MODULE IV

ECONOMIC PLANTWIDE CONTROL DESIGN PROCEDURE AND CASE STUDIES

With an appreciation of the regulatory and economic considerations in plantwide control system design, we are now ready to develop a systematic plantwide control system design procedure. We develop and present such a design procedure, which is a natural extension of the pioneering work of Page Buckley (DuPont), William Luyben (Lehigh), Jim Downs (Eastman) and Charlie Moore (Tennessee). Its application to four realistic processes, namely, a recycle process with side reaction, an ethyl benzene process, a cumene process and a C_4 isomerization process is also demonstrated. The last two examples are very comprehensive in that the performance of the economic plantwide control structure synthesized from our procedure is compared with a conventional plantwide control structure.

Chapter 13. Systematic Economic Plantwide Control Design Procedure

With the preliminaries on regulatory and economic operation considerations in plantwide control, we are now ready to develop a systematic procedure for designing an economic plantwide control system for integrated chemical processes. For completeness, we review the major contributors to plantwide control research before developing the procedure.

The design of effective plantwide control systems for safe, stable and economic process operation of complex chemical processes with material and energy recycle has been actively researched over the last two decades. The ready availability of dynamic process simulators has been crucial in fostering the research. Over the years, Luyben and co-workers have done seminal work in highlighting key regulatory control issues such as the snowball effect [15] in reactor-separator recycle systems and suggesting practical control system structuring guidelines (Luyben's rules [16]) for ensuring robust process stabilization in light of the same. Based on several case-studies, a nine-step general procedure has been developed for synthesizing effective plantwide control structures for integrated chemical processes [14]. In their procedure, economic concerns are addressed indirectly in the form of requiring 'tight' control of expected economic variables such as product impurity, process yield etc. The control objectives are obtained using engineering insights and heuristics.

Skogestad [24] has developed a more systematic steady state optimization based approach for obtaining the control objectives. Typically, at the optimum steady state, multiple process constraints are active so that these constraints must be controlled tightly. For managing the remaining unconstrained steady state degrees of freedom, the control of self-optimizing controlled variables [23] (CVs) is recommended. By definition, when self-optimizing variables are held constant at appropriate values, near-optimal operation is achieved in spite of disturbances. The quest for the best self-optimizing CV set is however not always straight-forward.

The combinatorial nature of the control structure design problem results in several possible structures that provide safe and stable process operation. A very simple example is a single-inlet single-outlet surge tank with two possible orientations for its level controller. In a simple distillation column, assuming the feed is fixed, the two orientations each for the reflux drum and bottom sump level controllers results in the well-known four basic regulatory control configurations. Other control configurations are possible if instead of the process feed, one of the other associated streams (distillate, bottoms, reflux or reboiler steam) is kept fixed. In a multi-unit chemical process, there would clearly be several possible reasonable control configurations. An obvious question then is which one is best for realizing economically (near) optimal process operation with robust stabilization over the expected process operating space. Further, is there a systematic methodology for synthesizing such an 'optimal' control structure?

A careful evaluation of the plantwide control literature reveals that most of the reported case studies consider process operation around the design steady state (see these example case studies [1,18,27]), although more recently, also at maximum throughput [2,3,11,22]. Around the base-case design steady state, usually all the process units are sufficiently away from any capacity constraints while at maximum throughput, typically, multiple units hit (hard) capacity constraints. The active constraint set progressively expands with throughput to the full set at maximum throughput. The expanding set partitions the throughput range into distinct regions. Much of the open plantwide control literature addresses control system design only for a fixed active constraint set, that is, only for a distinct region. This is surprising given that a plant must be operated over a wide throughput range with different active constraints over its life-span.

In this work, we develop a systematic approach for designing a simple and robust plantwide control system for near-optimal process operation over a wide throughput range with an expanding active constraint set. The approach has evolved out of very recent comprehensive case-studies from our group [7-9]. While the principles on which it is based may be well-known, our main contribution is in bringing these scattered principles together into a meaningful, holistic and practical top-down plantwide control system design framework. The application of the proposed framework is demonstrated on three realistic example processes.

13.1. Degrees of Freedom (DOFs) and Plantwide Control Structures

The plantwide control system design problem may be viewed as seeking the best possible way of managing the available control valves (control DOFs) for ensuring safe, stable and economic process operation in the face of principal disturbances that include large changes in the production rate (throughput) as well as variability in raw material quality, ambient conditions, equipment characteristics and economic conditions (e.g. volatility in the energy prices etc). If we discount the valves used to control nonreactive material inventories (surge tank levels, given column pressures etc), the number of independent control valves remaining equals the steady state operational DOFs for the process, which by definition, is the number of independent specifications necessary to solve for the steady state solution. For a given process, one may use alternative sets of independent specification variables. From the control perspective, each such DOF specification variable is an independent CV (excluding non-reactive material inventory controllers) in the plantwide control system. Note that one setpoint gets used to set the process throughput and is referred to as the throughput manipulator (TPM).

Figure 13.1 provides an illustration of the one-to-one correspondence between the independent CV setpoints (including TPM; excluding non-reactive material inventory controllers) and the steady state DOF specification variable set for a simple reactor-recycle process with five steady-state operation DOFs. The 5 DOFs are related to 1 fresh feed, 2 reactor specifications (level and temperature) and 2 specifications for the column. Four alternative DOF specification sets are shown in Figure 13.1. Implicit in each set is an inventory control system for balancing of the process material and energy inventories as well as appropriate pairings for controlling the specification variable. We have used the radiation rule [20] for material inventory control which gives the orientation of the level controllers upstream and downstream of the TPM respectively, opposite and in the direction of process flow, respectively. Note that for a given DOF specification set, multiple possibilities exist for the choice of the pairings for controlling the specification variables as well as for the inventory loops. Lastly, there exists flexibility in the choice of the DOF specification variable set (CV set) itself. There thus exists tremendous flexibility in designing the plantwide control system which must be gainfully exploited for achieving the twin objectives of robust stabilization and economic operation.

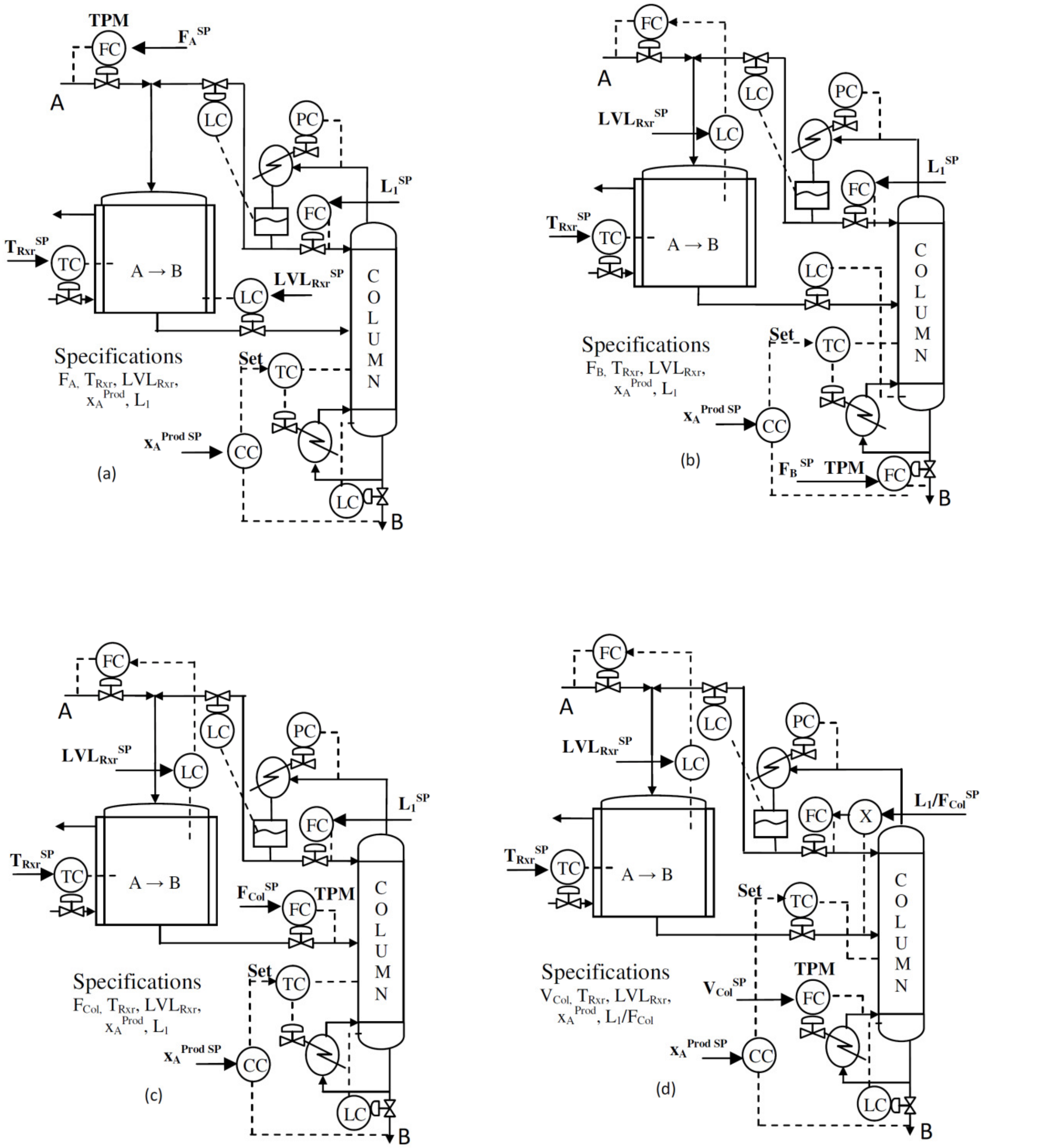

Figure 13.1. One-to-one correspondence between CV setpoints and steady state specification variables for a simple recycle process

13.2. Two-Tier Plantwide Control System Design Framework

The control system of a process plant has two main objectives:

1. Optimal economic operation: Control economic CVs
2. Stable operation: Control drifting inventories (i.e. material balance control)

'Inventory' is interpreted here in its most general sense to include material, phase, component and energy inventories in the different units as well as the overall process. The CVs for process inventory regulation (material balance control) are usually obvious. They typically include liquid levels and pressures, as well as selected temperatures, for example, a sensitive temperature in a distillation column. The best CVs for economic operation at a given throughput may be obtained from steady state optimization. Alternatively, process insight or operating experience may also suggest economically sound CVs that should be controlled.

Optimal operation requires operating the process at the optimal point, that is, *at* all the optimally active constraints as well as at the optimum value for decision variables corresponding to any remaining unconstrained DOFs. Typically, multiple constraints are active at the optimum solution. The choice of the unconstrained decision variable (CV) should be such that its optimum value is relatively insensitive to disturbances, for example, in feed rate or composition. This is the idea of 'self-optimizing' control where the economic loss due to no reoptimization for the disturbance is acceptably small. Purely from the steady state operation perspective, a constant setpoint operating policy with such CVs provides near-optimal operation in the face of disturbances. In summary, the economic CVs for optimal operation are the active constraints at the optimum plus the self-optimizing CVs corresponding to any unconstrained DOFs.

Once the set of economic CVs for a specified throughput are known (tier 1), either from economic optimization or from heuristics, the economic and regulatory loop pairings must be selected (tier 2). Which one of the two objectives (economic control or regulatory control) should have priority when designing the control system pairings (structure)? In the commonly used 'bottom-up' approach, process regulation is given priority over economic control. A 'basic' or 'regulatory' control layer with focus on inventory control (stabilization), usually with the feed rate as the throughput manipulator (TPM), is first designed. On top of this, one adds an 'advanced' or 'supervisory' control layer, often implemented using model predictive control, which aims at achieving optimal economic operation by adjusting the setpoints into the regulatory layer.

A problem with the 'bottom-up' approach is that it can yield slow control of the economic variables due to unfavorable pairings, since control valves are already paired up for regulatory control. This results in economic losses mainly because slow control requires back-off from hard active constraint limits, which can be especially costly when it is optimal to maximize throughput. As illustrated in Figure 13.2, the back-off and consequent economic penalty is primarily determined by the severity of transients in the active constraint for the worst-case disturbance. Even if the constraint is a soft one, tight regulation of the same may be desirable due to the often very non-linear nature of the process with highly skewed deviations in only one direction.

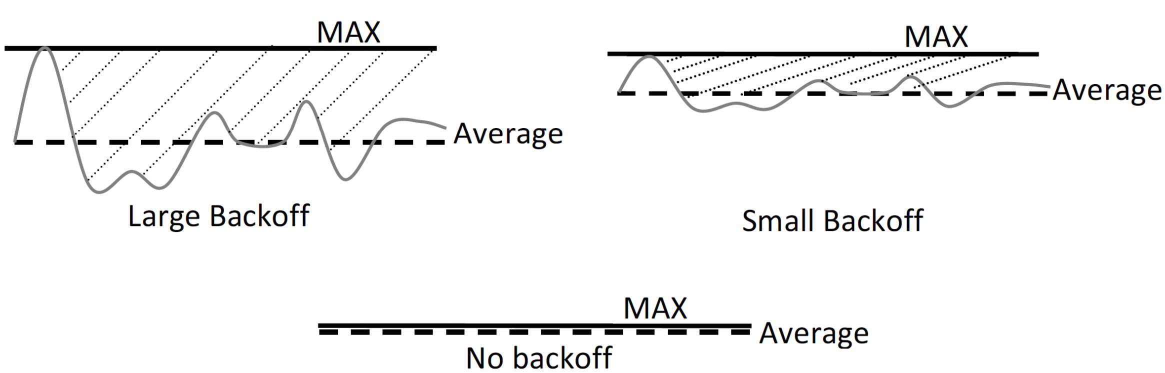

Figure 13.2. Illustration of tightness of active constraint control and back off

In this work, we consider the alternative 'top-down' approach for selecting the control pairings with higher priority to economic control over regulatory control. Such a reprioritization is natural in light of the global push towards green / sustainable / efficient process operation. In this approach, the best possible pairings for tight control of the economic CVs are obtained first followed by pairings for inventory (material balance) control. It attempts to accomplish economic and regulatory control in a single layer. The same is made possible as many-a-times controlling an economic CV accomplishes a regulatory task (and vice versa). Also, processes are designed to have sufficient number of surge capacities and the associated control valves remain available for dynamic control (including inventory control) with no steady state economic impact.

Regardless of the specific pairing philosophy (bottom-up or top-down), the application of the two-tiered framework is relatively straightforward for a given active constraint set, implying a fixed set of economic CVs that must be controlled. For most plants however, the active constraint set expands or contracts depending primarily on the plant throughput. The best economic CV set would then depend on the active constraint set (operating region) and conflicts can arise with a control valve being most suitable for robust inventory control in one region and economic CV control in another. Also, pairings done without considering the impact of a constraint going active can result in loss of crucial control functions such as product quality control or component inventory control with consequent snowballing. Additional override controllers that alter the material balance control structure may need to be configured to ensure a seamless transition and stable operation in the different regions. Alternatively, one can exploit *apriori* knowledge of the full active constraint set to devise a plantwide control system that ensures control of all critical economic and regulatory control objectives regardless of which constraints in the full active constraint set are active. Such a control system is appealing in that its basic regulatory structure remains fixed regardless of the operating region while also avoiding the need for complex over-ride controllers. The two-tiered framework must be appropriately modified to systematically devise such a control structure.

13.3. Active Constraint Regions for a Wide Throughput Range

A process is typically designed for a design throughput, where no hard constraints are active due to over-design of the different processing units. Over its life span, economic considerations necessitate sustained operation at throughputs much below and above the design throughput, usually including operation at maximum achievable throughput. As throughput increases above the design throughput, different processing units reach their (typically hard) capacity constraints, usually one after the other. These active constraints partition the entire throughput range into distinct regions. There are many disturbances in a plant, but throughput is usually considered the principal disturbance because of its wide range encompassing multiple active constraints. A control system that works well for such a large throughput range would also handle other routine disturbances well.

Figure 13.3 illustrates active constraint regions with respect to throughput for a process with 5 steady state DOFs. The active constraints divide the entire throughput range into three regions corresponding to low (2 active constraints), intermediate (3 active constraints) and high throughputs (4 active constraints). At the maximum achievable throughput (5 active constraints), all the steady state DOFs are used up to drive as many constraints active in this hypothetical example. Alternatively, one may have unconstrained DOFs remaining at maximum throughput (i.e. throughput decreases on moving the unconstrained variable away from its optimum value).

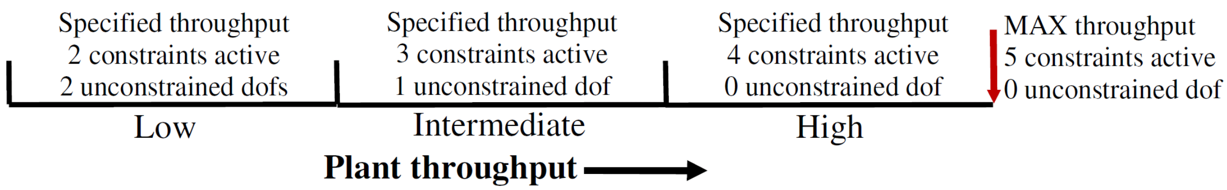

Figure 13.3. Active constraint regions with respect to throughput

Let us assume that the full active constraint set, corresponding to maximum throughput operation, does not change for a given process [a]. To design a truly top-down control system where economic objectives are given the highest priority, loops for the tightest possible control of all the active constraints would first be designed. We would then have the fewest number of control valves left for process regulation, specifically material (total, component and phase) and energy inventory control of the different units and the plant as a whole. If we can achieve effective inventory regulation for maximum throughput operation along with the tightest possible control of the economic CVs, the control system would most certainly work at lower throughputs with additional DOFs (setpoints) available for control due to constraints becoming (optimally) inactive. The reason we emphasize tight economic CV control at maximum throughput is that this is where the economic benefits of improved operation are usually the largest.

13.4. Systematic Control System Design Procedure

Based on the above arguments, the two-tier plantwide control system design framework is modified to designing a robust control system for process operation at maximum achievable throughput with tight economic CV control, arguably the most difficult to stabilize due to the highest number of active constraints, and then designing loops for taking up additional control tasks using constraints (setpoints) that become optimally inactive at lower throughputs. The additional control task may be economic CV control or throughput manipulation A step-by-step 'top-down' procedure for designing the overall control system for near optimum operation over a wide throughput range is then:

Step 0: Obtain active constraint regions for the wide throughput range
Step 1: Pair loops for tight control of economic CVs at maximum throughput
Step 2: Design the inventory (regulatory) control system
Step 3: Design loops for ‘taking up’ additional economic CV control at lower throughputs along with appropriate throughput manipulation strategy
Step 4: Modify structure for better robustness / operator acceptability

Each of these distinct steps is now elaborated upon.

[a] This appears to be a reasonable assumption.

13.4.1. Step 0: Obtain active constraint regions for the wide throughput range

Steady state optimization of the available steady state DOFs is performed to obtain the expanding set of active constraints with increasing throughput. A wide throughput range, from below design throughput to the maximum achievable, is considered. The active constraints partition the entire throughput range into distinct regions. To assess the economic impact of a back-off in any hard active constraints, obtain the economic sensitivity of the hard active constraints at maximum throughput, which corresponds to the full active constraint set. The sensitivities dictate the prioritization as to which constraints must be controlled the tightest.

Corresponding to the unconstrained DOFs in an active constraint region (including maximum throughput), propose self-optimizing CVs that give near-optimal operation with constant setpoint. Sometimes such self-optimizing CVs are not forthcoming. This is acceptable with the implicit understanding that these setpoints are adjusted by a real-time optimizer.

13.4.2. Step 1: Pair loops for tight maximum throughput economic CV control

The economic CVs at maximum throughput are all the active constraints (full active constraint set) and self-optimizing CVs corresponding to any unconstrained steady state DOFs. Typically constraints on maximum allowable product impurity, maximum allowable effluent discharge etc. would be active along with hard capacity constraints such as column operation at flooding limit, furnace operation at maximum duty etc. The full active constraint set may include direct MVs (e.g. a fully open valve). Direct MVs that are optimally at a constraint limit should be left alone at the limit and not used for conventional control tasks. Other active output constraints should be selected as CVs and tightly controlled using close-by MVs that are not active (saturated). For direct MV active constraints, the back-off is then eliminated while for active output constraints, the back-off is mitigated by the tight control.

After implementing loops for tight active constraint control (including leaving a direct MV at its limit), design loops for tight control of self-optimizing CVs. The economic optimum with respect to these unconstrained variables is often 'flat' so that the economic penalty for small deviations from the optimum setpoint is likely to be smaller than for a back-off from an active constraint limit. The loops for self-optimizing CV control are therefore implemented only *after* the loops for tight active constraint control. The flexibility in the input-output (IO) pairings then gets utilized for the tightest control of the economically most important CVs.

There may be situations where the best self-optimizing CV exhibits extremely slow and difficult dynamics. The control implementation may then be decomposed into a faster loop that controls a dynamically better behaved close-by secondary CV, which is not the best self-optimizing CV, with a cascade loop above adjusting its setpoint to ensure that the best self-optimizing CV is maintained close to its (optimum) setpoint over the long-term.

We also note that economic optimality usually requires maximizing reactive inventory hold up, for example, liquid (gas) phase reactor operation at maximum level (pressure). The best pairings for tight control of these inventories should be implemented in this step itself with the remainder of the inventory control system being synthesized in the next step (Step 2).

13.4.3. Step 2: Design the inventory (regulatory) control system

Given loops for tight economic CV control at maximum throughput, implement appropriate loops for consistent inventory control [4] of the different units and the overall process. Inventory is interpreted in its most general sense to include total amount of material, phases (e.g. liquid or vapour), components as well as energy held within the individual units and the overall process. Ensuring consistency of the inventory control system then accounts for tricky regulatory plantwide issues such as the snowball effect due to the integrating nature of component inventories in recycle systems. As recommended in Luyben et al. [14, 16], a 'Downs Drill' must be performed to ensure the control system guarantees that no chemical component (and energy) builds up within the process.

We note that processes are designed with sufficient number of surge capacities to smoothen flow imbalances and facilitate start-up / shut-down. Thus, even if all steady state DOFs are exhausted at maximum throughput to drive as many constraints active, these surge capacities with their associated independent control valves ensure availability of control valves for inventory regulation. An example is a simple distillation column with two steady state DOFs and five control valves (excluding feed). Let us say that to minimize energy consumption, the light key and heavy key in respectively the bottoms and distillate should be at their maximum limits. The 2 steady state DOFs thus get exhausted in driving as many constraints active. If two valves (e.g. reflux and reboiler steam) are paired for maintaining the light-key and heavy key impurities in the two product streams at their maximum limits, three valves (e.g. distillate, bottoms and condenser duty) remain available for controlling the three inventories (reflux drum level, bottom sump level and column pressure).

In a top-down sense, inventory regulation (stabilization) is a lower objective than economic control. The economic CV control loops are therefore put in place first (Step 1) followed by the inventory control system (Step 2). In the inventory loops, local unit specific pairings should be used \to the extent possible. However since valves already paired in Step 1 for tight economic CV control are unavailable, some of the inventory loop pairings may possibly be unconventional non-local 'long' loops.

It is important that, at least in the first pass, a truly 'top-down' plant-wide control structure with such unconventional inventory loops be synthesized. In situations where the inventory control turns out to be fragile due to these unconventional loops, the economic CV loop and inventory loop pairings can always be appropriately revised (this is Step 4 of the procedure). Many a times, these unconventional and seemingly unworkable inventory loops actually work surprisingly well in practice. An example is bottom sump level control of a column with a very small bottoms stream, akin to a leak compared to the internal column flows. Conventional wisdom would suggest using such a leak stream for bottoms level control is unworkable and therefore ill-advised. If however a stripping section tray temperature is well controlled e.g. by adjusting the boilup or feed, the seemingly unworkable pairing provides acceptable sump level control [25]. Level control would be lost only when the temperature loop is put on manual. In our opinion, the unconventional level controller pairing is acceptable with the caveat that the stripping temperature loop be viewed as part of the overall inventory control system and never put on manual. One of the case-studies provides another example where an unconventional inventory control loop pairing works surprisingly well.

13.4.4. Step 3: Design loops for additional economic CV control at lower throughputs along with throughput manipulation strategy

In the control structure for process operation at maximum throughput, one setpoint (TPM) must be used to reduce the process throughput below maximum. Usually, the setpoint for the last constraint to go active is an immediate choice for the TPM. Moving this TPM setpoint away from its active constraint limit would reduce the throughput. As throughput is reduced, additional active constraints become optimally inactive, typically, one after the other. The unconstrained setpoints of the corresponding constraint controllers are now MVs that may be used to control additional self-optimizing CVs for near-optimal operation at lower throughputs. For dynamic reasons, the new CV should be close to the MV (constraint controller setpoint) that becomes available. If such a close-by pairing is not forthcoming, the new unconstrained setpoint may alternatively be considered for use as the TPM in that active constraint region, while using the 'old' TPM (from the more constrained higher throughput region) to control the new CV. The best throughput manipulation strategy across the wide throughput range would then depend on the specific full active constraint set.

To develop such a scheme, list the MV setpoints that become unconstrained along with close-by CVs whose control can be taken-up for more economical operation. Usually, conventional control tasks are best taken up by these MV setpoints. An example is a column moving away from its flooding limit and the resulting unconstrained boilup (MV) taking up column tray temperature control for better energy efficiency. In this list, the unconstrained MV setpoint that gives the dynamically poorest economic CV control may be used as the TPM. In the special case where this MV setpoint is the last constraint to go active and its optimal variation with throughput is monotonic, this single setpoint can be used as the TPM over the entire throughput range. If optimality requires holding this MV setpoint constant in a lower throughput region, the TPM must be shifted to the setpoint of the constraint variable that becomes inactive in that lower throughput region. The shifting may have to be repeated depending on the nature of the next constraint that goes inactive on decreasing throughput.

Referring back to Figure 13.3, we note that the next constraint to become active as throughput is increased can always be used as the TPM in that operating region. If we keep shifting the TPM to the next constraint to go active as throughput is increased, the back-off from the active constraint limit is mitigated. In particular, using the unconstrained setpoint of a constraint control loop as the TPM allows the setpoint to be left closest to its active limit with the least back-off. If the constraint is economically dominant (i.e. large economic penalty per unit back-off), both throughput manipulation and reduced economic penalty due to mitigated back-off get achieved. Another pairing possibility that allows the same is using the unconstrained setpoint of the constraint control loop to control a self-optimizing CV, and not a critical CV such as product quality (critical for economic reasons) or a process inventory (critical for process stabilization). When the constraint limit is reached (e.g. when throughput is increased), control of the non-critical self-optimizing CV is simply given up and the constraint variable setpoint is left closest to the constraint limit with the least back-off. In the special case where the active constraint is a saturated valve, the valve gets left at its saturated position with no back-off.

The point is that there is nothing sacrosanct about fixing the TPM location, although it may be desirable that operators have a single handle to adjust the throughput. This flexibility should be gainfully exploited for eliminating / mitigating the back-off in economically dominant active constraints, obtaining pairings for tight control of the additional unconstrained economic

CVs at lower throughputs as well as simplifying the overall plantwide control system. The throughput manipulation strategy is therefore best considered along with the additional unconstrained economic CV loop pairings in a single step. The best throughput manipulation strategy usually becomes self evident in light of the particular full active constraint set.

13.4.5. Step 4: Modify structure for better robustness / operator acceptance

The control structure obtained from Step 1-3 corresponds to a fully top-down design approach where tight economic CV control at maximum throughput is given precedence over regulatory inventory control, for which control valves are typically available by the design of the process. Through carefully chosen input-output (IO) pairings, the structure attempts to transform all the process variability to the surge capacities and utilities, while maintaining economic CVs at their constrained / optimum setpoints. In such a structure, we may have inventory control loops that are quite unconventional with long loops across units. These may result in fragile inventory (including energy inventory) control.

A surge drum overflowing or drying for even moderately large flow disturbances is a typical result of inventory control fragility. Another example is temperature control of a highly exothermic CSTR with maximum reactor cooling duty being an active constraint. If the cooling duty is left alone at maximum (as it is active) and the CSTR temperature is controlled using the reactor feed, there is the possibility of a thermal runaway with reactants slowly building up inside the reactor when the temperature is below setpoint and the accumulated reactants lighting up once the temperature starts to rise back-up due to the exponential dependence of reaction rate on temperature. The energy inventory inside the reactor then blows up, which is unacceptable. The IO pairings must then be revised to improve inventory control robustness.

To revise the pairings, in the control structure obtained for maximum throughput operation (Step 1-3), tight control of one or more economic CVs must first be given up to free appropriate control valves that then get paired for robust / conventional inventory control. The valves (or setpoints) that become available in lieu may be used for less tight or loose control of the economic CVs whose control was earlier given up. In this exchange of economic CV and unconventional inventory loop MVs for a more robust / conventional inventory control system, it is preferable that the economic CV with the least economic impact (lowest sensitivity) be used to minimize the economic penalty. Instead of unconventional 'long' inventory loops, the revised structure would then have more conventional inventory loops with 'long' economic CV loops.

In most chemical processes, only a few active constraints are dominant with a large economic penalty per unit back-off. With appropriate iteration between Step 1-3, it should be possible to synthesize a control system for tight control of the few dominant active constraints with a not-too-unconventional (i.e. acceptable) and robust inventory control system along with well-behaved additional unconstrained economic CV loops at lower throughputs.

The application of the systematic approach for economic plantwide control system design is demonstrated on four realistic process examples. The first example process is a hypothetical reactor-separator-recycle process with side reaction. The second example process is a C_4 isomerization process. The ethyl benzene manufacturing process is the third example considered. We finally consider two alternative processes for cumene manufacture.

Chapter 14. Economic Plantwide Control of Recycle Process with Side Reaction

14.1. Process Description

The process flowsheet is shown in Figure 14.1 and consists of a cooled liquid phase CSTR followed by a stripper and a distillation column. The main reaction A + B → C and the minor side reaction C + B → D occur in the CSTR. Reaction kinetics and other modelling details are available in Jagtap et al. [7]. The unreacted A and B in the reactor effluent are stripped, condensed and recycled along with some C. The stripper bottoms is fractionated to recover 99% pure C as the distillate (main product) and D with some C as the bottoms (side product). The process has 7 steady state DOFs (2 fresh feeds, reactor level and temperature, 1 stripper DOF and 2 column DOFs) and there are 13 independent control valves. Thus even if all steady state DOFs are exhausted at maximum throughput, 6 valves would still remain available for dynamic control, including inventory control.

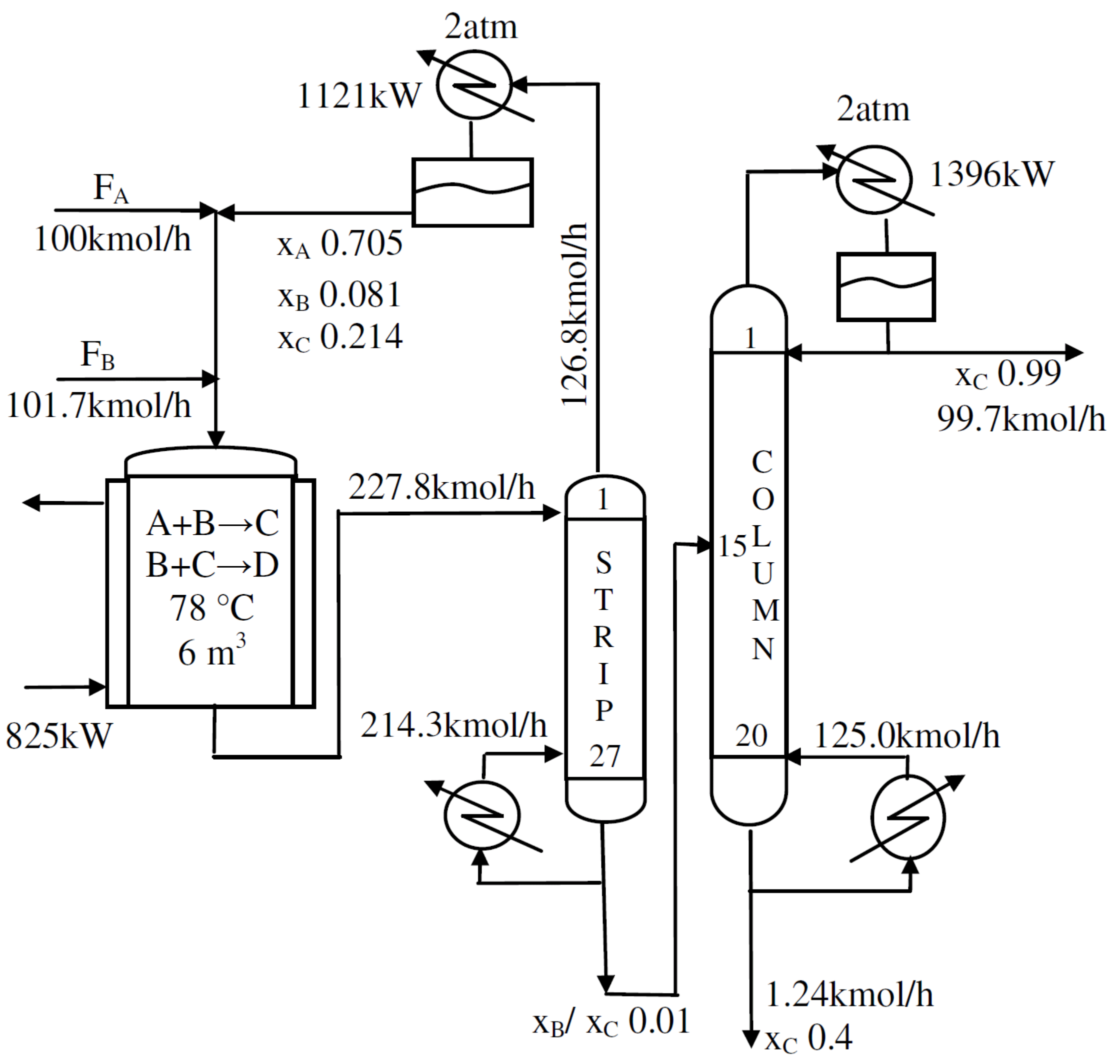

Figure 14.1. Schematic of recycle process with design and base operating conditions

14.2. Economic Plantwide Control System Design

Table 14.1 neatly summarizes the step-by-step implementation of the four-step economic plantwide control system design procedure to this process. A reasonably detailed explanation of the steps is provided in the following.

Table 14.1. Economic Plantwide Control Structure Synthesis for Recycle Process

Step 0: Active Constraint Regions and Economic CV's				
Region	I	II	III	Max Throughput
Additional Active Constraints*	-	V_1^{MAX}	V_1^{MAX} T_{Rxr}^{MAX}	V_1^{MAX} T_{Rxr}^{MAX} V_2^{MAX}
Unconstrained DOF's	2	1	0	0
Self-Optimizing CV's	x_B^{Rxr}, T_{Rxr}	x_B^{Rxr}	-	-
Step 1: Maximum Throughput Economic Control Loops				
Active Constraint Control Loops	$T_{Rxr}^{MAX}\leftrightarrow Q_{Rxr}$	$V_1^{MAX}\leftrightarrow Q_{Reb1}$	$V_2^{MAX}\leftrightarrow Q_{Reb2}$	$T_S^{Col}\leftrightarrow B_1$
	$x_B^{ColD}\leftrightarrow T_{Stp}^{SP}\leftrightarrow F_{Stp}^{SP}$	$x_D^{ColD}\leftrightarrow L_2/B_1^{SP}\leftrightarrow L_2^{SP}$		$LVL_{Rxr}^{MAX}\leftrightarrow F_{Tot}^{Rxr}\leftrightarrow F_A$
Self-Optimizing Loops	none			
Step 2: Maximum Throughput Inventory Loops				
$LVL_{Reb2}\leftrightarrow B_1$	$LVL_{Cnd1}\leftrightarrow F_{Rcy}$	$P_{Cnd1}\leftrightarrow Q_{Cnd1}$		
$LVL_{Reb1}\leftrightarrow x_B^{Rxr\ SP}\leftrightarrow(F_B/F_{Tot}^{Rxr})^{SP}\leftrightarrow F_B$	$LVL_{Cnd2}\leftrightarrow D_2$	$P_{Cnd2}\leftrightarrow Q_{Cnd2}$		
Step 3: Additional Self-Optimizing CV Loops at Reduced Throughput				
Region III	Region II	Region I		
TPM: V_2^{SP}	TPM: V_2^{SP} $x_B^{Rxr\ SP\#}\leftrightarrow(F_B/F_{Tot}^{Rxr})^{SP}\leftrightarrow F_B$ $LVL_{Reb1}\leftrightarrow T_{Rxr}^{SP}\leftrightarrow Q_{Rxr}$	TPM: V_2^{SP} $T_{Rxr}^{SP\#}\leftrightarrow Q_{Rxr}$ $x_B^{Rxr\ SP\#}\leftrightarrow(F_B/F_{Tot}^{Rxr})^{SP}\leftrightarrow F_B$ $LVL_{Reb1}\leftrightarrow V_1^{SP}\leftrightarrow Q_{Reb1}$		
Step 4: Modifications for Conventional Inventory Control Loop				
$LVL_{Reb1}\leftrightarrow B_1$; $T_S^{Col2}\leftrightarrow V_2^{SP}$(with sufficient back-off in V_2^{SP})				
Region III	Region II	Region I		
TPM: $x_B^{Rxr\ SP}$	TPM: T_{Rxr}^{SP} $x_B^{Rxr\ SP\#}$	TPM: V_1^{SP} $T_{Rxr}^{SP\#}$ $x_B^{Rxr\ SP\#}$		

*: LVL_{Rxr}^{MAX}, x_B^{ColD}, x_D^{ColD}, T_S^{Col} are always active; #: Set point value is the optimized value

14.2.1. Step 0: Active Constraint Regions and Economic Operation

To avoid product give-away, the product C impurity mol fractions are fixed at their specified upper limits of 0.98% B (x_B^{ColD}) and 0.02% D (x_D^{ColD}) for the desired 99 mol% pure C (x_C^{ColD}) product. At maximum throughput, the active constraints are maximum column boilup (V_2^{MAX}), reactor temperature (T_{Rxr}^{MAX}), stripper boilup (V_1^{MAX}) and reactor level (LVL_{Rxr}^{MAX}). Further, to prevent loss of precious C with the side product, the average temperature of three adjacent sensitive stripping trays (T_S^{Col}) is maintained [a]. The four equipment capacity constraints, the two product impurity mol fractions and the product column stripping section temperature specification exhaust all 7 steady state DOFs.

At lower throughputs, it is economically near optimal to hold the two product impurity mol fractions and the column stripping section temperature at their maximum throughput values. Also, the LVL_{Rxr}^{MAX} constraint is active at all throughputs as it maximizes the reaction conversion at a given reactor temperature. As throughput is reduced below maximum, the capacity constraints become optimally inactive in the order V_2^{MAX}, T_{Rxr}^{MAX} and V_1^{MAX}. The entire throughput range thus gets partitioned into three active constraint regions (see Table 14.1, Step 0). The number of unconstrained steady state DOFs corresponding to the low throughput (only LVL_{Rxr}^{MAX} active), intermediate throughput (LVL_{Rxr}^{MAX} and V_1^{MAX} active) and high throughput (LVL_{Rxr}^{MAX}, V_1^{MAX} and T_{Rxr}^{MAX} active) regions is respectively, 2, 1 and 0. The V_2^{MAX} constraint going active represents the loss of DOF corresponding to specifying the throughput. The process throughput is then determined by the actual 7 equality / inequality constraint variable values. Jagtap et al. [11] have shown that in the low throughput region, holding the reactor temperature (T_{Rxr}) and the CSTR inlet B (limiting reactant) concentration (x_B^{Rxr}) at appropriate constant values provides near-optimal steady operation. In other words, T_{Rxr} and x_B^{Rxr} are self-optimziing CVs corresponding to the two unconstrained DOFs. In the intermediate throughput region, holding x_B^{Rxr} constant ensures near optimum steady operation (T_{Rxr}^{SP} is not held constant and adjusted for either active constraint control or throughput manipulation). In the high throughput region, there are no unconstrained steady state DOFs left.

14.2.2. Step 1: Loops for Tight Control of Full Active Constraint Set

We now design the control system for maximum throughput operation, where all constraints in the full active constraint set are active. At maximum throughput, there is no TPM as all steady state DOFs are exhausted implying the DOF related to throughput is used for active constraint control. V_2^{MAX} and V_1^{MAX} are active hard constraints with significant economic penalty. Any back-off from V_2^{MAX} causes a large loss in throughput and any back-off in V_1 causes a reduction in the recycle rate and hence a loss in selectivity. Accordingly, V_1 and V_2 are controlled tightly using the respective reboiler steam valves. The back-off necessary from V_1^{MAX} and V_2^{MAX} is then almost negligible.

It is economically important to have tight control of the impurities in the product. The product impurity D mol fraction (x_D^{ColD}) is controlled using the column reflux. The composition controller manipulates the reflux-to-feed ratio setpoint [b]. Maintaining product impurity B mol

[a] This ensures that C composition in the byproduct stream remains small

[b] In practice, the composition controller would cascade a setpoint to a rectifying tray temperature controller which manipulates the L/F ratio setpoint.

fraction (x_B^{ColD}) requires tight control of the B dropping down the stripper as all of it ends up in the product. Since V_1^{MAX} is active, V_1 cannot be used for stripper tray temperature control. The stripper temperature (T_{Stp}) controller then manipulates the stripper feed (F_{Stp}), which provides tight temperature control. The temperature setpoint is adjusted by a cascade x_B^{ColD} controller.

LVL_{Rxr}^{MAX} and T_{Rxr}^{MAX}, the other active equipment capacity constraints imply LVL_{Rxr} and T_{Rxr} must be controlled tightly. Controlling LVL_{Rxr} and T_{Rxr} (at their maximum limits) would also stabilize the reactor material and energy inventories, respectively. For tight control, T_{Rxr} is controlled using reactor cooling duty (Q_{Rxr}), the MV with the best dynamic response (fast dynamics and high open loop gain). We assume T_{Rxr}^{MAX} to be a soft constraint and set $T_{Rxr}^{SP} = T_{Rxr}^{MAX}$. The orientation of the reactor level controller must be opposite to process flow since the reactor effluent (F_{Stp}) is already paired for stripper temperature control. The total flow to the reactor (F_{Tot}^{Rxr}) is a good MV for tight reactor level control. Accordingly, LVL_{Rxr} is controlled by adjusting $F_{Tot}^{Rxr\ SP}$, which in turn is maintained by manipulating the fresh A feed (F_A).

Lastly, it is economically important to maintain an appropriate column stripping section temperature (T_S^{Col}) to ensure loss of precious C in the bottoms is kept small. The active V_2^{MAX} constraint implies column boilup is unavailable for temperature control. Accordingly, the column feed (B_1) is manipulated for the purpose. The active constraint control loops are shown in Figure 14.2. The constrained setpoints at maximum throughput are highlighted in brown.

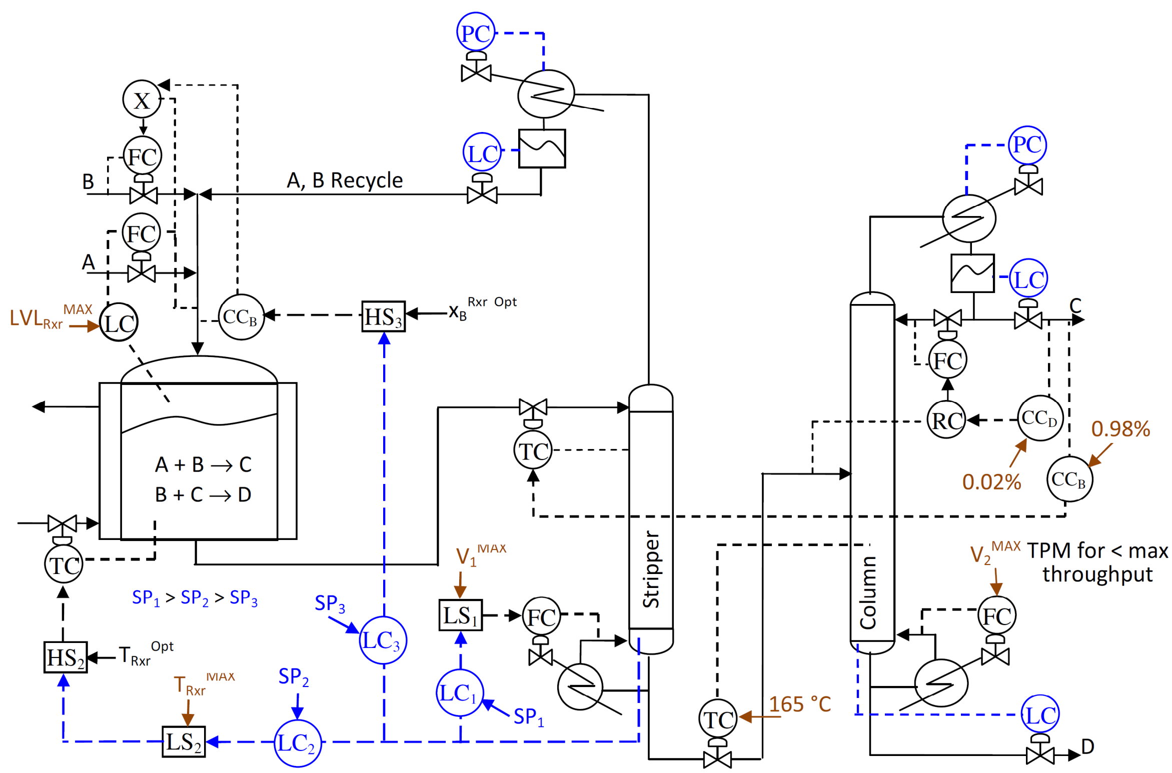

Figure 14.2. Plantwide control structure for maximum throughput operation of recycle

14.2.3. Step 2: Inventory (Regulatory) Control System

Control loops to stabilize the liquid, vapour and component inventories in the process are now implemented using the available unpaired valves (reactor level and energy is already stabilized by the LVL_{Rxr} and T_{Rxr} loops). The inventory loops are shown in blue in Figure 14.2. We need to control the column reflux drum and sump levels, the stripper sump level and the recycle condenser level. The column and the recycle condenser pressures also need to be controlled.

The existing loops for tight active constraint control in Figure 14.2 imply obvious loop pairings for inventory control. The column reflux drum level (LVL_{Cnd2}) is controlled using the distillate (D_1). The recycle and column condenser pressures (P_{Cnd1} and P_{Cnd2}) are controlled using the respective cooling duty valves (Q_{Cnd1} and Q_{Cnd2}). The column sump level (LVL_{Bot}) is controlled using the feed from the stripper (B_1). To mitigate transients in the reactor composition, F_B is maintained in ratio with F_{Tot}^{Rxr}. To ensure A or B component inventory does not build up inside the recycle loop (snowball effect), the B mol fraction in the reactor inlet (x_B^{Rxr}) is maintained by adjusting the F_B to F_{Tot}^{Rxr} ratio setpoint ($F_B/F_{Tot}^{Rxr\ SP}$).

With these pairings, no close-by valves are left for controlling stripper sump level (LVL_{Stp}). The only available option is to adjust the $x_B^{Rxr\ SP}$. The pairing makes sense in that the reaction products accumulate in the stripper sump for downstream separation. The sump level is then an indirect indication of the reactor production rate. If this level is falling, the reactor production needs to be increased. Increasing the $x_B^{Rxr\ SP}$ causes the limiting reactant B composition in the reactor to increase with consequent increase in generation of product C and hence in the stripper sump level.

The stripper level controller is the most unconventional in the scheme. Will it work in practice? That depends on the hold up in the CSTR. If the reactor is too big, the dynamic effect of a change in the $x_B^{Rxr\ SP}$ on stripper sump level would be slow and it may run dry or overflow during worst case transients. The robustness of the control system is tested for a ±5 5% step bias in the F_B sensor (control system tuning details in Appendix A). In the transient response, all the levels are well controlled with the maximum deviation in the stripper sump level being < 4%. The inventory control scheme, though unconventional, is quite robust and acceptable.

14.2.4. Step 3: Additional Economic CV Control Loops and Throughput Manipulation

At lower throughputs, the additional unconstrained economic CVs whose control must be taken up are x_B^{Rxr} and T_{Rxr}. Both are associated with the reactor. Since maximum column boilup (V_2^{MAX}) is the last constraint to go active and its optimal variation with throughput is monotonic, we consider using it as the TPM over the entire throughput range. Now as V_2^{SP} is reduced below V_2^{MAX}, the production rate would decrease below maximum with x_B^{Rxr} reducing. The excess A inside the reactor then increases to further suppress the side reaction for improved yield to the desired product. When x_B^{Rxr} reduces to its optimal value, it must be held constant for optimal operation. LVL_{Stp} then gets controlled using T_{Rxr}^{SP}, in lieu of x_B^{Rxr}. T_{Rxr}^{SP} would reduce below T_{Rxr}^{MAX} as V_2^{SP} is decreased. When T_{Rxr}^{SP} decreases to its optimum value, it must be held constant. LVL_{Stp} then gets controlled using V_1^{SP} in lieu of T_{Rxr}^{SP}). V_1^{SP} would reduce below V_1^{MAX} as V_2^{SP} is reduced to decrease the throughput. The stripper bottom sump level controller pairing thus switches from $x_B^{Rxr\ SP}$ to T_{Rxr}^{SP} to V_1^{SP} as throughput is reduced. Referring to the throughput regions in Table 14.1, at high throughputs, x_B^{Rxr} floats to the appropriate value determined by V_2^{SP} via the action of the inventory control system. At intermediate throughputs, x_B^{Rxr} is maintained at its optimum and T_{Rxr} floats to the appropriate value. Finally, at low throughputs, x_B^{Rxr} and T_{Rxr} are held at their near optimum values and V_1^{SP} floats to the appropriate value.

A simple override scheme to accomplish the switching between the operating regions with three separate PI stripper sump level controllers (LC_1, LC_2 and LC_3) is shown in Figure 14.2. The MVs for LC_1, LC_2 and LC_3 are respectively, V_1^{SP}, T_{Rxr}^{SP} and $x_B^{Rxr\ SP}$. At maximum throughput, since T_{Rxr}^{MAX} and V_1^{MAX} are active, LC_1 and LC_2 are inactive and sump level control is performed by LC_3. As V_2^{SP} (TPM) is reduced below V_2^{MAX}, LC_3 decreases $x_B^{Rxr\ SP}$. When $x_B^{Rxr\ SP}$ reduces below its optimum value, the high select block, HS_3, passes the optimum value to the x_B^{Rxr} controller. LC_3 then becomes inactive and stripper sump level control is lost. The level then increases beyond LC_2 setpoint and the LC_2 output starts to decrease. When the output decreases below T_{Rxr}^{MAX}, level control is taken over by LC_2. When T_{Rxr}^{SP} decreases below its optimum value, the high select block, HS_2, passes the optimum value and LC_2 becomes inactive and the stripper sump level again rises beyond LC_1 setpoint. LC_1 output then reduces and on decreasing below V_1^{MAX}, the low select block, LS_1, causes LC_1 to take over level control. A complementary logic causes proper switching from LC_1 to LC_2 to LC_3 as throughput is increased.

Note that the decreasing level setpoint order ($LC_1 > LC_2 > LC_3$) is necessary to enforce the proper switching order. For example, when LC_1 is active, the level would be close to LC_1 setpoint and the I action in LC_2 and LC_3 would cause the respective controller output signals to be sufficiently high ensuring the respective (high) select blocks pass the appropriate signal (optimum T_{Rxr}^{SP} and $x_B^{Rxr\ SP}$ respectively). It is also highlighted that in the given scheme, LC_1 is reverse acting and nested with the stripper temperature loop. As LVL_{Stp} decreases, V_1^{SP} increases (reverse action) which causes the stripper temperature to increase. The temperature controller then increases the stripper feed which causes the LVL_{Stp} to return to setpoint.

Rigorous dynamic simulations are performed to test the synthesized control structure in Hysys. Unless specified otherwise, all flow / pressure PI controllers are tuned tight for a fast and snappy servo response. The non-reactive level controllers are P-only with a gain of 2. The only exception is the unconventional stripper sump level controller with overrides. For the three different pairings in the three operating regions, distinct conservative (non-aggressive) tunings are used to dampen flow variability. The CSTR level is controlled using a PI controller for offset free level tracking. The approximate controller tuning is first obtained using the Hysys autotuner and then adjusted for a fast and not-too-oscillatory servo response at maximum throughput. All temperature measurements are lagged by 2 mins to account for sensor and cooling / heating circuit dynamics. To tune the temperature loops, the open loop step response at maximum throughput is obtained and the reset time set to $1/3^{rd}$ of the approximate 95% response completion time. The gain is then adjusted for a slightly underdamped servo response with mild oscillations. The composition controllers are similarly tuned. A sampling time and delay time of 5 mins each is applied to all composition measurements. Salient controller parameters are reported in Table 14.2.

The dynamic response of salient process variables of this control system to a throughput transition from the base-case throughput (F_A = 100 kmol/h) to the maximum throughput (F_A = 188.7 kmol/h) and back is shown in Figure 14.3. Tight product purity control is achieved along with smooth plantwide transients. The control system is also tested for a ±5% step bias in the F_B measurement signal at maximum throughput operation. The dynamic response is plotted in

Figure 14.4. Notice the tight control of the product impurities as well the C loss in the by-product stream. The synthesized plantwide control system is thus suitable for economic process operation across the wide throughput range.

If a conventional control system with the TPM at the fresh feed were to be implemented, the need for a back-off from V_1^{MAX} and V_2^{MAX} during worst case transients results in significant throughput (economic) loss (~4-7%) [8]. The synthesized plantwide control system thus achieves significantly superior economic operation for the same plant equipment.

Table 14.2. Salient controller tuning parameter for recycle process

CV	K_C	τ_i (min)	Sensor Span
x_B^{Rxr}	0.8	400	0 – 1
T_{Rxr}^*	1	10	60 – 130 ºC
LVL_{Rxr}	0.5	25	0-100%
T_{Stp}	0.5	15	100 – 160 ºC
T_S^{Col}	0.6	25	140 – 180 ºC
x_B^{ColD}	0.1	40	0 – 0.02
x_D^{ColD}	0.1	30	0 0.0004
Tuning for LVL_{Reb1} override control			
LVL_{Reb1}^1	0.8	200	0-100%
LVL_{Reb1}^2	0.6	250	0-100%
LVL_{Reb1}^3	0.5	400	0-100%

All level loops use K_C = 2 unless otherwise specified
Pressure/flow controllers tuned for tight control
All composition measurements: deadtime = 5 min; sampling time = 2 min;
*: Derivative action used with τ_D = 2 min
All temperatures measurements lagged by 2 mins
1: MV= V_1; 2: MV= T_{Rxr},; 3: MV= x_B^{Rxr}

14.2.5. Step 4: Modifications for a More Conventional Inventory Control System

Given that the control system works well with the unconventional stripper bottoms level control loop, Step 4 (control system modification for a more conventional inventory control system) is not necessary. It is however instructive to develop a control system with conventional local inventory control loops.

The stripper sump level control loop in Figure 14.2 is arguably the most controversial inventory control loop. For a more conventional local pairing, the column stripping section temperature (T_S^{Col}) loop is broken to free the stripper bottoms valve, which is then paired to control the stripper sump level. T_S^{Col} may then be maintained by adjusting $x_B^{Rxr\ SP}$ in a long loop. Even as the steady state economic penalty with such a long economic loop is small, the penalty during transients is likely to be severe. Due to the V_2^{MAX} active constraint, the precious C that could not be boiled off would accumulate at the bottom of the product column and get discharged in the by-product stream by the action of the column sump level controller. Since the optimum C leakage in the bottom stream is very small to begin with, one would expect transient deviations in the direction of higher than optimum C leakage to be significantly more severe than in the opposite (lower than optimum C leakage) direction, where there is little / no leeway. The long column stripping section temperature loop is then susceptible to large loss of precious C during transients. To mitigate the same, a local temperature control loop is needed. Accordingly,

T_S^{Col} is controlled using the column boilup (V_2^{SP}). For maximum throughput operation without loss of control of C leaking down the product column bottoms, the $x_B^{Rxr\ SP}$ would be set at a value such that V_2^{MAX} constraint is just hit during the worst case transient. The back-off from V_2^{MAX} then represents an unrecoverable economic loss, which is the price that must be paid for a more conventional inventory control system.

In the original control system (Figure 14.2), V_2^{SP} was used as the TPM in all regions. With the revised pairings where V_2^{SP} is used for T_S^{Col} control, an alternative throughput manipulation strategy is needed. To reduce throughput below maximum (Region III), $x_B^{Rxr\ SP}$ gets used as the TPM. Once $x_B^{Rxr\ SP}$ is reduced to its optimum value, the TPM shifts to T_{Rxr}^{SP} which is reduced below T_{Rxr}^{MAX} (Region II). Once T_{Rxr}^{SP} is reduced to its optimum value, the TPM shifts to V_1^{SP}, which is reduced below V_1^{MAX} (Region I). Note that in this TPM shifting scheme, the back-off from V_1^{MAX} is negligible. Also, the transient variability in T_{Rxr} for operation at T_{Rxr}^{MAX} is minimal as T_{Rxr}^{SP} is not adjusted by any master cascade loop once T_{Rxr}^{MAX} is hit. The revised control system is shown in Figure 14.5 (Step 4 in Table 14.1).

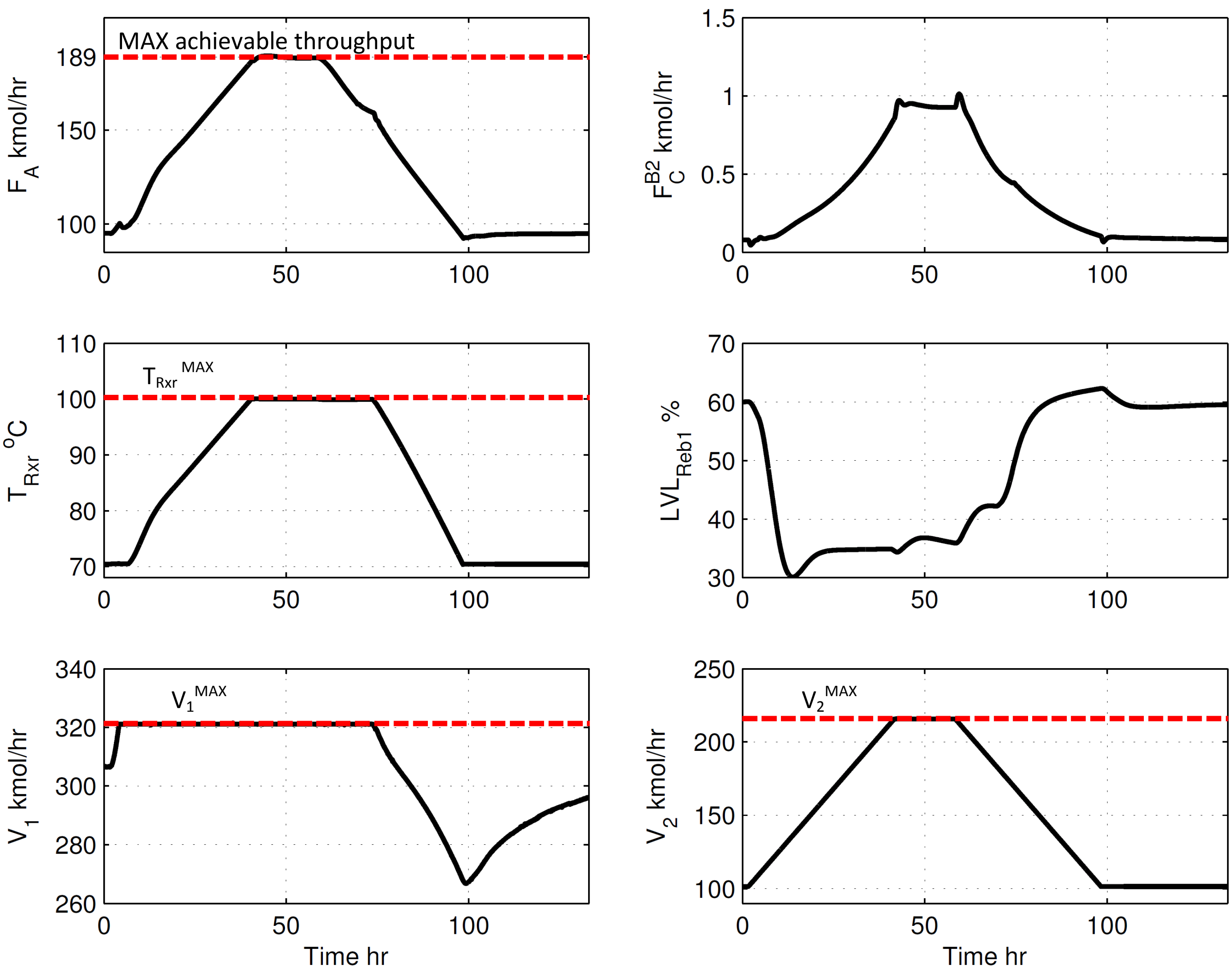

Figure 14.3. Throughput transition with stripper sump level override control scheme

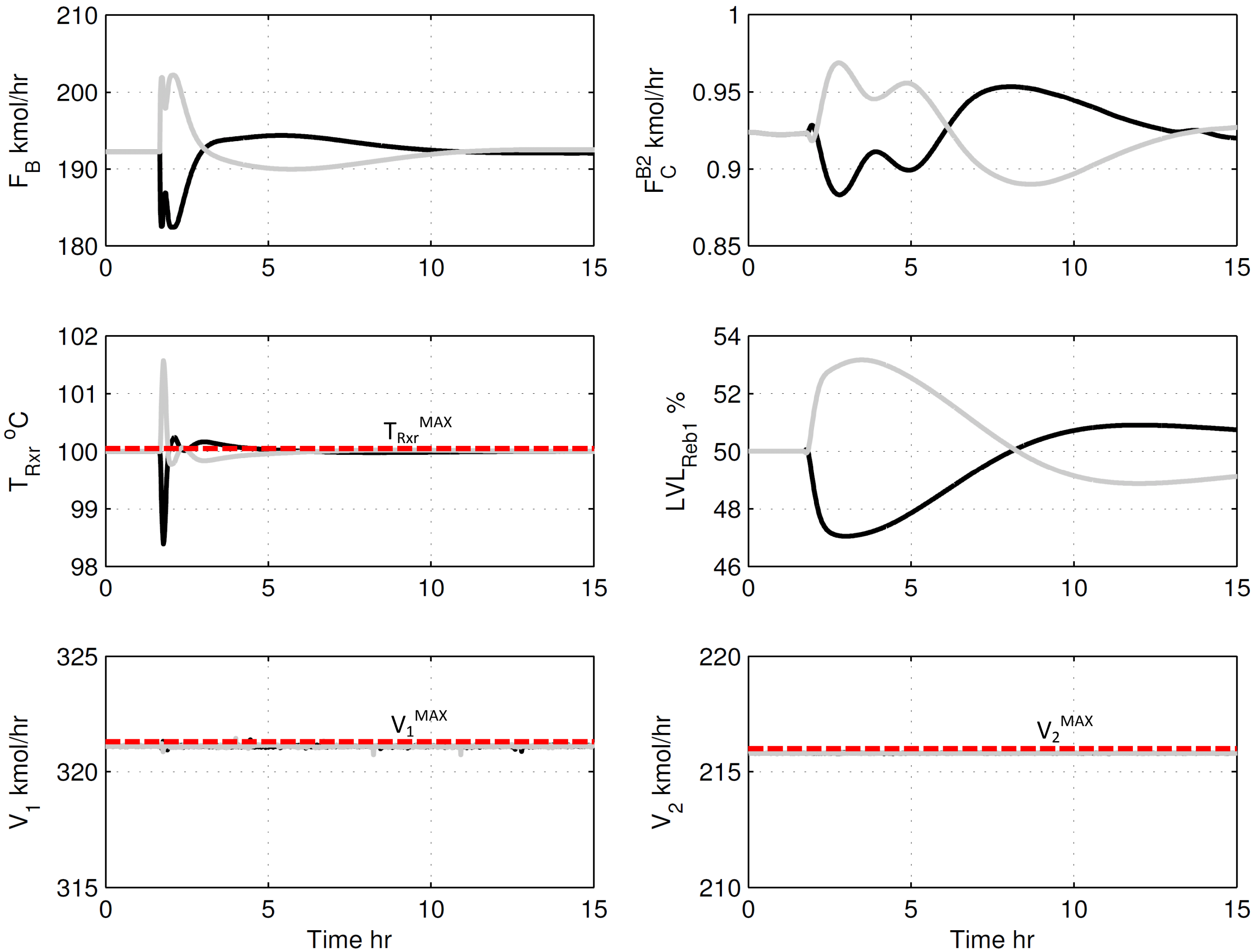

Figure 14.4. Transient response for ±5% step bias in F_B flow sensor

: +5% bias; : -5% bias

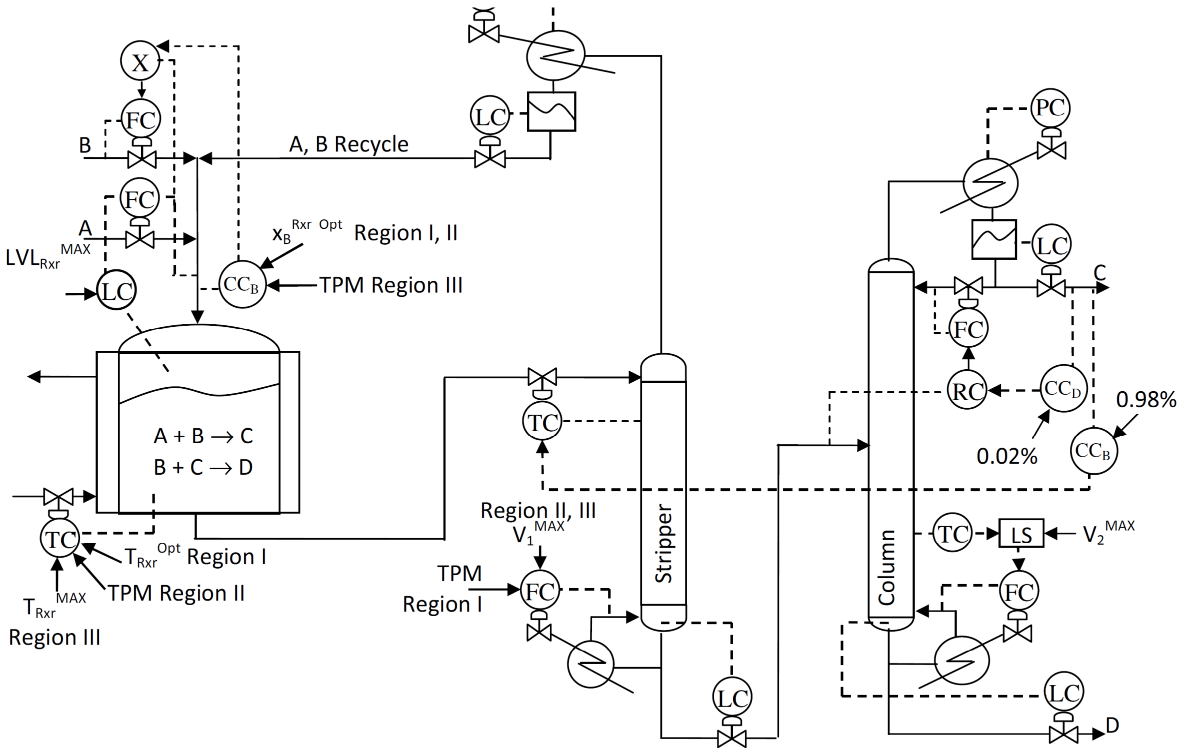

Figure 14.5. Recycle process modified control structure for conventional inventory control

Chapter 15. Economic Plantwide Control of Ethyl Benzene Process

15.1. Process Description

The process consists of two reactors and two columns along with two liquid recycle streams, as in Figure 15.1. The reaction chemistry consists of three reactions

C_6H_6 +	C_2H_4	$\rightarrow$	C_8H_{10}	Main Reaction
Benzene	Ethylene		Ethyl Benzene	
C_8H_{10} +	C_2H_4	$\rightarrow$	$C_{10}H_{14}$	Side Reaction
Ethyl Benzene	Ethylene		Diethyl Benzene	
$C_{10}H_{14}$ +	C_6H_6	$\rightarrow$	$2\ C_8H_{10}$	Transalkylation
Diethyl Benzene	Benzene		Ethyl Benzene	

The reaction kinetics and other modeling details are available in Jagtap and Kaistha [8]. The first two reactions occur primarily in the first coil cooled CSTR while transalkylation primarily occurs in the second adiabatic CSTR. Near complete ethylene conversion occurs in the two CSTRs. The reaction section effluent is fractionated in the recycle column to recover and recycle unreacted benzene back to the first CSTR. The bottoms is fractionated in the product column to recover 99.9 mol% pure ethyl benzene (EB) as the distillate. The diethyl benzene (DEB) drops down the bottoms and is recycled to the second CSTR. The DEB is allowed to build in the recycle loop so that the DEB formation rate by the side reaction exactly matches the DEB transalkylation rate for no net DEB formation. The DEB is thus recycled to extinction.

15.2. Economic Plantwide Control System Design

The step-by-step synthesis of the economic plantwide control system is summarized in Table 15.1. The major steps are briefly described below.

15.2.1. Step 0: Active Constraint Regions and Optimal Operation

With fixed pressures, the process has nine steady state degrees of freedom: 2 fresh feeds, 2 DOFs for the first reactor (level and temperature), 1 for the second reactor (level) and 4 DOFs for the two columns. At maximum throughput, there are 8 active constraints: maximum recycle column boilup (V_1^{MAX}) and reflux (L_1^{MAX}), maximum product column boilup (V_2^{MAX}), first reactor maximum temperature (T_{rxr1}^{MAX}) and level (LVL_{rxr1}^{MAX}), second reactor maximum level (LVL_{rxr2}^{MAX}) plus maximum product impurity levels $x_{Bz}^{D2\ MAX}$ (benzene mol fraction) and $x_{DEB}^{D2\ MAX}$ (DEB mol fraction) for no product give-away. This leaves one unconstrained steady state DOF at maximum throughput, which is related to the optimal DEB recycle (L_1^{MAX} fixes benzene recycle). Of the active constraints, T_{rxr1}^{MAX}, LVL_{rxr1}^{MAX} and LVL_{rxr2}^{MAX} are active regardless of throughputs. As throughput is increased, L_1^{MAX}, V_2^{MAX} and V_1^{MAX} become active, in that order. These three active constraints are treated as hard while the remaining ones are treated as soft.

In this process, unlike previous examples, an unconstrained DOF remains at maximum throughput. The DEB recycle flow rate (B_2) is considered as a self-optimizing CV. We have shown that holding B_2 fixed at its optimal maximum throughput value results in only a maximum 0.35% operating profit loss at lower throughputs [8]. The loss is deemed acceptable and is a consequence of energy being significantly cheaper than products or raw material (Douglas' doctrine [5]). At lower throughputs, overrefluxing in the two columns is mitigated by maintaining L_1 in ratio with the recycle column feed (F_{col1}) and maintaining a sensitive stripping tray temperature (T_S^{col2}) using V_2. The self-optimizing CVs corresponding to unconstrained L_1 and and V_2 are L_1/F_{col1} and T_S^{col2} respectively.

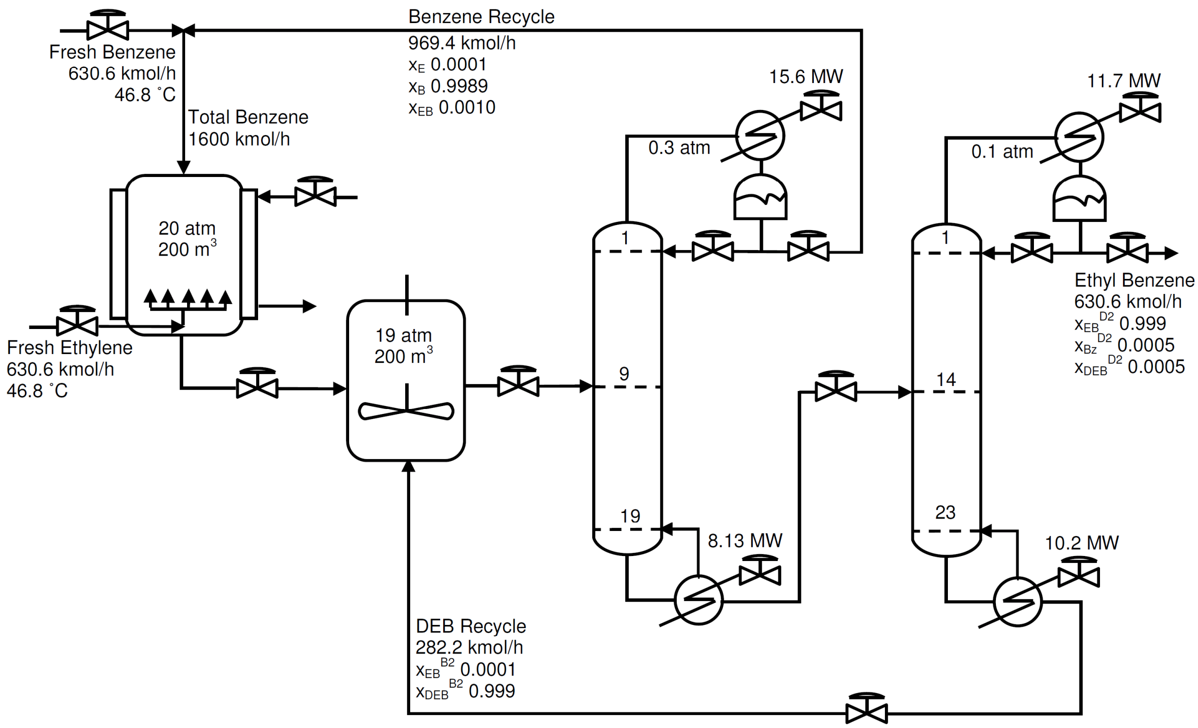

Figure 15.1. Schematic of ethyl benzene process with design and operating conditions

15.2.2. Step 1: Loops for Maximum Throughput Economic CV Control

The full active constraint set consists of LVL_{rxr1}^{MAX}, T_{rxr1}^{MAX}, LVL_{rxr2}^{MAX}, L_1^{MAX}, V_2^{MAX}, V_1^{MAX} $x_{DEB}^{D2\ MAX}$ and $x_{Bz}^{D2\ MAX}$. Of these, L_1^{MAX}, V_2^{MAX} and V_1^{MAX} are hard constraints. For negligible back-off from their hard constraint limits, V_1 and V_2 are controlled using the respective reboiler steam valves (Q_{reb1} and Q_{reb2}) while L_1 is flow controlled. T_{rxr1}^{MAX} is controlled using the reactor cooling duty (Q_{rxr}), a conventional pairing for tight temperature control. For tight control of x_{DEB}^{D2} (product impurity), the column reflux to feed ratio is adjusted. For tight control of x_{Bz}^{D2} (product impurity) another cascade loop arrangement is implemented where the composition controller adjusts a sensitive recycle column stripping tray temperature controller setpoint, which in turn manipulates the column feed (F_{col1}). With the recycle column feed (F_{Col1}) paired for temperature control, the level controllers in the two reactors must be oriented opposite to the process flow. Accordingly, LVL_{rxr2} is controlled using its feed (F_{rxr2}). Similarly, for tight level control of the first reactor (LVL_{rxr1}), the reactor liquid feed (fresh + recycle benzene, F_{TotBz}) is adjusted. F_{TotBz} is maintained by adjusting the fresh benzene so that that the fresh benzene is fed as a make-up stream (Luybens' rule). Lastly, B_2 (self optimizing CV) is flow controlled.

Table 15.1. Economic Plantwide Control Structure Synthesis for Ethyl Benzene Process

Step 0: Active Constraint Regions and Economic CV's				
Region	I	II	III	Max Throughput
Additional Active Constraints*	-	L_1^{MAX}	V_2^{MAX} L_1^{MAX}	V_1^{MAX} V_2^{MAX} L_1^{MAX}
Unconstrained DOF's	3	2	1	1
Self-Optimizing CV's	B_2, L_1/F_1, T_S^{col2}	B_2, T_S^{col2}	B_2	B_2
Step 1: Maximum Throughput Economic Control Loops				
Active Constraint Control Loops	$T_{rxr1}^{MAX} \leftrightarrow Q_{rxr1}$ $x_{Bz}^{D2} \leftrightarrow T_S^{col1\ SP} \leftrightarrow F_{col1}^{SP}$	$V_1^{MAX} \leftrightarrow Q_{reb1}$ $x_{DEB}^{D2} \leftrightarrow L_2/B_1^{SP} \leftrightarrow L_2^{SP}$	$V_2^{MAX} \leftrightarrow Q_{reb2}$	$LVL_{rxr1}^{MAX} \leftrightarrow F_{TotBz} \leftrightarrow F_{Bz}$ $LVL_{rxr2}^{MAX} \leftrightarrow F_{rxr2}$
Self-Optimizing Loops	none			
Step 2: Maximum Throughput Inventory Loops				
$LVL_{cnd1} \leftrightarrow D_1$ $LVL_{cnd2} \leftrightarrow D_2$	$LVL_{reb1} \leftrightarrow F_{C2}/F_{TotBz}^{SP} \leftrightarrow F_{C2}$ $LVL_{reb2} \leftrightarrow B_1$	$P_{cnd1} \leftrightarrow Q_{cnd1}$ $P_{cnd2} \leftrightarrow Q_{cnd2}$		
Step 3: Additional Self-Optimizing CV Loops at Reduced Throughput				
Region III	Region II	Region I		
TPM: V_1^{SP}	TPM: V_1^{SP} $T_S^{col2} \leftrightarrow V_2^{SP\ \#}$	TPM: V_1^{SP} $T_S^{col2} \leftrightarrow V_2^{SP\ \#}$ $L_1/F_1 \leftrightarrow L_1^{\#}$		
Step 4: Modifications for Conventional LVL_{Reb1} Control Loop				
$LVL_{reb1} \leftrightarrow B_1$				
Region III	Region II	Region I		
TPM: V_1^{SP} $B_2 \leftrightarrow F_{TotBz}/F_{C2}^{SP}$	TPM: V_1^{SP} $T_S^{col2} \leftrightarrow V_2^{SP\ \#}$ $B_2 \leftrightarrow F_{TotBz}/F_{C2}^{SP\ \#}$	TPM: V_1^{SP} $T_S^{col2} \leftrightarrow V_2^{SP\#}$ $B_2 \leftrightarrow F_{TotBz}/F_{C2}^{SP} \leftrightarrow F_{C2}$ $L_1/F_1 \leftrightarrow L_1^{\#}$		

*: T_{rxr1}^{MAX}, LVL_{rxr1}^{MAX}, LVL_{rxr2}^{MAX}, $x_{Bz}^{D2\ MAX}$, $x_{DEB}^{D2\ MAX}$ are always active; #: Is unconstrained from MAX limit

15.2.3. Step 2: Inventory (Regulatory) Control System

The remaining inventories to be controlled include the four column levels (LVL_{cnd1}, LVL_{cnd2}, LVL_{bot1}, LVL_{bot2}) and the two column pressures (P_{cnd1} and P_{cnd2}). The column pressures are controlled conventionally using the respective condenser duty valves (Q_{cnd1} and Q_{cnd2}). The reflux drum levels of the two columns (LVL_{cnd1} and LVL_{cnd2}) are controlled using the respective distillate stream (D_1 and D_2). On the product column, since the B_2 is under flow control as a self-optimizing variable and therefore unavailable, the sump level (LVL_{bot2}) is controlled using the product column feed (B_1). This leaves no close-by valves for controlling the recycle column sump level (LVL_{bot2}). The only pairing possibility is to adjust the fresh ethylene feed rate (F_{C2}). To mitigate the transients in the reactor composition, F_{C2} is maintained in ratio with the F_{TotBz} with the LVL_{bot2} controller adjusting the ratio setpoint, $F_{C2}/F_{TotBz}{}^{SP}$. As in the recycle process case study (Case Study 1), this is an unconventional long inventory loop and makes sense in that the reaction products (EB and DEB) accumulate in the bottom sump of the recycle column. LVL_{Bot2} thus indirectly indicates the production rate. A decreasing level implies the reaction production rate must be increased, which is accomplished by increasing F_{C2} (limiting reactant) via appropriate adjustment in $F_{C2}/F_{TotBz}{}^{SP}$ by the level controller.

15.2.4. Step 3: Additional Economic CV Loops and Throughput Manipulation

To reduce throughput below maximum, we consider using V_1^{SP} as the TPM across the entire throughput range as V_1^{MAX} is the last constraint to go active. When optimally inactive, L_1^{SP} is maintained in ratio with the recycle column feed to mitigate overrefluxing in the recycle column [e]. Similarly, V_2^{SP} takes up tight control of a sensitive product column stripping tray temperature, whenever feasible at lower throughputs.

15.2.5. Step 4: Modifications for a More Conventional Inventory Control System

The economic plantwide control structure synthesized by the application of Step 1-3 of our procedure is shown in Figure 15.2. In this control system, we have an unconventional and long loop for controlling the recycle column sump level. For this process, the total reactor residence time is ~2 hrs so that the dynamic response of LVL_{bot2} to a change in $F_{C2}/F_{TotBz}{}^{SP}$ (MV) is quite sluggish resulting in the recycle column sump overflowing or running dry even for the mildest of disturbances such as a 1% step change in B_2^{SP}. Clearly the inventory control system is very fragile so that the economic CV and inventory loop pairings must be appropriately revised.

To revise the pairings, we first consider giving up on tight control of the self-optimizing CV, B_2. The product column sump level (LVL_{bot2}) is then paired with B_2 which frees up the recycle column bottoms flow (B_1) which is then used for robust control of LVL_{bot1}. This frees up $F_{C2}/F_{TotBz}{}^{SP}$ which takes up 'loose' control of the self-optimizing variable, B_2. The long inventory loop, LVL_{bot1} - $F_{C2}/F_{TotBz}{}^{SP}$, in Figure 15.2 (Step 2 row in Table 15.1) thus gets replaced by a long B_2 - $F_{C2}/F_{TotBz}{}^{SP}$ loop after the re-pairing exercise to provide a conventional and robust inventory control system. The revised control system is shown in Figure 15.3.

[e] Alternatively, L_1^{SP} can take up rectifying temperature control for dual ended control.

To transition to lower throughputs, V_1^{SP}, the last constraint to go active is used as the TPM over the entire throughput range. Also, to prevent overrefluxing in the two columns at low throughputs, V_2^{SP} takes up product column stripping tray temperature control and L_1 is maintained in ratio with the recycle column feed (F_{col1}). These two loops take-up control as and when the controller output becomes implementable (i.e. $V_2^{SP} < V_2^{MAX}$ and $L_1^{SP} < L_1^{MAX}$).

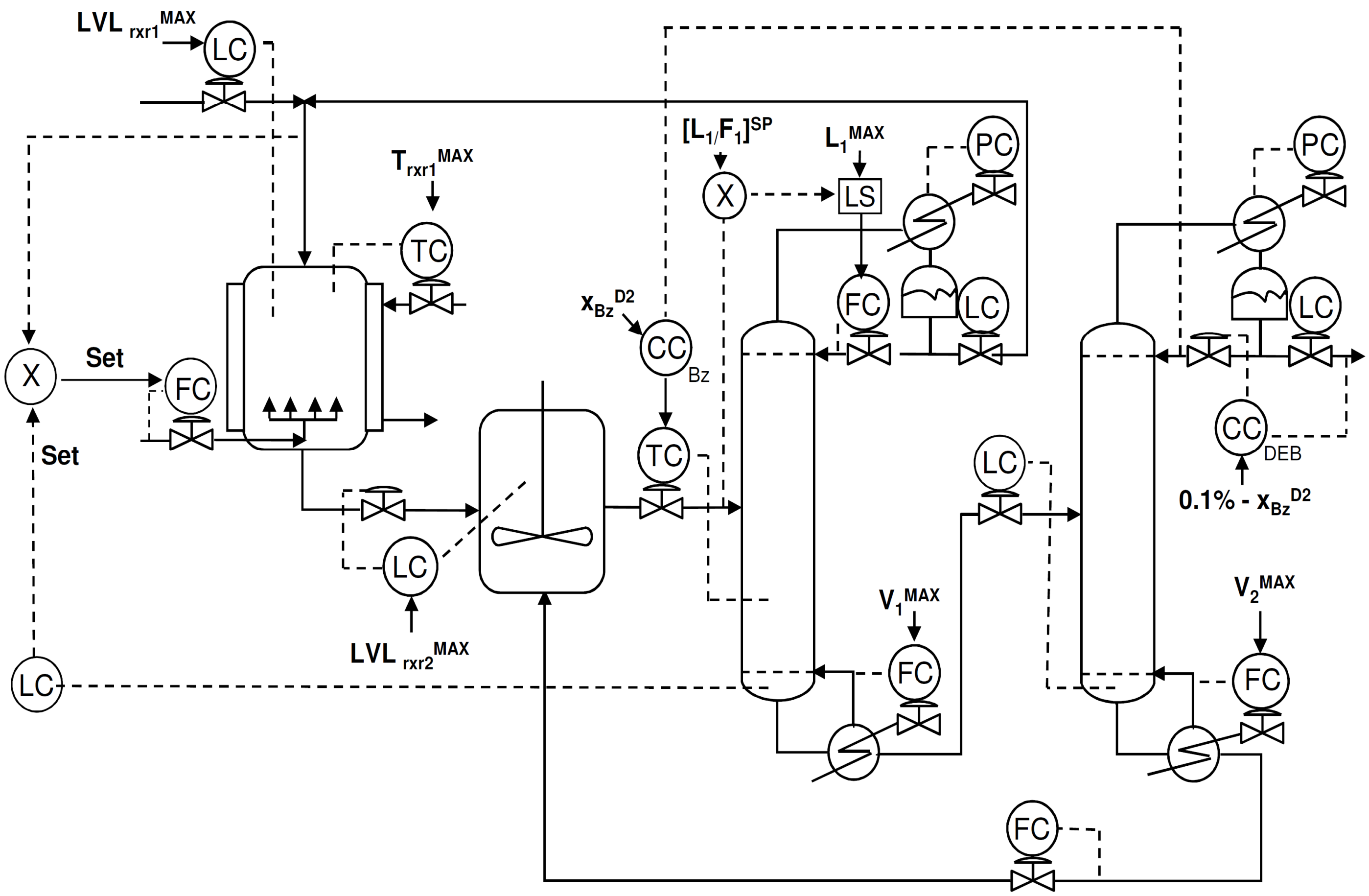

Figure 15.2. Ethyl benzene process economic plantwide control structure (with long inventory loop)

It is highlighted that in the revised pairings for more conventional inventory control (Step 4 in Table 15.1), B_2 must be controlled (by adjusting $F_{C2}/F_{TotBz}{}^{SP}$) and not allowed to float as it can result in a snowballing problem. This is because V_2^{MAX} is an active constraint at maximum throughput implying limited capacity to boil-off EB in the product column. Any EB that could not be boiled off in the product column would necessarily drop down the bottoms causing the DEB recycle rate (B_2) to slowly increase. To prevent this slow drift (snowballing), it must be ensured that only as much EB is produced in the reaction section as can be boiled off in the product column. This gets accomplished by adjusting the $F_{C2}/F_{TotBz}{}^{SP}$ to maintain B_2, which ensures the fresh ethylene feed to the process matches the EB boil-off rate. A seemingly innocuous recommendation of allowing a self-optimizing CV to float and accepting the consequent economic loss results in a very severe consequence of potential process instability. This highlights the importance of Down's drill in ensuring the recommended control structure does not suffer from such hidden instabilities due to slow accumulation of component inventories.

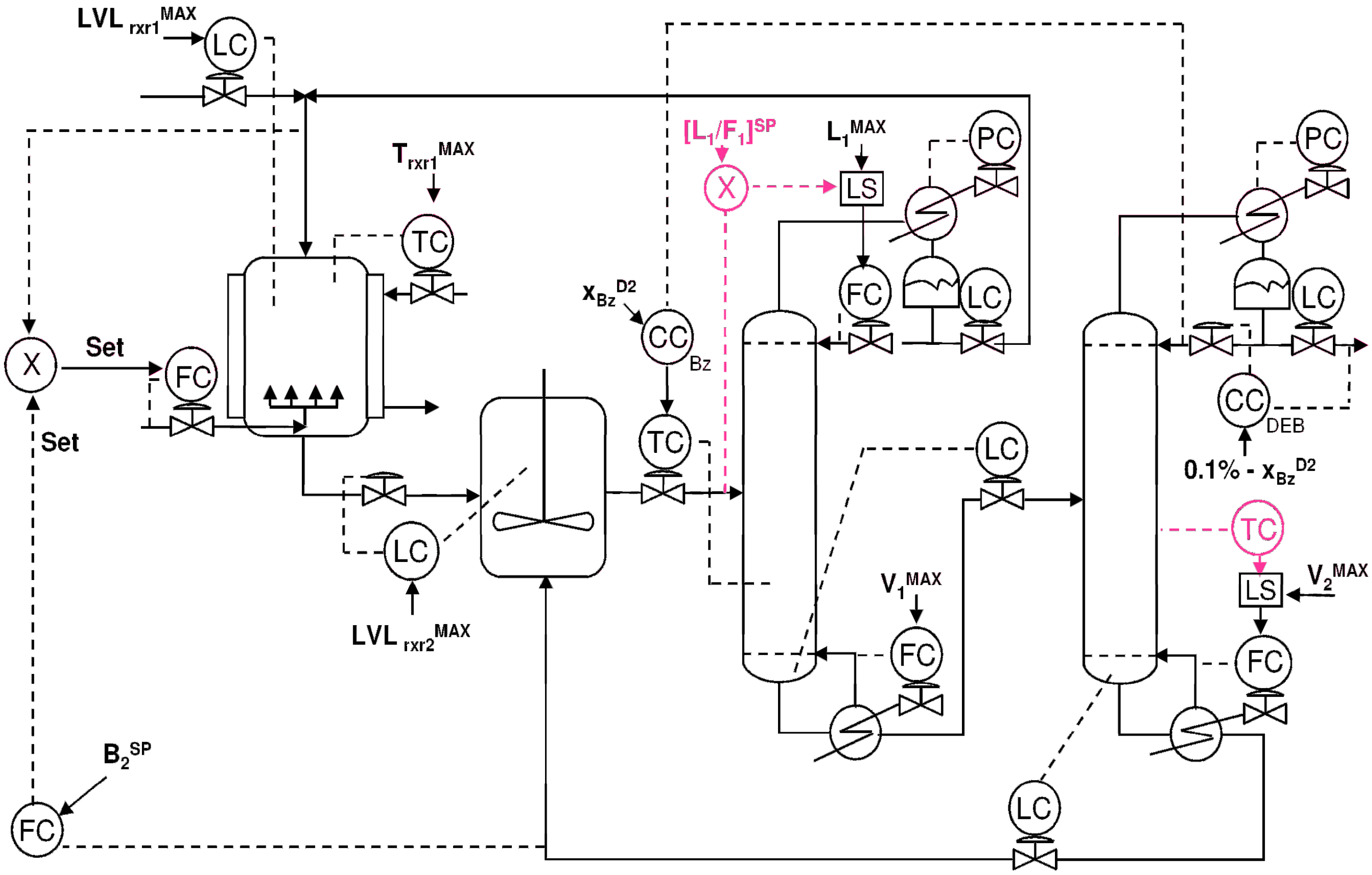

Figure 15.3. Modified economic plantwide control structure for ethyl benzene process

If a conventional control system was designed for process operation around the design condition, V_2 would get used for maintaining a product column stripping temperature. As long as the loop is functioning, the EB would get boiled-off and not accumulate in the DEB recycle loop. However, once V_2^{MAX} goes active, product column stripping temperature control would be lost. To ensure that the process does not succumb to snowballing in the DEB recycle loop, one would have to design an override scheme that alters the material balance structure all the way up to the process feed resulting in an inherently complicated scheme for constraint handling. In contrast, the synthesized control structure is much simpler with no overrides and appealing in that the way inventory is regulated remains the same regardless of the operating region.

Rigorous dynamic simulations are performed to test the synthesized control structure in in Aspen Plus. All flow / pressure PI controllers are tuned tight for a fast and snappy servo response, unless specified otherwise. The long B_2 loop is tuned by hit-and-trial for a smooth overall plantwide response. The non-reactive level controllers are P-only with a gain of 2. The CSTR levels are controlled using a PI controller for offset free level tracking. The relay feedback test feature with Tyreus-Luyben settings is used to obtain the CSTR level controller tuning parameters at maximum throughput. All temperature measurements are lagged by 2 mins to account for sensor and cooling / heating circuit dynamics. To tune the temperature loops, the open loop step response at maximum throughput is obtained and the reset time set to 1/3rd of the approximate 95% response completion time. The gain is then adjusted for a slightly underdamped servo response with mild oscillations. The composition controllers are similarly tuned. A sampling time and delay time of 5 mins each is applied to all composition measurements. The tuning parameters of salient loops are reported in Table 15.2.

The closed loop dynamic response of the synthesized plantwide control system to a throughput transition from the design throughput (F_{C2} = 630 kmol/h) to maximum throughput (F_{C2} = 970 kmol/h) is shown in Figure 15.4. The product impurity is tightly controlled and the transients in the process variables are smooth implying the suitability of the control structure for near optimal operation over the wide throughput range.

Table 15.2. Salient Controller tuning parameter for Ethyl Benzene process

Controlled Variable	K_C	τ_i (min)	Sensor Span
LVL_{rxr1}	5	250	0 – 100%
LVL_{rxr2}	5	250	0 – 100%
T_{rxr1}	4.8	25	0 – 400°C
T_{col1}	3.2	18.5	77 °C – 157 °C
T_{col2}	2	11	0 .0 – 244.7 °C
x_{Bz}^{D2}	0.3	100	0 – 0.0016
x_{DEB}^{D2}	0.8	88.5	0 0.002
B_2	0.2	1200	0 – 500 kmol/h

All level loops use K_C = 2 unless otherwise specified
Pressure/flow controllers tuned for tight control
All composition measurements use a deadtime of 5 minutes and a sampling time of 5 mins

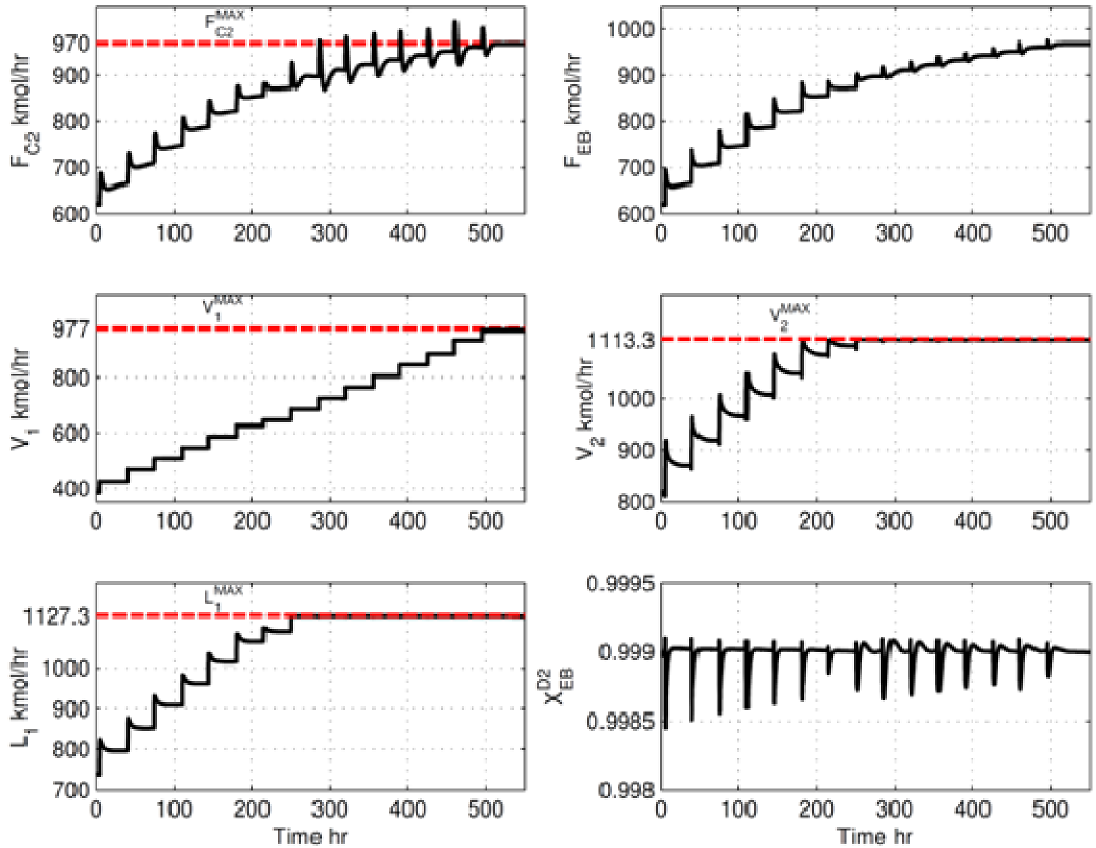

Figure 15.4. Low to maximum throughput transition of ethyl benzene process using modified economic plant-wide control structure