate the process at maximum (economic) throughput. Plants immediately after commissioning are often operated at maximum throughput to maximize revenue and pay-off debts. First-to-patent product / process monopolies may also be operated at maximum throughput given sufficient product demand.

### **11.3. Process Constraints and Economic Operation**

The discussion on economic considerations hints at economic process operation requiring operation at or close to constraints. The constraints may be soft, where short duration constraint violations are acceptable, or hard, where constraint violations are unacceptable or not possible. Process operation at the maximum allowed product impurity constraint for no product give-away is an example of a soft constraint. Hard constraints usually correspond to equipment capacity constraints. Examples include operating a gas recycle compressor at maximum duty to maximize gas recycle rate and hence minimize fresh gas consumption, operating a distillation column at its flooding limit (maximum boilup) to maximize the recycle of the excess reactant for suppressing a side reaction etc.

At the design throughput, hard equipment capacity constraints are usually not active (due to equipment overdesign). However, as throughput is increased, equipment successively hit capacity constraints. For example, the boilup in a distillation column is commonly manipulated for stripping tray temperature control. As throughput is increased sufficiently, the boilup would increase to a point where the column approaches its flooding limit with the high boilup not allowing liquid to drop down the trays. Upon hitting the flooding limit (maximum boilup,  $V^{\text{MAX}}$ ), tray temperature control would be lost. The loss in tray temperature control would imply loss in regulation of the light key dropping down the column. Let us say the column bottoms stream is a product stream. Product light key impurity control is then lost, which is unacceptable. If the bottoms is a recycle stream, the light key inventory in the recycle stream is unregulated and can

build-up (snowballing) unless the throughput is cut. The point is that as constraints go active, regulation of crucial control tasks may be lost.

## **11.4. Approaches for Handling Equipment Capacity Constraints**

### ***11.4.1. Backed-off Operation***

How does one handle equipment capacity constraints going active? Consider the simple distillation column with conventional single ended temperature control using boilup and maximum boilup ( $V^{\text{MAX}}$ ) representing a capacity constraint. The simplest thing to do would be to back-off the column feed sufficiently so that  $V^{\text{MAX}}$  does not go active for the worst expected disturbance. This is illustrated in Figure 11.1(a). The maximum achievable steady throughput would then be lower, representing an economic loss.

### ***11.4.2. Use of Valve Positioning (Optimizing) Controller***

To automate the back-off in throughput, one may implement a valve positioning controller that maintains the boilup at a specified value by manipulating the feed rate. This is shown in Figure 11.1(b). Since an adjustment in feed by the VPC would affect the boilup reasonably quickly through the action of the temperature controller, the back-off would be lower than what was necessary using the strategy in Figure 11.1(a). Even so, some back-off would be necessary representing a loss in maximum throughput ie an economic loss.

In the control system in Figure 11.1(b), the VPC setpoint sets the feed to the column and thus indirectly acts as the TPM. A simple and effective control scheme for handling the  $V^{\text{MAX}}$  constraint is to directly use the boil-up flow setpoint as the TPM and control tray temperature using the column feed, as shown in Figure 11.1(c). Increasing the boilup would cause the tray temperature to increase and the temperature controller would increase the cold fresh feed to bring the increasing temperature back to setpoint. The temperature control would be reasonably tight as long as the control tray is not too far below the feed tray. Notice that due to tight control of the boilup using reboiler duty, little/no back-off from the  $V^{\text{MAX}}$  limit would be necessary so that the process can be operated at  $V^{\text{MAX}}$  with no (or negligible) loss in maximum achievable throughput.

### ***11.4.3. Altering Material Balance Control Structure Using Overrides***

There is also the conventional approach of handling constraints using override controllers. The  $V^{\text{MAX}}$  constraint on a distillation column is conventionally handled by a slower override tray temperature controller with its setpoint slightly below the nominal setpoint and its output passing to the column feed valve through a low select, as shown in Figure 11.2(a). When  $V^{\text{MAX}}$  is inactive, the nominal temperature controller controls tray temperature close to the nominal setpoint. The tray temperature is then higher than the override temperature controller setpoint so that its output increases in an attempt to put more cold feed to reduce the tray temperature to its setpoint. The output is then high and the low select on the signal to the feed valve passes the desired feed throughput signal (column feed as TPM). When the  $V^{\text{MAX}}$  constraint goes active on eg sufficiently increasing column feed rate, the tray temperature would

decrease causing a decrease in the override controller output with the low select eventually passing feed manipulation to the override temperature (column feed under temperature control). The override scheme thus alters the control structure from fixed feed – manipulated boilup to fixed boilup – manipulated feed.

In case the feed to the column is being set by an upstream process eg by the level controller of the upstream reactor, the temperature override taking up column feed manipulation would imply loss of level control on the reactor. The reactor level would then increase and an override level controller with its setpoint slightly higher than the nominal level controller setpoint must now take up manipulation of reactor feed to regulate its level. Appropriate overrides will have to be implemented all the way back to the process feed, as shown in Figure 11.2(b-c). Regardless of the number of intervening units between the process feed and the constrained unit, what the override scheme does is alter the material balance control structure from fixed process feed – varying constraint variable (boilup in the distillation example) to fixed constraint variable – varying process feed.

#### ***11.4.4. Using Constraint Variable as Throughput Manipulator***

The use of overrides for altering the material balance control structure on hitting a constraint can be avoided as illustrated in Figure 11.3. Here, the constraint variable is the TPM and the material balance control loops are oriented around it using the radiation rule. Clearly, this gives a much simpler control system with no overrides. Also, no (minimum) back-off is needed from the active constraint limit. In contrast, a major disadvantage of using overrides is the need for appropriate offset in override controller setpoints. In the Figure 11.2 examples, the nominal reactor level setpoint would necessarily be lower than maximum implying that the nominal process operation would be at a lower than maximum single pass conversion due to lower than maximum holdup with consequent higher recycle cost. Similarly, the offset in the column temperature override controller would imply higher steady loss of the light-key down the bottoms once  $V^{\text{MAX}}$  goes active. The overrides also introduce an inherent dynamic disadvantage with the overrides taking time to take-over and give up control and also an element of on-off control with potential repeated misfiring causing unnecessary plantwide transients, particularly when the final steady state is not at the constraint limit but slightly below it. In our considered view, the use of overrides should be minimized as far as possible and using a (hard) equipment capacity constraint variable controller setpoint as the TPM and orienting the material balance control system around constitutes a simple and effective way of handling one such hard constraint variable for negligible back-off and consequent economic loss.

Typically the maximum throughput solution has multiple hard active constraints. The economic loss due to a back-off from these constraints would usually be the largest only with respect to a particular constraint. We refer to this constraint as the economically dominant constraint. For economic operation, we choose this constraint variable (or setpoint of the loop that controls it) as the TPM and put in place the total material balance control system around it. This minimizes the back-off in the economically dominant constraint mitigating the consequent economic loss. The loss in control dofs due to the remaining hard active constraints is then managed with sufficient back-off from the constraint limits which causes only an acceptably small steady economic loss, since these constraints are not economically dominant.

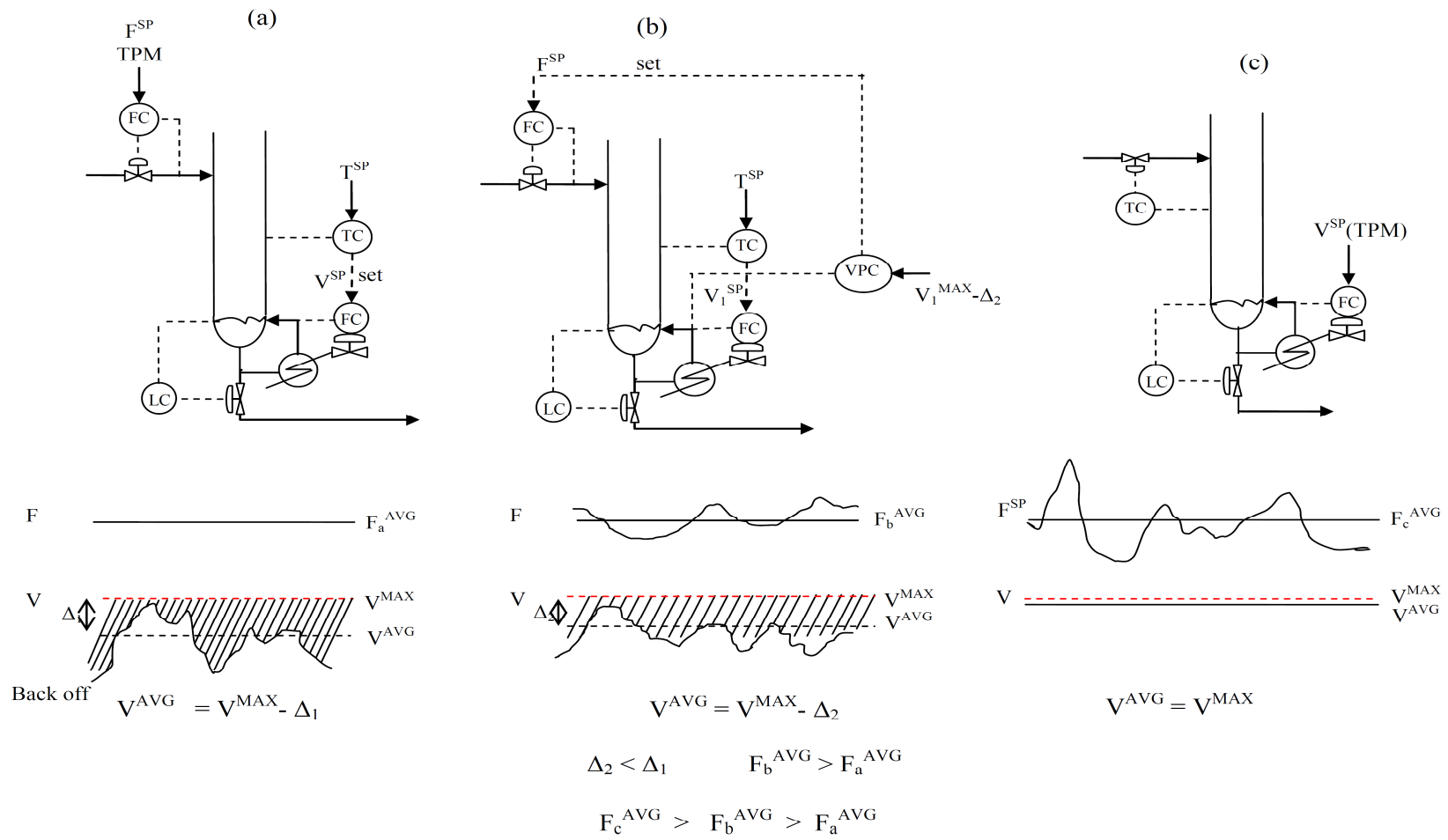


Figure 11.1. Various control scheme for handling equipment capacity constraint

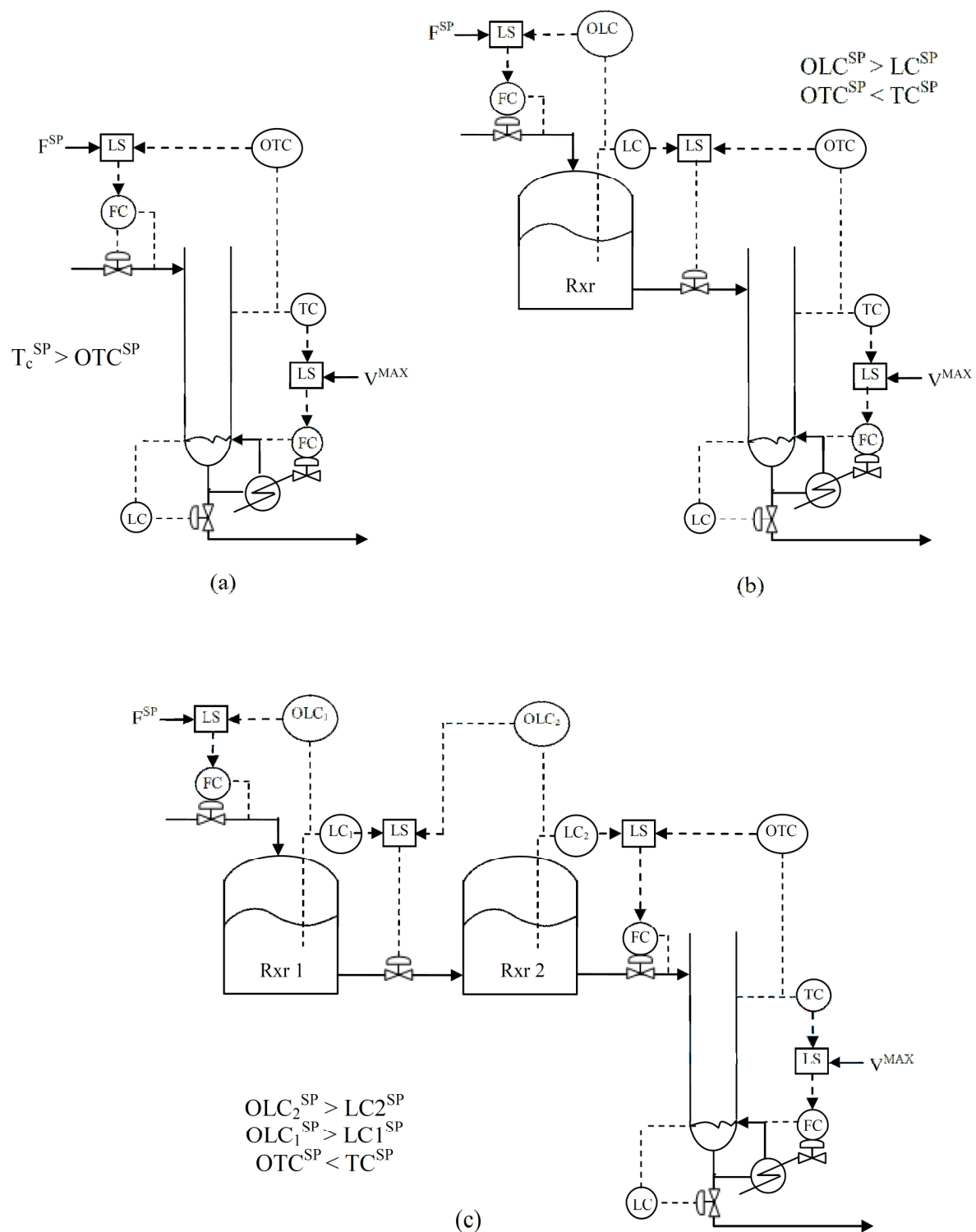
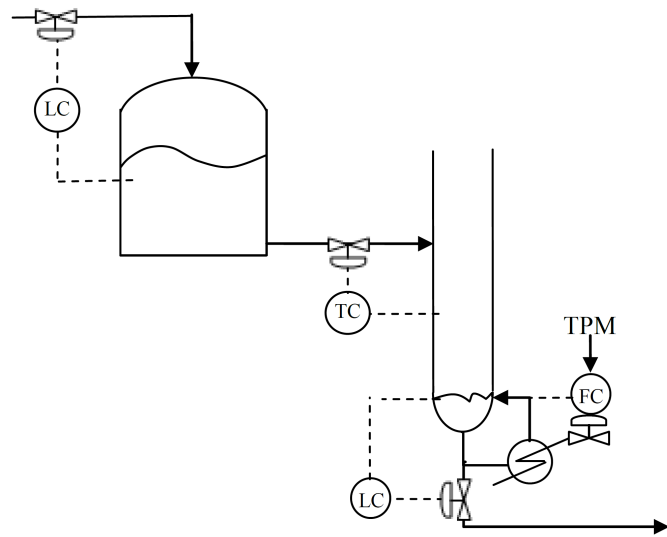
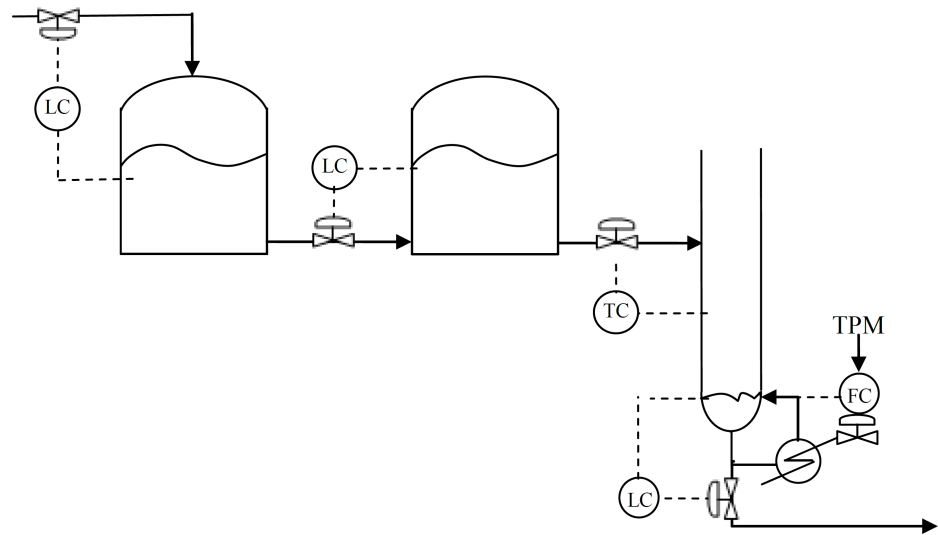


Figure 11.2. Override control scheme for handling capacity constraint



(a)



(b)

Figure 11.3. Choosing TPM at the constraint variable to avoid overrides

## Chapter 12. Economic Plantwide Control Examples

We are now ready to synthesize a plantwide control structure for economic operation of the two example chemical processes in Figure 9.4 using the engineering heuristics discussed above.

### 12.1. Single Column Recycle Process

The material, component, phase and energy inventories have already been discussed previously. We now bring in economic considerations. The process has 6 steady state dofs. Since there are no side reactions in this toy-problem, economic operation corresponds to minimizing energy consumption (i.e. column reboiler duty). If the separation in the column is relatively easy (likely as C is formed by the addition of A to B and is therefore significantly heavier than both reactants), minimizing energy consumption per kg throughput would correspond to maximizing single pass conversion and hence minimizing the recycle load. Accordingly, the reactor should be operated at maximum level ( $LVL_{\text{rxr}}^{\text{MAX}}$ ) and temperature ( $T_{\text{rxr}}^{\text{MAX}}$ ). Also, no product giveaway requires the B impurity in the product to be at its maximum allowed limit ( $x_B^{\text{B1 MAX}}$ ). These three constraints would be active regardless of throughput (ie both in Mode I and Mode II) and account for three steady state dofs.

In Mode I, the throughput ( $F_A$ ) is specified leaving 2 unconstrained dofs. These correspond to the C leakage in the recycle stream and the B composition in the reactor ( $x_B^{\text{rxr}}$ ) or more generally, in the recycle loop. If too little C leaks up the top (sharp separation), the boil-up increases (higher reflux for the sharper rectification). On the other hand, if too much C leaks up the top, the reactor gets diluted with the recycle C and the reactor reactant composition goes down for lower single pass conversion and consequent higher recycle cost. Sufficient reflux thus needs to be provided in the column so that too much C does not leak up the top. This is achieved by maintaining the reflux in ratio with the column feed ( $L_1/F_1$ ) ensuring adequate C regulation at all throughputs.

With respect to  $x_B^{\text{rxr}}$ , we note that the conversion would be maximized for comparable reactor A and B mol fractions as the irreversible reaction kinetic expression is

$$r = k x_A^{\text{rxr}} x_B^{\text{rxr}}$$

Now since the reactor contains C (generated by reaction) and its amount varies with throughput (generation rate), the optimal value of  $x_B^{\text{rxr}}$  that ensures  $x_B^{\text{rxr}} \approx x_A^{\text{rxr}}$ , would vary with throughput. Care must then be exercised that the specified  $x_B^{\text{rxr}}$  setpoint is not infeasible due to the variation in  $x_C^{\text{rxr}}$ . The optimum  $x_B^{\text{rxr}}$  would be the smallest at maximum production (largest  $x_C^{\text{rxr}}$ ) large. To ensure feasibility the desired setpoint over the entire throughput range, we may choose to implement this setpoint value at all throughputs. At low throughputs ( $x_C^{\text{rxr}}$  small due to low generation,  $x_B^{\text{rxr}}$  specified to be small), the reactor then gets operated in significant excess A environment implying higher than necessary reboiler duty.

One way around this problem is to realize that the recycle stream contains mostly A and B with only a small amount of C. If instead of holding  $x_B^{\text{rxr}}$  constant, we ensure that  $x_B^{\text{DI}} \approx 50\%$  (ie comparable A and B in recycle stream), then  $x_B^{\text{rxr}}$  would automatically float to be comparable to  $x_A^{\text{rxr}}$ . Now since B is heavier than A and therefore requires more energy to boil-off, a reasonable specification for near optimal operation over the entire throughput range would be holding  $x_B^{\text{DI}}$  slightly but not too far below 50% (say at 45%). Such a choice would ensure

reactor operation close to maximum achievable single pass conversion (an economic objective) across the entire throughput range.

As throughput is increased, let us say that the column approaches flooding. The maximum boilup ( $V_1^{MAX}$ ) then limits the maximum achievable throughput (Mode II operation). We take the two regulatory plantwide control structures synthesized earlier (Figure 10.2) and adapt them for economic operation over the entire throughput range.

In Figure 12.1(a), we take the conventional plantwide control structure with  $F_B$  as the TPM and modify it for economic operation. The setpoints for  $T_{rxr}$  and  $LVL_{L_{rxr}}$  loops are specified to be  $T_{rxr}^{MAX}$  and  $LVL_{L_{rxr}}^{MAX}$  (for maximum single pass conversion). A slow  $x_B^{D1}$  controller is implemented that adjusts the  $x_B^{rxr}$  composition loop setpoint to hold  $x_B^{D1}$  at its (near) optimal value (chosen as 45% here) for the entire throughput range. Similarly,  $L_1/F_1^{SP}$  is set at an appropriate value for ensuring too much C does not leak in the recycle stream over the entire throughput range. For maximum throughput operation with  $V_1^{MAX}$  as the bottleneck constraint, an override scheme for altering the material balance control structure is implemented. Notice that the setpoint of the nominal and override temperature controllers on the column comes from the master  $x_B^{B1}$  (product B impurity) controller. The override temperature controller setpoint is always slightly lower than the nominal setpoint via the negative bias. When the temperature override gets triggered, the product impurity would increase (as override temperature setpoint is lower) and the action of the  $x_B^{B1}$  controller would slowly bring it back to the appropriate level. On the other hand, when the nominal controller takes up temperature control ( $V_1^{MAX}$  goes inactive), since its setpoint is higher than the override setpoint, the impurity leakage would decrease (below maximum allowed) and then get back to the desired value via the action of the  $x_B^{B1}$  controller. Clearly, product impurity control becomes loose due to the overrides ‘taking over’ or ‘giving-up’ control.

To avoid the disadvantages associated with overrides, one may insist on having a fixed control structure regardless of throughput. If the conventional regulatory control loops are already implemented and are not modifiable, the only free setpoint available for maintaining the constraint variable ( $V_1$ ) at a desired value is  $F_B^{SP}$ . This loop is shown in Figure 12.1(b) and is a long one. When coupled with the snowball effect,  $V_1$  would only get controlled loosely around the desired setpoint implying a large back-off from  $V_1^{MAX}$  and consequent throughput loss.

We may also take the regulatory control structure synthesized using Luyben’s approach and adapt it for economic operation. Figure 12.2(a) shows the adapted control structure along with a material balance altering override scheme for handling the  $V_1^{MAX}$  constraint for maximum throughput operation. Figure 12.2(b) shows a long  $V_1$  constraint control loop manipulating  $F_{rxr}$  to avoid the use of override controllers. These modifications to the basic regulatory control structure are very similar to those for the conventional control structure and are therefore not elaborated upon. It is however worth mentioning that tighter  $V_1$  control by the long  $V_1$ - $F_{rxr}$  loop would be achieved as the snowball effect is mitigated with the fresh reactants being fed as make-up streams. The back-off from  $V_1^{MAX}$  would then be lower and the control scheme would achieve higher maximum throughput than the one in Figure 12.1(b).



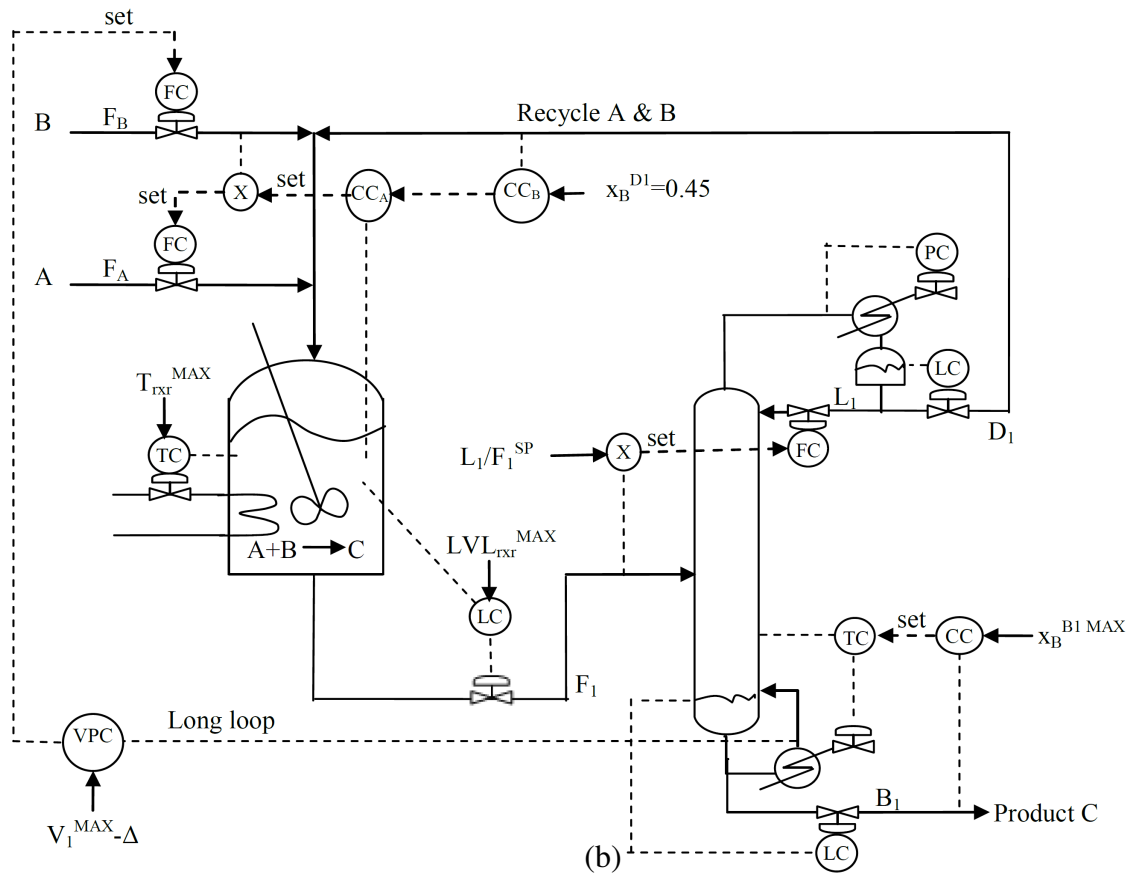
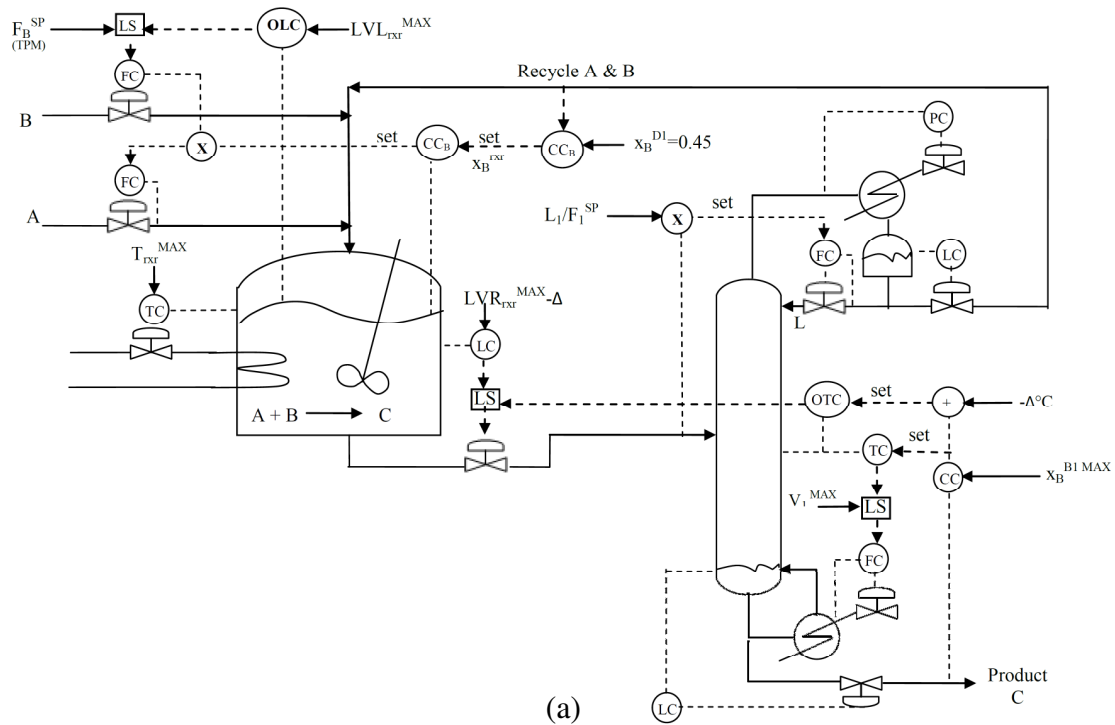


Figure 12.1. Handling capacity constraint in single column process (Conventional Process)  
 (a) Using overrides (b) Using long active constraint control loop

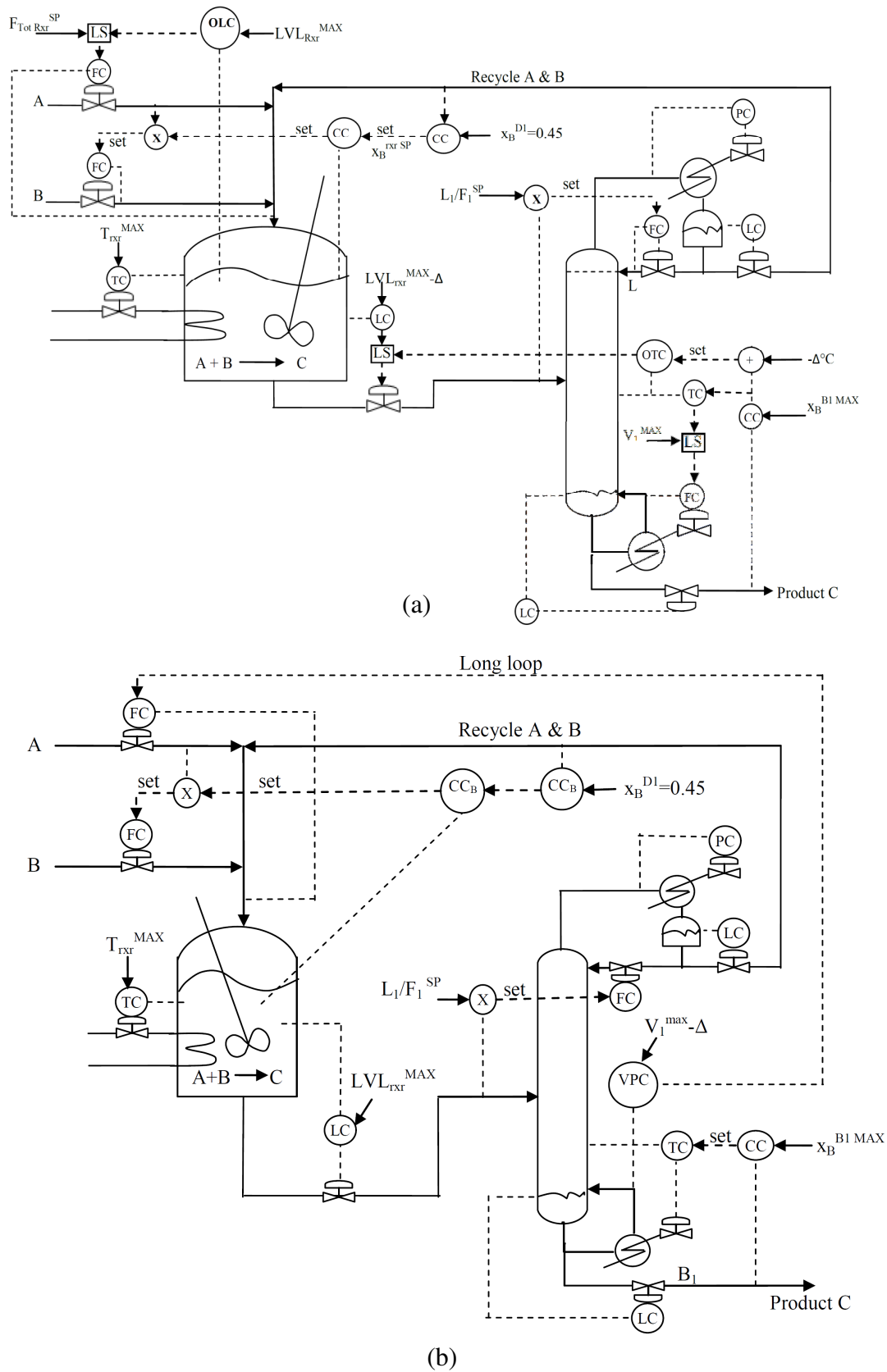


Figure 12.2. Handling capacity constraint in single column process (Luyben structure)  
 (a) Using override (b) Using long active constraint control loop

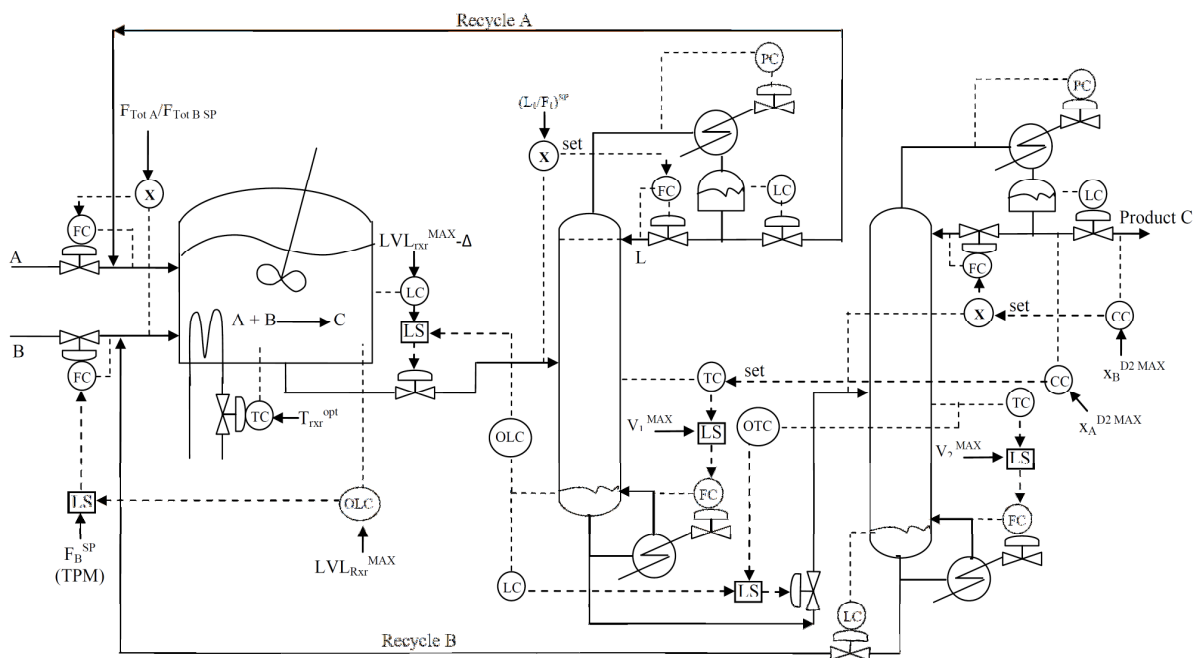
The diagram illustrates a chemical process involving a reactor and a distillation column. Two feed streams, A and B, enter the reactor. Stream A is controlled by a Level Controller (LC) with a setpoint  $LVL_{rxr}^{MAX}$  and a Feed Controller (FC) with a setpoint 'set'. Stream B is controlled by a Temperature Controller (TC) with a setpoint  $T_{rxr}^{MAX}$ . The reactor output, labeled  $A + B \rightarrow C$ , is pumped to a distillation column. The column has several control loops: a Top Product (TPM) flow  $V_1$  is controlled by a Level Setpoint (LS) with a setpoint  $V_1^{SP\ TPM}$  and a maximum value  $V_1^{MAX}$ ; a Bottom Product (BP) flow is controlled by a Level Setpoint (LS) with a setpoint  $V_1^{MAX}$  and a maximum value  $V_1^{MAX}$ ; a Recycle stream is controlled by a Level Setpoint (LS) with a setpoint  $V_1^{MAX}$  and a maximum value  $V_1^{MAX}$ ; a Bottoms (B) flow is controlled by a Level Setpoint (LS) with a setpoint  $V_1^{MAX}$  and a maximum value  $V_1^{MAX}$ . The distillation column also has a Temperature Controller (TC) and a Level Controller (LC) for the bottoms. The final product is labeled 'Product C'.

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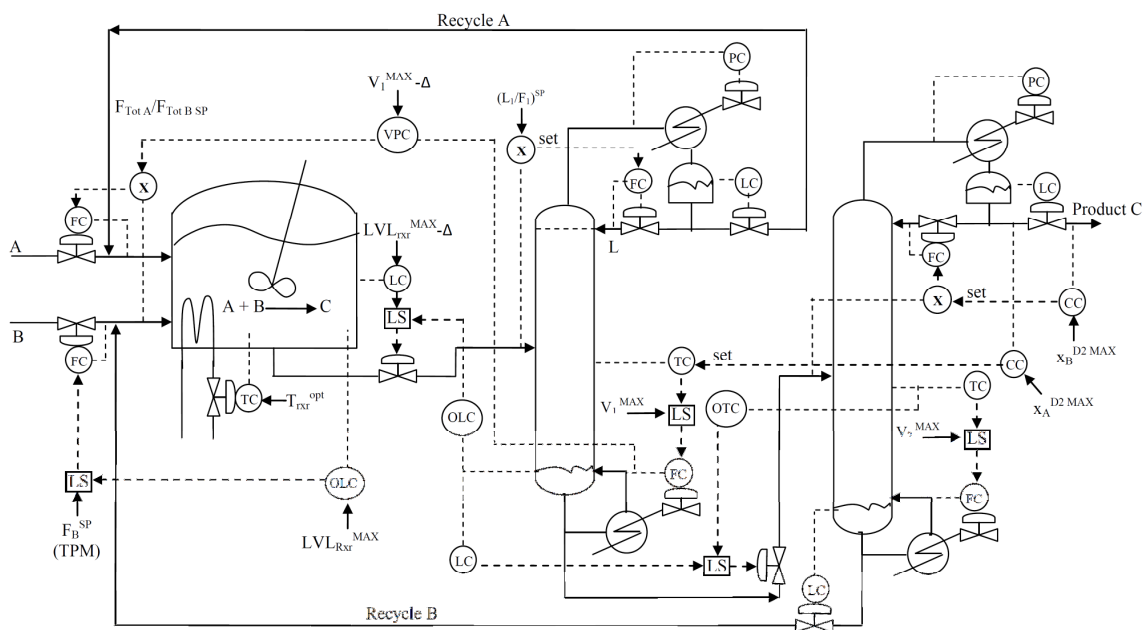
## 12.2. Two Column Recycle Process

This process has 8 steady state dofs, as discussed earlier. Purely for the sake of a more interesting discussion, let us assume that there is a side reaction (assume side product volatility is such that it leaves with product C stream) and that this side reaction is suppressed by operating the reactor in excess A environment (B limiting). Economic process operation then requires maximizing the reactor excess A environment, which requires operating the first column at maximum boilup ( $V_1^{MAX}$ ) so that the A recycle rate is as high as possible. To maximize single-pass conversion with no yield penalty, it should be operated at maximum level ( $LVL_{TXR}^{MAX}$ ). Also, the A and B impurities in the product should be at their maximum limits for no product give-away ( $x_A^{D2 MAX}$  and  $x_B^{D2 MAX}$ ). These four constraints are active at all throughputs. In Mode I (given throughput), we have a specified throughput leaving 3 unconstrained steady state dofs. These correspond to the optimum reactor temperature (conversion-yield trade-off) along with the C leakage in the A recycle stream and in the B recycle stream. This C leakage must be kept small enough at all throughputs. As throughput is increased, let us say the maximum boilup on the second column ( $V_2^{MAX}$ ) constraint is hit, which fixes the maximum achievable throughput (Mode II).

We now adapt the conventional plantwide regulatory control structure ( $F_B$  TPM) for economic operation (Figure 10.3a). The adapted control structure is shown in Figure 12.4(a). In the regulatory control structure, the product impurity control loops are already in place and their setpoints are set at the maximum acceptable impurity level ( $x_A^{D2 MAX}$  and  $x_B^{D2 MAX}$ ). The reactor level setpoint is specified at  $LVL_{TXR}^{MAX}$ . To operate close to  $V_1^{MAX}$ , a  $V_1$  controller is implemented which manipulates  $F_{TotA}/F_{TotB}^{SP}$  in a long loop. Its setpoint will require sufficient back-off from  $V_1^{MAX}$  to ensure A impurity regulation is never lost. The reactor temperature setpoint is specified at an appropriate value that ensures the yield is always sufficiently high. On the first column,  $L_1/F_1$  setpoint is fixed at a value that ensures too much C does not leak up the top over the entire throughput range. On the second column, the stripping tray temperature setpoint is chosen to regulate C leakage down the bottoms at an acceptably small value. For handling the bottleneck  $V_2^{MAX}$  constraint that limits maximum throughput, a material balance altering control scheme with overrides from the second column back to the fresh A feed is implemented. Note that  $V_2^{MAX}$  represents a capacity constraint on the amount of product C that can be boiled off. If too much C is generated in the reactor than can be boiled off in the second column, the extra C would necessarily accumulate in the B recycle stream. The override scheme acts to cut the fresh B feed to the appropriate value so that the C generation in the reactor exactly matches what is boiled off in the second column. If the override scheme for altering material balance structure is to be avoided,  $F_B^{SP}$  must get adjusted to hold  $V_2$  (constraint variable) in a long loop. While it may be acceptable to let the C impurity in the recycle stream float for short durations till the long  $V_2$  loop sufficiently reduces  $F_B^{SP}$  after  $V_2^{MAX}$  goes active, large plantwide transients due to adjustment in  $F_B$  (snowball effect) are likely and conservative operators may simply back-off  $V_2^{SP}$  sufficiently to ensure  $V_2^{MAX}$  never goes active.



(a)



(b)

Figure 12.4 Use of overrides for handling capacity constraints for the two column recycle process. (a) Conventional structure (b) Luyben's structure

Figure 12.4(b) shows the adapted control structure for economic operation with regulatory plantwide control structure from Luyben's approach. The adaptations are very similar to the conventional structure (Figure 12.4a). Note that the  $L_2/B_1$  ratio controller must be specified to a value that ensures too much C does not leak down the second column bottoms over the entire throughput range. To avoid the override scheme for altering material balance control when  $V_2^{MAX}$  goes active, one can adjust  $F_{TotB}^{SP}$  to maintain  $V_2$  in a long loop. The plantwide transients are expected to be smooth as  $F_{TotB}$  is inside the recycle loop so that  $F_B$  is always fed as a makeup stream mitigating the snowball effect and the back-off from  $V_2^{MAX}$  would be smaller.

In this example, we have two hard equipment capacity constraints,  $V_1^{MAX}$  and  $V_2^{MAX}$ . In the synthesized control structures, some back-off from  $V_1^{MAX}$  and  $V_2^{MAX}$  is needed to avoid loss of product quality control and snowballing issues. The back-off from  $V_1^{MAX}$  causes a loss in selectivity while and back-off from  $V_2^{MAX}$  causes throughput loss. The latter can be a significant economic loss and to avoid the same we may use  $V_2^{SP}$  (last constraint to go active) as the TPM and orient the material balance control system around it as shown in Figure 12.5.  $T_{col2}^S$  is controlled using  $B_1$ ,  $LVL_{bot1}$  is controlled using  $F_{col1}$  and  $LVL_{trf}$  is controlled using  $F_{TotB}$ . As before,  $F_{TotA}$  is maintained in ratio with  $F_{TotB}$  to ensure the reactor feed composition does not vary too much. The ratio controller also ensures tight reactor level control with the total reactor feed varying in response to a change in its level. The rest of the control system is self explanatory.

Can we further alter the control structure to ensure the back-off from  $V_1^{MAX}$  is also eliminated. We show one possible control structure (there are other possibilities too) in Figure 12.6. Here,  $V_2^{SP}$  is used as the TPM as before. Since  $V_1^{MAX}$  is active, it is not used for controlling  $T_{col1}^S$  and  $F_{col1}$  is adjusted instead to ensure the A impurity in the product is always regulated.  $LVL_{bot1}$  is then controlled using  $B_1$  and  $LVL_{RD1}$  is controlled using  $D_1$ . Similarly  $LVL_{RD2}$  and  $LVL_{bot2}$  are regulated using  $D_2$  and  $B_2$  respectively.  $LVL_{trf}$  is controlled using  $F_{TotA}$  with  $F_{TotB}$  maintained in ratio to ensure the proper A excess in the feed to the reactor. The column pressures are controlled using the respective condenser duty valves. For product impurity control, the  $x_A^{D2}$  controller adjusts the  $T_{col1}^S$  controller setpoint while the  $x_B^{D2}$  controller adjusts  $L_2/B_1$ , as before. On the second column, no close by valves are available for stripping tray temperature control and the C leakage in  $B_2$  remains unregulated.  $V_2^{SP}$  (TPM) fixes the product C boil-off from the second column and if more C is being generated in the reactor than what is boiled-off, it would drop down the second column and  $B_2$  can show a very large increase (snowballing). To mitigate the same,  $B_2$  is loosely regulated by adjusting the  $F_{TotB}/F_{TotA}^{SP}$ . If  $B_2$  increases, the ratio setpoint is increased causing a decrease in  $F_B$  with  $F_A$  also eventually decreasing so that only as much C is produced in the reactor as is being boiled off in the second column. Loose control of  $B_2$  flow rate is acceptable as it is a recycle stream and not an exit (product, byproduct or purge) stream. This example illustrates that economic considerations, in particular, tight control of equipment capacity constraints, results in a plantwide control structure that is very different from structures synthesized using the conventional approach or Luyben's approach.

The two toy problems considered here illustrate how economic considerations impact plantwide control structure design. We also hope that the elaborate discussion for the two case studies convinces the readers that common sense based process engineering principles clearly bring out the major considerations in economic / efficient process operation, at least at the qualitative level. These economic considerations, including equipment capacity constraints,

translate to economic control objectives, which then govern the pairings to be implemented for achieving economic plantwide control. In the next Chapter, we consolidate the qualitative discussions here into a systematic step-by-step procedure for synthesizing an economic plantwide control system. The application of the procedure to five example processes with rigorous dynamic simulation results is presented in the subsequent chapters.

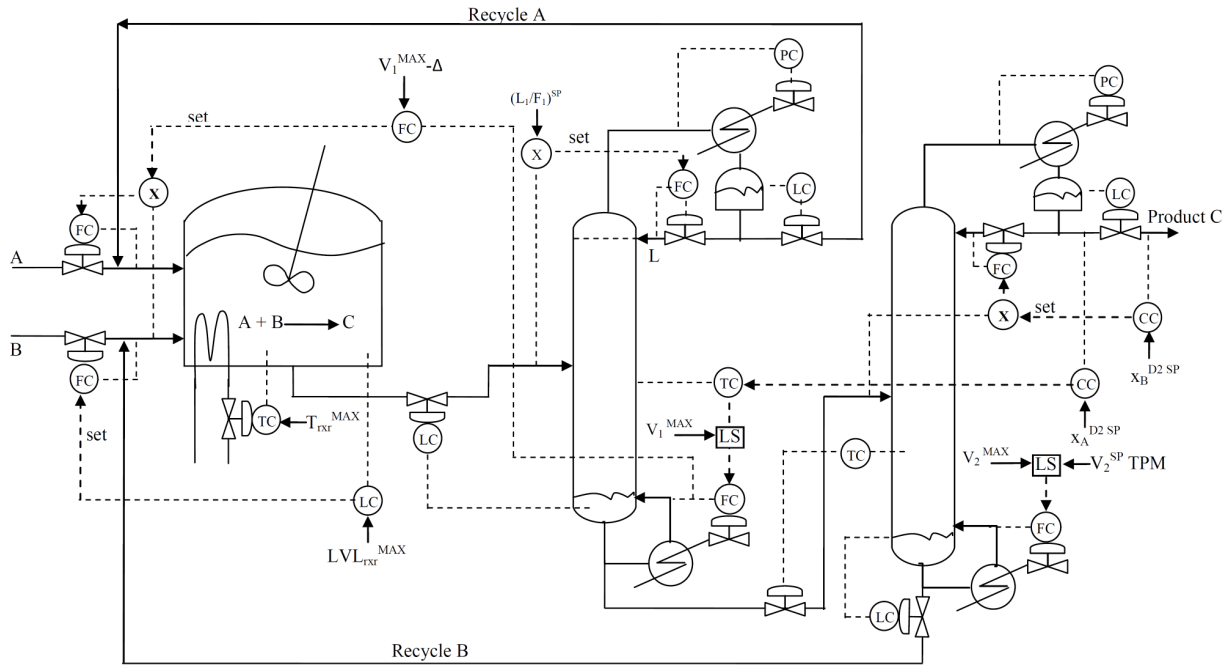


Figure 12.5. Use of bottleneck constraint as TPM to reduce overrides in the two column recycle process example

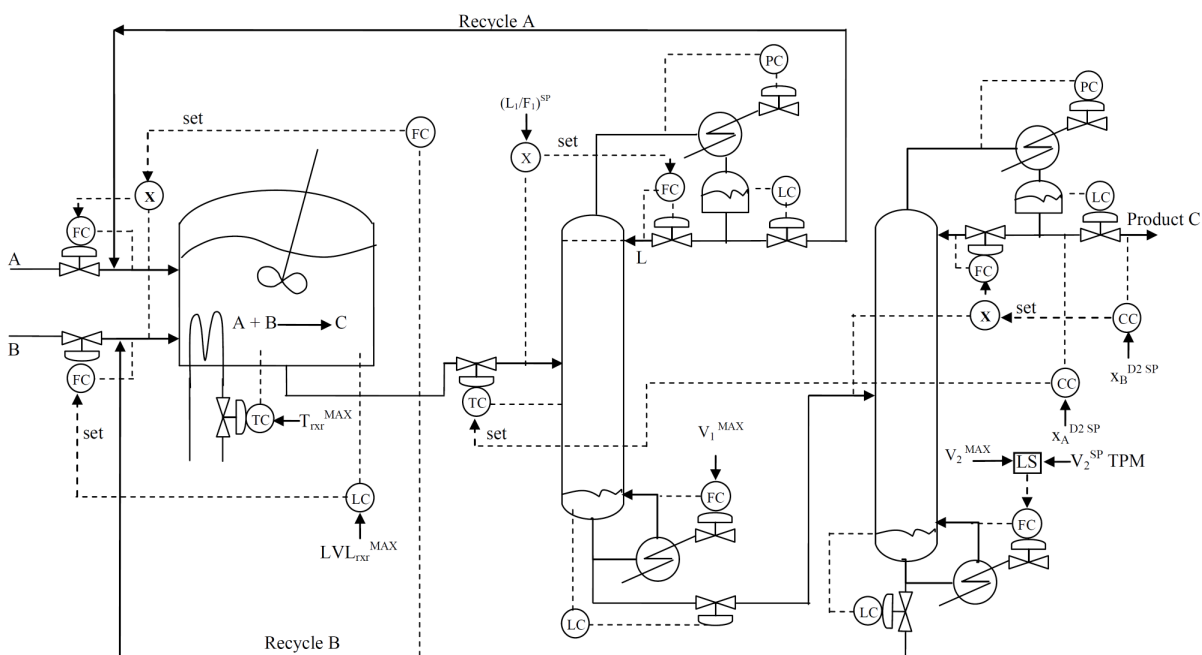


Figure 12.6. A control structure for the two column recycle process that allows operation at  $V_1^{\text{MAX}}$  and  $V_2^{\text{MAX}}$  with no back-off