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Economic Plantwide Control Over a Wide Throughput Range: A Systematic Design Procedure

Rahul Jagtap and Nitin Kaistha
Dept. of Chemical Engineering, Indian Institute of Technology, Kanpur 208016, India

Sigurd Skogestad
Dept. of Chemical Engineering, Norwegian University of Science and Technology, Trondheim, Norway

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A systematic top-down procedure to design an effective plantwide control system for economically (near) optimum process operation over a wide throughput range is developed. The proposed procedure focuses on devising a control system for optimal operation at maximum throughput, where usually the highest number of constraints is active and the economic benefits of improved operation are the greatest. To do so, loops for tight control of all the active constraints and economically sound controlled variables (CVs) corresponding to unconstrained degrees of freedom are first implemented (Step 1) followed by a consistent inventory control system (Step 2). This control system is adapted with setpoints that become unconstrained at lower throughputs taking up additional economic CV control or throughput manipulation (Step 3). Iteration between the three steps may be necessary to eliminate fragile/unconventional inventory loops (Step 4). The application of the methodology is demonstrated on three realistic example processes. © 2013 American Institute of Chemical Engineers *AIChE J*, 59: 2407–2426, 2013
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Introduction

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 coworkers have done seminal work in highlighting key regulatory control issues such as the snowball effect[1] in reactor–separator recycle systems and suggesting practical control system structuring guidelines (Luyben's rules[2]) 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.[3] In their procedure, economic concerns are addressed indirectly in the form of requiring "tight" control of expected economic variables such as product impurity and process yield. The control objectives are obtained using engineering insights and heuristics.

Skogestad[4] 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 (DOFs), the control of self-optimizing controlled variables[5] (CVs) is recommended. By definition, when self-optimizing variables are held constant at appropriate values, near-optimal operation is achieved despite disturbances. The quest for the best self-optimizing CV set is, however, not always straightforward.

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 multiunit chemical process, there would clearly be several possible reasonable control configurations. An obvious question then is which one is the 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 in Refs. 6–8), although more recently, also at maximum throughput.[9–12] 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

Correspondence concerning this article should be addressed to N. Kaistha at nkaistha@iitk.ac.in.

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 application of the proposed approach is demonstrated on three realistic example processes. The approach has evolved out of very recent comprehensive case studies from our group.[13–15] Although 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 of general applicability to all continuous chemical processes. In particular, rational justification for the different sequential steps in the framework as well as for the key structural input–output (IO) pairing decisions (including throughput manipulation strategy) is provided. To the best of our knowledge, such a systematic and general economic plantwide control system design framework for achieving near-optimal process operation over a large throughput range encompassing multiple (hard) equipment capacity constraints is not forthcoming in the extant literature. That is the major research contribution of this work.

The novelty of the proposed approach, in contrast to the existing plantwide control literature, lies in designing a control system for the most constrained maximum throughput steady state and then adapting it for less constrained lower throughput operation. Extant approaches of Luyben[3] and Skogestad[4] devise the control system for the design throughput, where no equipment capacity constraints are active. Further, unlike the conventional bottom-up pairing approach, a top-down pairing approach with higher prioritization to economic objectives over regulatory objectives is used for realizing significantly improved economic operation.

In what follows, we develop the economic plantwide control system design procedure with particular emphasis on the conflicts that often arise between regulatory and economic control objectives and how they are resolved. It is the resolution of these conflicts that lies at the core of designing effective plantwide control systems. The application of the systematic procedure is then demonstrated on three example processes. Each example has been included to highlight the variety of conflicts between regulatory and economic objectives and their resolution within the proposed framework.

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 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 nonreactive 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 1 provides an illustration of the one-to-one correspondence between the independent CV setpoints (including TPM and excluding nonreactive material inventory controllers) and the steady-state DOF specification variable set for a simple reactor–recycle process with five steady-state operation DOFs. The five DOFs are related to one fresh feed, two reactor specifications (level and temperature), and two specifications for the column. Four alternative DOF specification sets are shown in Figure 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[16] for material inventory control that gives the orientation of the level controllers upstream and downstream of the TPM, respectively, opposite and in the direction of process flow. 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. Thus, there exists tremendous flexibility in designing the plantwide control system that must be gainfully exploited for achieving the twin objectives of robust stabilization and economic operation.

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/energy 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

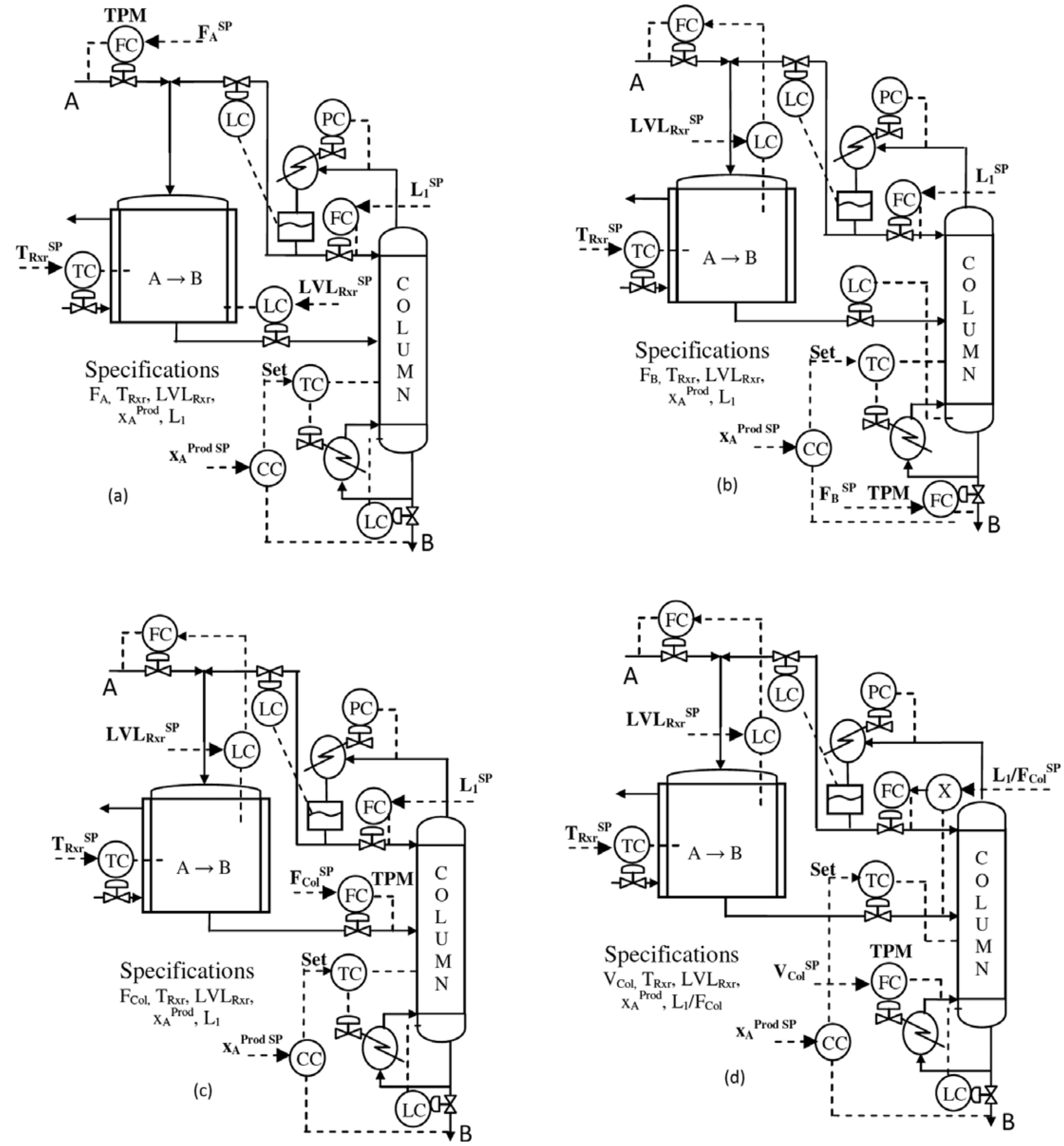

Figure 1. One-to-one correspondence between CV setpoints and steady-state specification variables for a simple recycle process.

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 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, because 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 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 nonlinear nature of the process with highly skewed deviations in only one direction.

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 toward

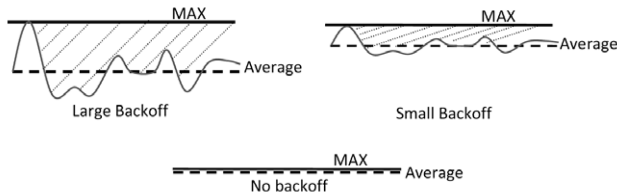

Figure 2. Illustration of tightness of active constraint control and back-off.

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/energy balance) control. It attempts to accomplish economic and regulatory control in a single layer. The same is made possible as controlling an economic CV usually 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 and 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 *a priori* 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 override controllers. The two-tiered framework must be appropriately modified to systematically devise such a control structure.

Control System Design for Optimal Operation over 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 due to 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 3 illustrates active constraint regions with respect to throughput for a process with five steady-state DOFs. The active constraints divide the entire throughput range into three regions corresponding to low (two active constraints), intermediate (three active constraints), and high throughputs (four active constraints). At the maximum achievable throughput (five 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).

Let us assume that the full active constraint set, corresponding to maximum throughput operation, does not change for a given process.* 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.

On the basis of the above-said arguments, the two-tier plantwide control system design framework is modified to design 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-optimal 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.

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

*This appears to be a reasonable assumption.

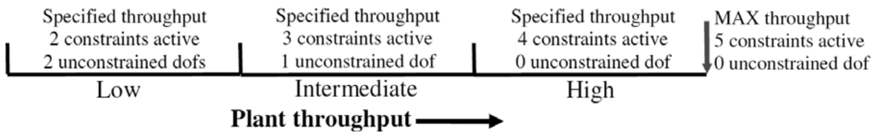

Figure 3. Active constraint regions with respect to throughput.

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.

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, and so forth would be active along with hard capacity constraints such as column operation at flooding limit and furnace operation at maximum duty. The full active constraint set may include direct Manipulated Variables (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, whereas 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 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. Similarly, sometimes tight control of an economic CV may require tight control of an inventory. A very common example is temperature-inferential quality control on a distillation column requiring tight column pressure control. The best pairings for both the economic CV and the associated inventory loop should then be implemented in this step itself. The remainder of the inventory control system is synthesized in the next step (Step 2).

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[17] 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 vapor), 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.,[2,3] 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 two 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, as 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 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 column internal flows. Conventional wisdom would suggest using such a leak stream for bottoms level control is unworkable and, therefore, ill-advised. However, if a stripping section tray temperature is well controlled, for example, by adjusting the boilup or feed, the seemingly unworkable pairing provides acceptable sump level control.[18] 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.

Step 3: design loops for "taking up" additional economic CV control at lower throughputs along with appropriate 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 the 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 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 noncritical 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, the 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.

Step 4: modify structure for better robustness/operator acceptability

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 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 Continuously Stirred Tank Reactor (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 Steps 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.

Case Studies

The application of the systematic approach for economic plantwide control system design is demonstrated on three 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 (EB) manufacturing process is the third and final example considered. To keep the discussion focussed on the IO pairing issue, optimization details that lead to the active constraint set and operating regions are deliberately omitted. These details are available in recent articles.[13–15]

Case study 1: recycle process with side reaction

The process flow sheet is shown in Figure 4 and consists of a cooled liquid-phase CSTR followed by a stripper and a distillation column. The main reaction A+B $\rightarrow$ C and the minor side reaction C+B $\rightarrow$ D occur in the CSTR. The unreacted A and B in the reactor effluent are stripped, condensed, and recycled along with some C. The stripper bottoms are fractionated to recover 99% pure C as the distillate (main product) and D with some C as the bottoms (side product). The process has seven steady-state DOFs (two fresh feeds, reactor level and temperature, one stripper DOF, and two column DOFs), and there are 13 independent control valves. Thus, even if all steady-state DOFs are exhausted at maximum throughput, six valves would still remain available for dynamic control, including inventory control. Table 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.

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

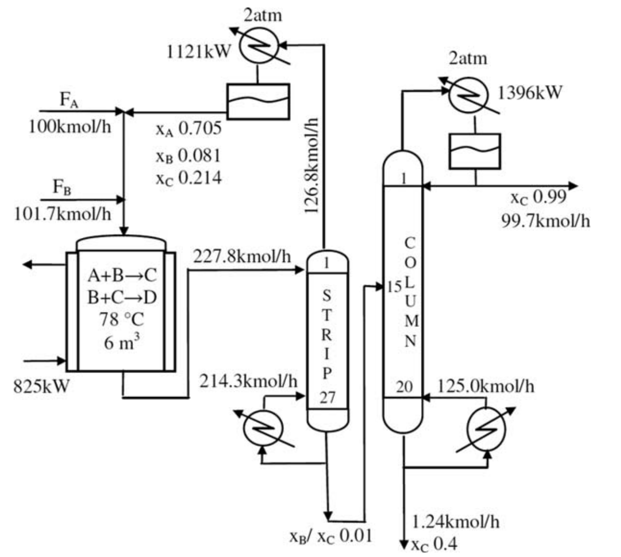

Figure 4. Schematic of recycle process with design and base operating conditions (Case Study I).

0.98% B $\left(x_B^{\mathrm{ColD}}\right)$ and 0.02% D $\left(x_D^{\mathrm{ColD}}\right)$ for the desired 99 mol % pure C $\left(x_C^{\mathrm{ColD}}\right)$ product. At maximum throughput, the active constraints are maximum column boilup (V_2^{MAX}), reactor temperature ($T_{\mathrm{Rxr}}^{\mathrm{MAX}}$), stripper boilup ($V_1^{\mathrm{MAX}}$), and reactor level ($\mathrm{LVL}_{\mathrm{Rxr}}^{\mathrm{MAX}}$). 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.† The four equipment capacity constraints, the two product impurity mol fractions, and the product column stripping section temperature specification exhaust all seven 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 $\mathrm{LVL}_{\mathrm{Rxr}}^{\mathrm{MAX}}$ 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}, $T_{\mathrm{Rxr}}^{\mathrm{MAX}}$, and V_1^{MAX}. The entire throughput range, thus, gets partitioned into three active constraint regions (see Table 1, Step 0). The number of unconstrained steady-state DOFs corresponding to the low-throughput (only $\mathrm{LVL}_{\mathrm{Rxr}}^{\mathrm{MAX}}$ active), intermediate-throughput ($\mathrm{LVL}_{\mathrm{Rxr}}^{\mathrm{MAX}}$ and V_1^{MAX} active), and high-throughput ($\mathrm{LVL}_{\mathrm{Rxr}}^{\mathrm{MAX}}$, V_1^{MAX}, and $T_{\mathrm{Rxr}}^{\mathrm{MAX}}$ 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 seven equality/inequality constraint variable values. Jagtap et al.[13] 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-optimizing CVs corresponding to the two unconstrained DOFs. In the intermediate-throughput region, holding x_B^{Rxr} constant ensures near optimum steady operation ($T_{\mathrm{Rxr}}^{\mathrm{SP}}$ 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.

†This ensures that *C* composition.

Table 1. Economic Plantwide Control Structure Synthesis for Recycle Process (Case Study I)

Step 0: active constraint regions and economic CV's				
Region	I	II	III	Max Throughput
Additional active constraints[a]	–	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	$T_{Rxr}^{MAX} \leftrightarrow O_{Rxr}$ $x_B^{ColD} \leftrightarrow T_{Stp}^{SP} \leftrightarrow F_{Stp}^{SP}$	$V_1^{MAX} \leftrightarrow Q_{Reb1}$ $x_D^{ColD} \leftrightarrow L_2/B_1^{SP} \leftrightarrow L_2^{SP}$	$V_2^{MAX} \leftrightarrow Q_{Reb2}$	$T_S^{Col} \leftrightarrow B_1$ $LVL_{Rxr}^{MAX} \leftrightarrow F_{Tot}^{Rxr} \leftrightarrow F_A$
Self-optimizing loops		None		
Step 2: maximum throughput inventory loops				
$LVL_{Reb2} \leftrightarrow B_1$ $LVL_{Reb1} \leftrightarrow x_B^{Rxr\,SP}\left(F_B/F_{Tot}^{Rxr}\right)^{SP} \leftrightarrow F_B$		$LVL_{Cnd1} \leftrightarrow F_{Rcy}$ $LVL_{Cnd2} \leftrightarrow D_2$		$P_{Cnd1} \leftrightarrow Q_{Cnd1}$ $P_{Cnd2} \leftrightarrow Q_{Cnd2}$
Step 3: additional self-optimizing CV loops at reduced throughput				
Region III TPM: V_2^{SP}		Region II TPM: V_2^{SP} $x_B^{Rxr\,SP\,b} \leftrightarrow (F_B/F_{Tot}^{Rxr})^{SP} \leftrightarrow F_B$ $LVL_{Reb1} \leftrightarrow T_{Rxr}^{SP} \leftrightarrow Q_{Rxr}$		Region I TPM: V_2^{SP} $T_{Rxr}^{SP\ b} \leftrightarrow Q_{Rxr}$ $x_B^{Rxr\,SP\,b} \leftrightarrow (F_B/F_{Tot}^{Rxr})^{SP} \leftrightarrow F_B$ $LVL_{Reb1} \leftrightarrow V_1^{SP} \leftrightarrow Q_{Reb1}$
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 TPM: $x_B^{Rxr\,SP}$		Region II TPM: T_{Rxr}^{SP} $x_B^{Rxr\,SP\,b}$		Region I TPM: V_1^{SP} $T_{Rxr}^{SP\ b}$ $x_B^{Rxr\,SP\,b}$

[a]LVL_{Rxr}^{MAX}, x_B^{ColD}, x_D^{ColD}, and T_S^{Col} are always active.
[b]Set point value is the optimized value.

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 from V_1^{MAX} 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.[‡] Maintaining product impurity B mol fraction (x_B^{ColD}) requires tight control of the B dropping down the stripper as all of it ends up in the product. As 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.

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

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, because 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 5. The constrained setpoints at maximum throughput are highlighted in brown.

Step 2: Inventory (Regulatory) Control System. Control loops to stabilize the liquid, vapor, 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 5. We need to control the

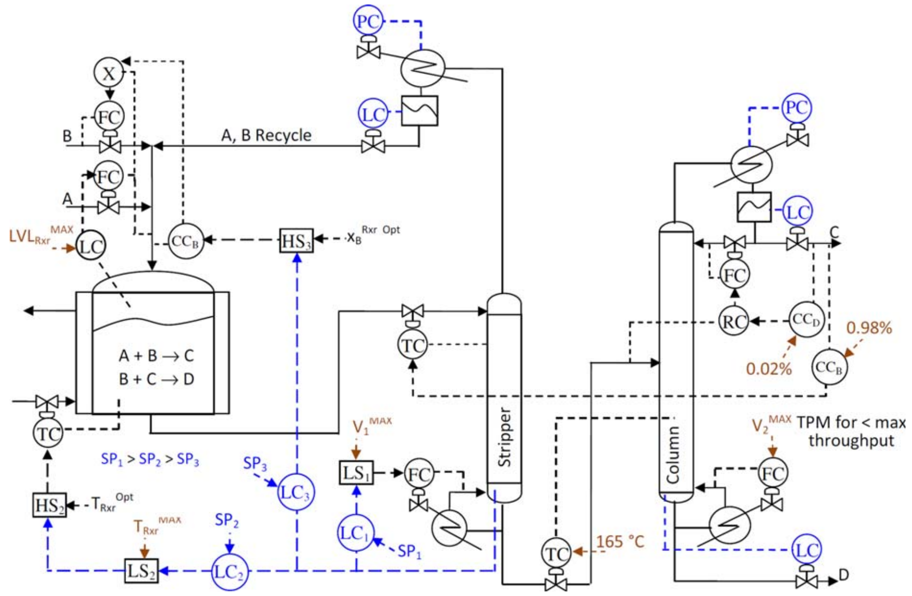

Figure 5. Plantwide control structure for maximum throughput operation of recycle process (Case Study I).

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

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 5 imply obvious loop pairings for inventory control. The column reflux drum level ($\mathrm{LVL_{Cnd2}}$) 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 ($\mathrm{LVL_{Bot}}$) is controlled using the feed from the stripper (B_1). To mitigate transients in the reactor composition, F_{B} is maintained in ratio with $F_{\mathrm{Tot}}^{\mathrm{Rxr}}$. 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 ($x_{\mathrm{B}}^{\mathrm{Rxr}}$) is maintained by adjusting the F_{B} to $F_{\mathrm{Tot}}^{\mathrm{Rxr}}$ ratio setpoint $\left(F_B/F_{\mathrm{Tot}}^{\mathrm{Rxr\,SP}}\right)$.

With these pairings, no close-by valves are left for controlling stripper sump level ($\mathrm{LVL_{Stp}}$). The only available option is to adjust the $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$. 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 $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$ 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 $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$ 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 $\pm 5\%$ step bias in the F_{B} sensor (control system tuning details in Appendix). 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, although unconventional, is quite robust and acceptable.

Step 3: Additional Unconstrained Economic CV Control Loops and Throughput Manipulation. At lower throughputs, the additional unconstrained economic CVs whose control must be taken up are $x_{\mathrm{B}}^{\mathrm{Rxr}}$ and T_{Rxr}. Both are associated with the reactor. As 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 $x_{\mathrm{B}}^{\mathrm{Rxr}}$ reducing. The excess A inside the reactor then increases to further suppress the side reaction for improved yield to the desired product. When $x_{\mathrm{B}}^{\mathrm{Rxr}}$ reduces to its optimal value, it must be held constant for optimal operation. $\mathrm{LVL_{Stp}}$ then gets controlled using $T_{\mathrm{Rxr}}^{\mathrm{SP}}$, *in lieu* of $x_{\mathrm{B}}^{\mathrm{Rxr}}$. $T_{\mathrm{Rxr}}^{\mathrm{SP}}$ would reduce below $T_{\mathrm{Rxr}}^{\mathrm{MAX}}$ as V_2^{SP} is decreased. When $T_{\mathrm{Rxr}}^{\mathrm{SP}}$ decreases to its optimum value, it must be held constant. $\mathrm{LVL_{Stp}}$ then gets controlled using V_1^{SP} *in lieu* of $T_{\mathrm{Rxr}}^{\mathrm{SP}}$. 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 $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$ to $T_{\mathrm{Rxr}}^{\mathrm{SP}}$ to V_1^{SP}, as throughput is reduced. Referring to the throughput regions in Table 1, at high throughputs, $x_{\mathrm{B}}^{\mathrm{Rxr}}$ floats to the appropriate value determined by V_2^{SP} via the action of the inventory control system. At intermediate throughputs, $x_{\mathrm{B}}^{\mathrm{Rxr}}$ is maintained at its optimum and T_{Rxr} floats to the appropriate value. Finally, at low throughputs, $x_{\mathrm{B}}^{\mathrm{Rxr}}$ 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 ($\mathrm{LC_1}$, $\mathrm{LC_2}$, and $\mathrm{LC_3}$) is shown in Figure 5. The MVs for $\mathrm{LC_1}$, $\mathrm{LC_2}$, and $\mathrm{LC_3}$ are, respectively, V_1^{SP}, $T_{\mathrm{Rxr}}^{\mathrm{SP}}$, and $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$. At maximum throughput, as $T_{\mathrm{Rxr}}^{\mathrm{MAX}}$ and V_1^{MAX} are active, $\mathrm{LC_1}$ and $\mathrm{LC_2}$ are inactive and sump level control is performed by $\mathrm{LC_3}$. As V_2^{SP} (TPM) is reduced below V_2^{MAX}, $\mathrm{LC_3}$ decreases $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$. When $x_{\mathrm{B}}^{\mathrm{Rxr\,SP}}$ reduces below its optimum value, the high select block, $\mathrm{HS_3}$, passes the optimum value to the $x_{\mathrm{B}}^{\mathrm{Rxr}}$ controller. $\mathrm{LC_3}$ then becomes inactive, and stripper sump level control is lost. The level then increases beyond $\mathrm{LC_2}$ setpoint, and the $\mathrm{LC_2}$ output starts to decrease. When the output decreases below $T_{\mathrm{Rxr}}^{\mathrm{MAX}}$, level control is taken over by $\mathrm{LC_2}$. When $T_{\mathrm{Rxr}}^{\mathrm{SP}}$ decreases below its optimum value, the high select block, $\mathrm{HS_2}$, passes the optimum value, and $\mathrm{LC_2}$

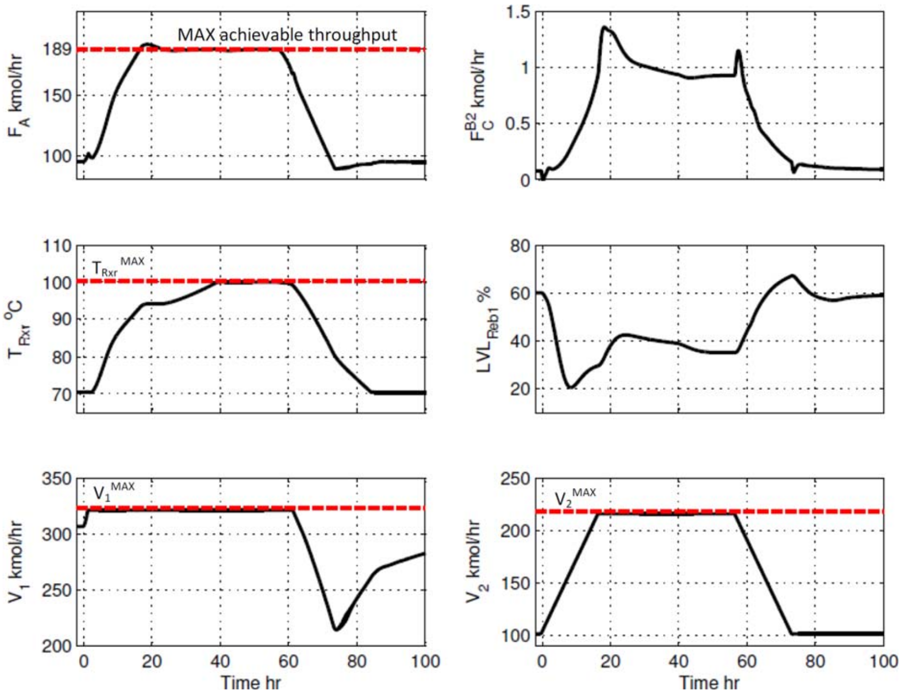

Figure 6. Throughput transition with stripper sump level override control scheme (Case Study I).

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

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.

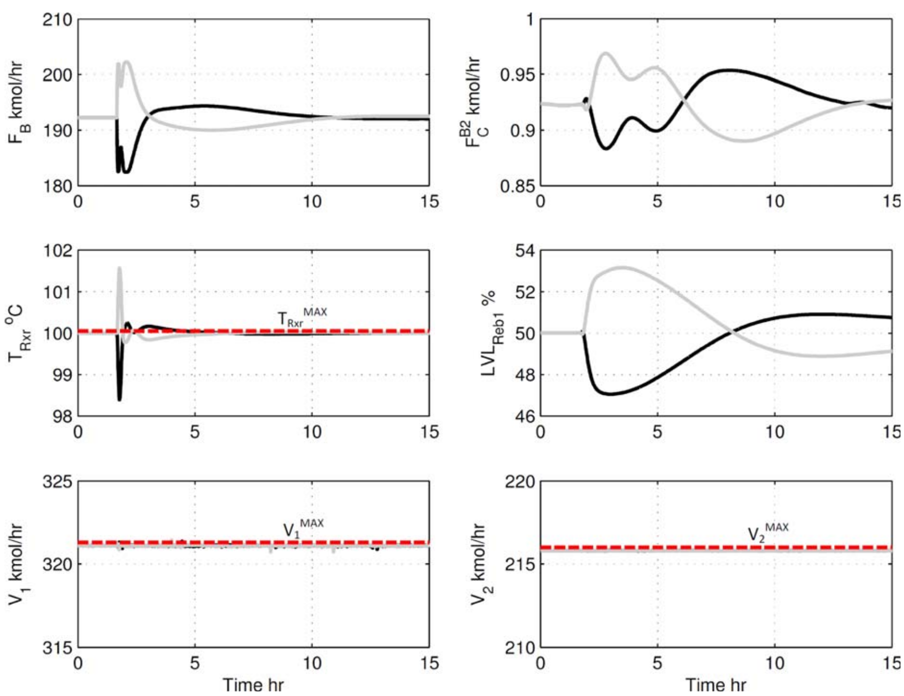

Figure 7. Transient response for ±5% step bias in F_B flow sensor (Case Study I). — +5% bias; —: −5% bias.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

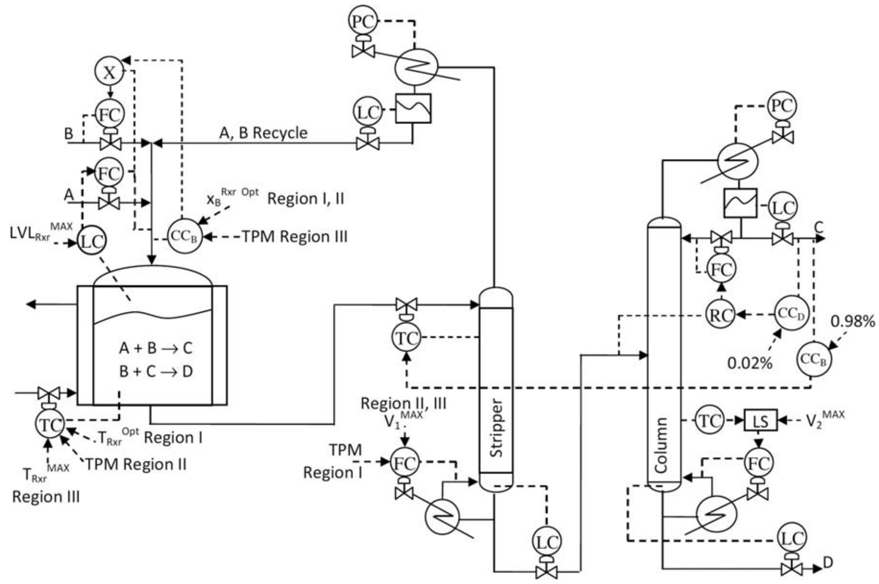

Figure 8. Recycle process (Case Study I) modified control structure for conventional inventory control system.

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 6 (salient loop tuning parameters in Appendix). The TPM V_1^{SP} is ramped up from its nominal value (101.4 kmol/h) to V_1^{MAX} (215.8 kmol/h) in 1000 min. For the ramp-down, the same ramp rate is used. The dynamic response shows that 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 7. 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%).[13] The synthesized plantwide control system, thus, achieves significantly superior economic operation for the same plant equipment.

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. However, it is instructive to develop a control system with conventional local inventory control loops.

The stripper sump level control loop in Figure 5 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. As 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 5), V_2^{SP} was used as the TPM in all regions. With the revised pairings where V_2^{SP}

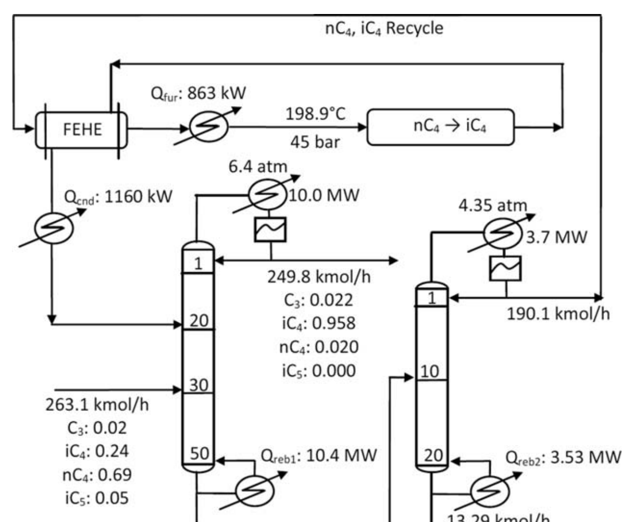

Figure 9. Schematic of C_4 isomerization process (Case Study II) with design and base operating conditions.

Table 2. Economic Plantwide Control Structure Synthesis for Isomerization Process (Case Study II)

Step 0: active constraint regions and economic CV's				
Region	I	II	III	Max Throughput
Additional active constraints[a]	–	$Q_{\text{fur}}^{\text{MAX}}$	V_1^{MAX} $Q_{\text{fur}}^{\text{MAX}}$	V_2^{MAX} V_1^{MAX} $Q_{\text{fur}}^{\text{MAX}}$
Unconstrained DOF's	2	1	0	0
Self-optimizing CV's	ΔT_{Col1}, V_2/B_1	V_2/B_1	–	–
Step 1: maximum throughput economic control loops				
Active constraint control loops	$Q_{\text{fur}}^{\text{MAX}} \leftrightarrow$ Fully open fuel valve	$V_2^{\text{MAX}} \leftrightarrow Q_{\text{Reb2}}$	$V_1^{\text{MAX}} \leftrightarrow Q_{\text{Reb1}}$	$T_{\text{Rxr}}^{\text{MAX}} \leftrightarrow D_2$ $T_{\text{Cnd}} \leftrightarrow F_{\text{cool}}$
	$P_{\text{Rxr}}^{\text{MAX}} \leftrightarrow \text{VLV}_2$	$x_{\text{nC4}}^{D1\text{MAX}} \leftrightarrow L_1/D_1^{\text{SP}} \leftrightarrow D_1$		$x_{\text{nC4}}^{B2} \leftrightarrow T_{\text{Col2}}^{\text{SP}} \leftrightarrow F_{\text{Col2}}^{\text{SP}}$
Self-optimizing loops		None		
Step 2: maximum throughput inventory loops				
$P_{\text{Cnd1}} \leftrightarrow Q_{\text{Cnd1}}$		$P_{\text{Cnd2}} \leftrightarrow Q_{\text{Cnd2}}$		$\text{LVL}_{\text{Cnd2}} \leftrightarrow L_2$
$\text{LVL}_{\text{Cnd1}} \leftrightarrow L_1$		$\text{LVL}_{\text{Reb1}} \leftrightarrow F_{\text{C4}}$		$\text{LVL}_{\text{Reb2}} \leftrightarrow B_2$
Step 3: additional loops at reduced throughput				
Region III		Region II		Region I
TPM: V_2/B_1^{SP}		TPM: $\Delta T_{\text{Col1}}^{\text{SP}}$		TPM: Q_{fur}
		V_2/B_1[b] $\leftrightarrow V_2 \leftrightarrow Q_{\text{Reb2}}$		ΔT_{Col1}[b] $\leftrightarrow V_1 \leftrightarrow Q_{\text{Reb1}}$
				V_2/B_1[b] $\leftrightarrow V_2 \leftrightarrow Q_{\text{Reb2}}$

[a]$T_{\text{Rxr}}^{\text{MAX}}$, $P_{\text{Rxr}}^{\text{MAX}}$, $x_{\text{nC4}}^{D1\text{MAX}}$, $T_{\text{Cnd}}=53°\text{C}$, and $x_{\text{nC4}}^{B2}=1\%$ are always active.
[b]Set point is the optimized value.

is used for T_S^{Col} control, an alternative throughput manipulation strategy is needed. To reduce throughput below maximum (Region III), $x_B^{\text{Rxr SP}}$ gets used as the TPM. Once $x_B^{\text{Rxr SP}}$ is reduced to its optimum value, the TPM shifts to $T_{\text{Rxr}}^{\text{SP}}$, which is reduced below $T_{\text{Rxr}}^{\text{MAX}}$ (Region II). Once $T_{\text{Rxr}}^{\text{SP}}$ 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 $T_{\text{Rxr}}^{\text{MAX}}$ is minimal, as $T_{\text{Rxr}}^{\text{SP}}$ is not adjusted by any master cascade loop once $T_{\text{Rxr}}^{\text{MAX}}$ is hit. The revised control system is shown in Figure 8 (Step 4 in Table 1).

Case study 2: C_4 isomerization process

Consider the isomerization process flow sheet shown in Figure 9 where the objective is to convert normal butane (n-C_4) to the more valuable product, *iso*-butane (*i*-C_4). The moderately exothermic irreversible isomerization reaction n-C_4 $\rightarrow$ *i*-C_4 (no side reactions) occurs in an adiabatic gas-phase tubular reactor. Fresh feed with a mix of mainly n-C_4 and *i*-C_4 with some C_3 and *i*-C_5 is fed to a deisobutanizer (DIB) column, which recovers *i*-C_4 product (along with feed C_3) at the top. The n-C_4 with some *i*-C_4 (and feed *i*-C_5) drops down the bottoms and is fed to the purge column, which recovers *i*-C_5 as a small bottoms stream with the n-C_4 and *i*-C_4 going up the top to the reactor after passing through a heating section consisting of a feed-effluent heat exchanger and a furnace. The hot reactor effluent is used to preheat the cold reactor feed. The cooler effluent is then condensed and fed to the DIB above the fresh C_4 feed (as it is richer in *i*-C_4). The step-by-step application of the economic plantwide control system design procedure is summarized in Table 2. Each of the steps is briefly explained in the following.

Step 0: Active Constraint Regions and Optimal Operation. The process has eight steady-state DOFs (one fresh feed, four column DOFs, one furnace DOF, one reactor effluent condenser DOF, and one reactor loop pressure). At maximum throughput, six constraints are active: maximum reactor temperature and pressure ($T_{\text{Rxr}}^{\text{MAX}}$ and $P_{\text{Rxr}}^{\text{MAX}}$), product impurity ($x_{\text{nC4}}^{D1\text{MAX}}$), furnace duty ($Q_{\text{fur}}^{\text{MAX}}$), purge column boilup ($V_2^{\text{MAX}}$), and DIB boilup ($V_1^{\text{MAX}}$). The former three are treated as soft constraints (small transient deviations beyond the limit permissible), whereas the latter three are hard constraints (deviations beyond limit not acceptable/possible). In addition, for economic reasons, the loss of precious n-C_4 in the bottoms of the purge column should be kept small. Accordingly, x_{nC4}^{B2} is maintained at a small value of 2%. Also, the reactor effluent condenser exit temperature (T_{Cnd}) is specified to be 53°C (set by cooling water). Then, this takes care of all the available steady-state DOFs at maximum throughput. Note that even as the process has eight steady-state DOFs, there are 14 independent control valves. The six additional control valves remain available for inventory control (stabilization), and we seek to devise an effective inventory management system along with tight control of the economic CVs.

In terms of economic sensitivity, the loss per unit back-off in the hard constraints decreases in order, $Q_{\text{fur}}^{\text{MAX}}$, V_1^{MAX}, and V_2^{MAX}. Also, $T_{\text{Rxr}}^{\text{MAX}}$, $P_{\text{Rxr}}^{\text{MAX}}$, and $x_{\text{nC4}}^{D1\text{MAX}}$ remain active constraints even at low throughputs with the first two constraints maximizing single pass conversion for reduced recycle/energy cost and the last constraint minimizing product give-away and DIB boilup. As throughput is decreased below maximum, V_1^{MAX}, V_2^{MAX}, and $Q_{\text{fur}}^{\text{MAX}}$ become optimally inactive. When V_2 becomes inactive, V_2/B_1 is held constant for mitigating over-refluxing in the purge column.[§] When V_1 becomes inactive, a sensitive difference between two stripping tray temperatures close to the fresh column feed ($\Delta T_{\text{Col1}}=T_{37}-T_{32}$) is held constant to mitigate over-refluxing in the DIB for energy efficiency. In other words, ΔT_{Col1} and

[§]Alternatively, an additional sensitive tray temperature may be held constant.

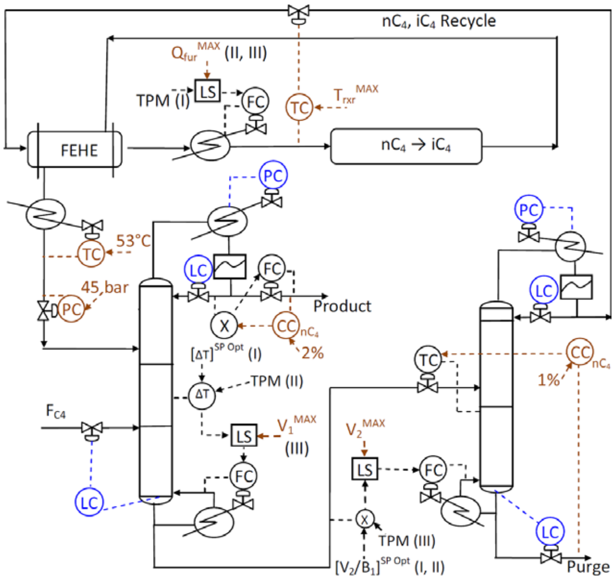

Figure 10. Economic plantwide control structure for isomerization process (Case Study II). Operating region shown in parentheses.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

V_2/B_1 are considered as self-optimizing CVs corresponding to unconstrained V_1 and unconstrained V_2, respectively.

Step 1: Loops for Tight Active Constraint Control at Maximum Throughput. To minimize the back-off in the hard active constraints at maximum throughput (Q_{fur}^{MAX}, V_2^{MAX}, and V_1^{MAX}), the furnace duty (Q_{fur}) valve is kept fully open, and V_2 and V_1 are controlled using the corresponding column reboiler steam valves. Now, T_{Rxr} and P_{Rxr} must always be controlled tightly around T_{Rxr}^{MAX} and P_{Rxr}^{MAX}, so that the reaction conversion is maintained at the maximum possible for minimum recycle cost. Q_{fur} cannot be used to control T_{Rxr}, as Q_{fur}^{MAX} is active. A possible close-by handle is the purge column distillate (D_2). The $T_{Rxr}-D_2$ open loop time constant is likely to be comparable to the conventional $T_{Rxr}-Q_{fur}$ pairing, so that the tightness of closed loop control is likely to be similar. P_{Rxr} is maintained at P_{Rxr}^{MAX} by manipulating the reactor effluent condenser outlet (F_{Cnd}). The pairing would achieve tight P_{Rxr} control.

To maintain x_{nC4}^{D1} at x_{nC4}^{D1MAX}, the column reflux-to-distillate ratio is adjusted. The conventional pairing gives tight x_{nC4}^{D1} control. To tightly control the n-C_4 loss down the purge column bottoms, x_{nC4}^{B2} is controlled by manipulating the setpoint of a sensitive purge column stripping section tray temperature (T_{Col2}). As V_2^{MAX} is active, the purge column feed (B_1) is used as the MV for controlling T_{Col2}. Lastly, the reactor effluent condenser duty (Q_{Cnd}) is manipulated to maintain the condenser outlet temperature at 53°C. Figure 10 shows all the active constraint control loops (in brown) for process operation at maximum throughput.

Step 2: Inventory (Regulatory) Control System. Now let us put in place the inventory control (stabilizing) loops. Four column levels and two column pressures need to be controlled. The DIB reflux drum level (LVL_{Cnd1}) is controlled using the reflux (L_1) as it is much larger than the distillate (D_1). The distillate is then maintained in ratio with the reflux with the ratio setpoint being adjusted by the x_{nC4}^{D1} controller. The reboiler steam and bottoms valves are already paired so that DIB sump level (LVL_{bot1}) is controlled using the fresh C_4 feed (F_{C4}). On the purge column, the existing pairings dictate that the sump level (LVL_{bot2}) be controlled using the bottoms stream (B_2) and the reflux drum level (LVL_{Cnd2}) be controlled using the reflux (L_2). The two column pressures (P_{Cnd1} and P_{Cnd2}) are controlled conventionally using their respective condenser duty valves (Q_{Cnd1} and Q_{Cnd2}). The inventory control loops for maximum throughput operation are shown in Figure 10 (in blue).

Step 3: Additional Unconstrained Economic CV Control Loops and Throughput Manipulation. At reduced throughputs, over-refluxing in the two columns must be mitigated for economic operation. If the purge or DIB column boilup, (V_2^{SP} or V_1^{SP}) were to be used for throughput manipulation over the entire throughput range, the furnace duty, Q_{fur}^{SP}, would have to be used as the manipulation handle for mitigating over-refluxing on one of the columns resulting in an undesirable long loop at lower throughputs. Clearly, available "local" MVs should be used for mitigating the over-refluxing. Accordingly, when inactive, V_2^{SP} takes up V_2/B_1 control and V_1^{SP} takes up ΔT_{Col1} control. To reduce throughput below maximum, V_2/B_1^{SP} is reduced to its optimum value. In response, V_2^{SP} would decrease below V_2^{MAX}. Then, ΔT_{Col1}^{SP} is increased to its optimum value causing V_1^{SP} to decrease below V_1^{MAX}. Finally, Q_{fur}^{SP} is reduced below Q_{fur}^{MAX} to further reduce the throughput. The TPM, thus, shifts from V_2/B_1^{SP} (or V_2^{SP}) to ΔT_{Col1}^{SP} (or V_1^{SP}) to Q_{fur}^{SP}. Conversely, to transition to maximum throughput, Q_{fur}^{SP} is ramped to Q_{fur}^{MAX} followed by decreasing ΔT_{Col1}^{SP} to drive V_1 to V_1^{MAX} and then increasing V_2/B_1^{SP} to drive V_2 to V_2^{MAX}. The throughput then floats at the maximum. The additional economic CV loops and TPM are indicated in Figure 10. Note that in this structure, the impurities (DIB bottoms i-C_4 and purge column distillate i-C_5) recirculating in the recycle, reduce when V_1 and V_2 are driven to maximum. This causes the recycle (D_2) flow to reduce. To keep T_{Rxr} at maximum, more fresh feed must be added. The throughput, thus, increases further.

The synthesized control structure is tested for a ±10% step change in the fresh feed n-C_4 mol fraction with a corresponding change in i-C_4 mol fraction. The tuning procedure and parameter details for the salient control loops are provided in Appendix. The plantwide response of important process variables is plotted in Figure 11 for operation at maximum throughput and around the base-case design throughput of F_{C4}=334 kmol/h. Tight product impurity control along with a smooth transient change in the different process flows is observed.

The control structure is also tested for a large ramped throughput change from Mode I to Mode II and back. Q_{fur}^{SP} is first ramped up to Q_{fur}^{MAX} from its nominal value (801 kW) till Q_{fur}^{MAX} (1294 kW) in 16 h. This is followed by ramping up ΔT_{Col1}^{SP} (from nominal value of 1.39–2.0°C in 8.4 h) and finally ramping up V_2/B_1^{SP} (from nominal 1.182 till 2.853 in 8.4 h). In response to these ramp-ups, V_2 saturates at V_2^{MAX} followed by V_1 saturating at V_1^{MAX}. After allowing the process to settle down at the maximum throughput steady state, the setpoints are ramped down (at ~75 h) at the same rates in reverse order. Figure 12 plots the response for important process variables to the throughput transition. Smooth transients in the process flows along with tight product impurity control are observed.

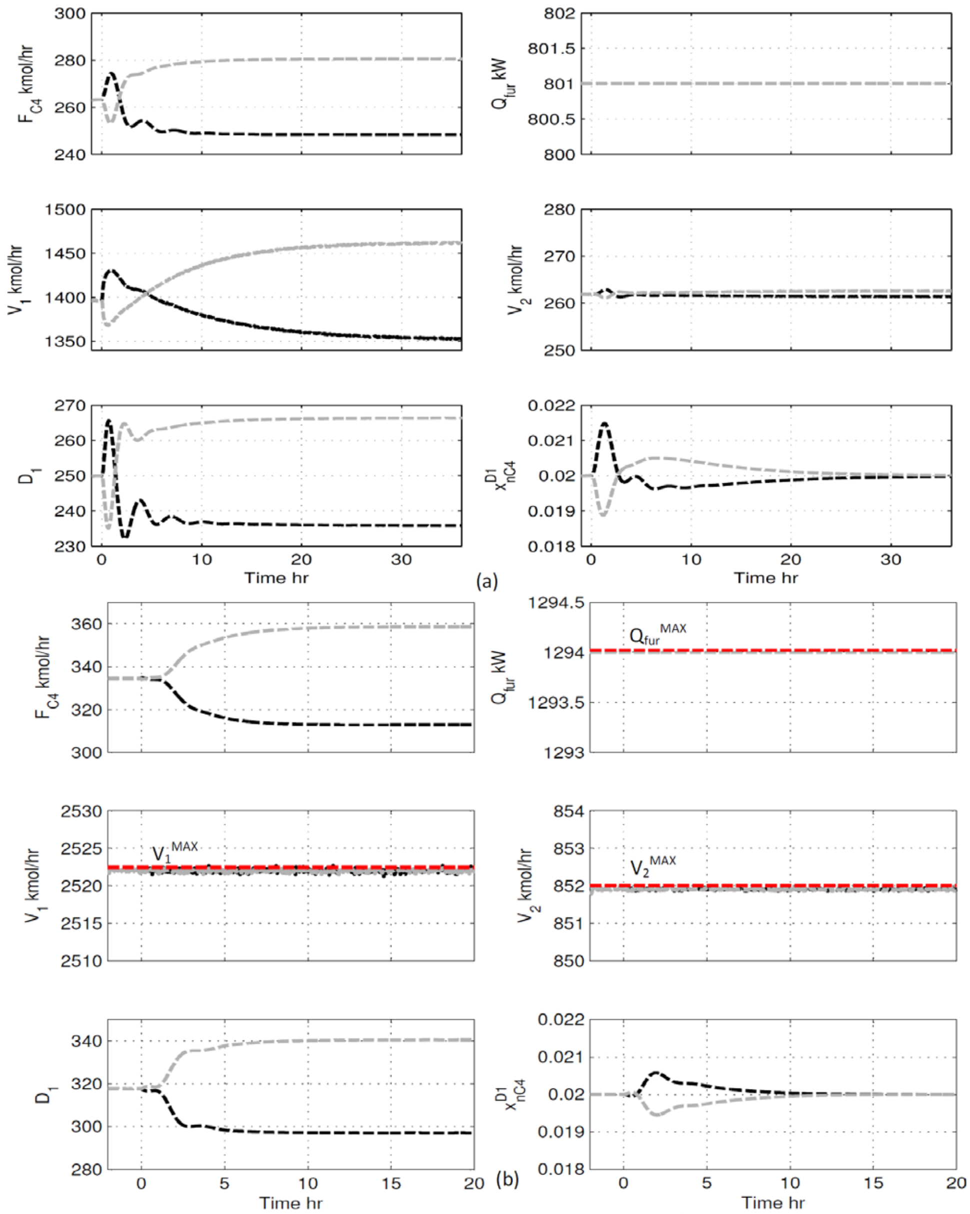

Figure 11. Plantwide transient response of isomerization process (Case Study II) using economic plantwide control structure at (a) design throughput (F_{C4}=263.1 kmol/h) and (b) maximum throughput. __: +5% step change in x_{nC4}^{feed}; _ _: −5% step change in x_{nC4}^{feed}.

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

The transient responses in Figures 11 and 12 demonstrate that at maximum throughput, all hard input constraints are maintained at their constraint limits with negligible back-off. Jagtap and Kaistha[14] show that with a conventional control structure with the fresh feed as the TPM over the entire throughput range, the back-off from V_2^{MAX}, V_1^{MAX}, and Q_{fur}^{MAX} results in ~2% reduction in maximum throughput. Lastly, it is noted that unlike the previous case study, there are no unconventional, long inventory control loops for this process. The application of Step 4 is, therefore, not necessary.

Case study 3: EB process

The process consists of two reactors and two columns along with two liquid recycle streams, as in Figure 13. The reaction chemistry consists of three reactions

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

The first two reactions occur primarily in the first coil cooled CSTR, whereas 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

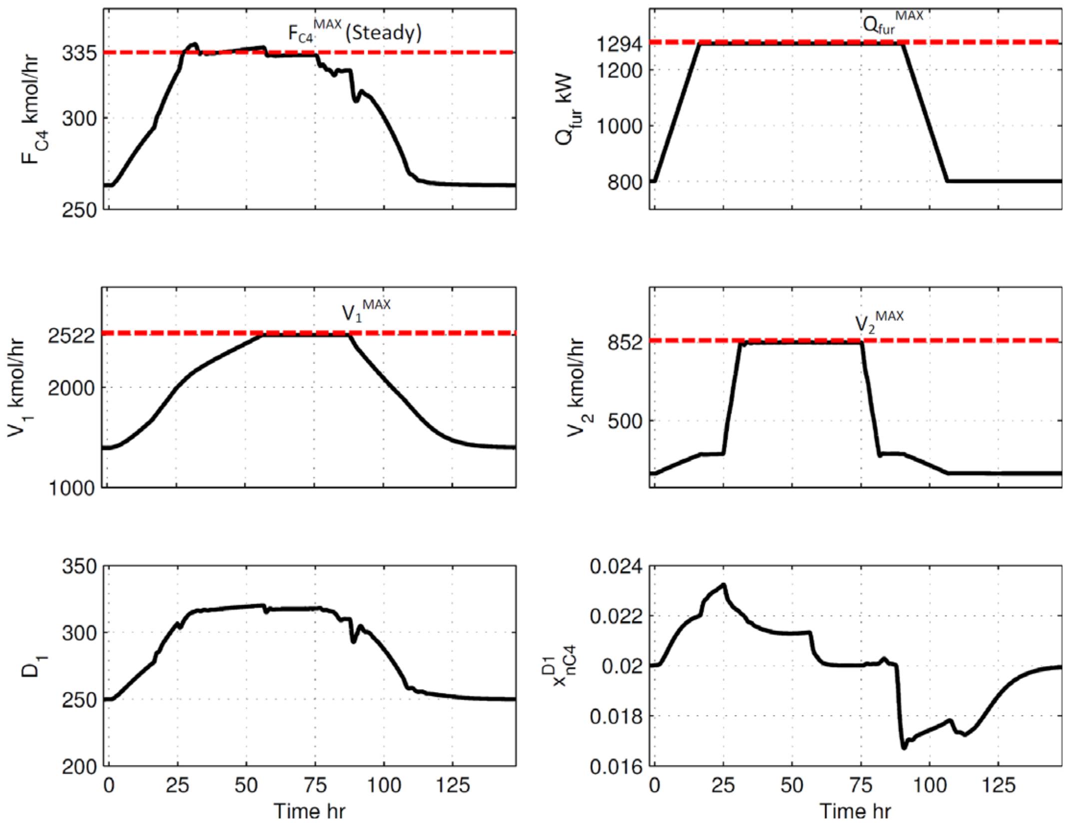

Figure 12. Low- to maximum-throughput transition of isomerization process (Case Study II).

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

fractionated in the product column to recover 99.9 mol % pure 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. The step-by-step synthesis of the economic plantwide control system is summarized in Table 3. The major steps are briefly described later.

Step 0: Active Constraint Regions and Optimal Operation. With fixed pressures, the process has nine steady-state DOFs: two fresh feeds, two DOFs for the first reactor (level and temperature), one for the second reactor (level), and four DOFs for the two columns. At maximum throughput, there are eight 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 ($T_{\mathrm{Rxr1}}^{\mathrm{MAX}}$) and level ($\mathrm{LVL}_{\mathrm{Rxr1}}^{\mathrm{MAX}}$), second reactor maximum level ($\mathrm{LVL}_{\mathrm{Rxr2}}^{\mathrm{MAX}}$) plus maximum product impurity levels $x_{\mathrm{Bz}}^{D2\mathrm{MAX}}$ (benzene mol fraction), and $x_{\mathrm{DEB}}^{D2\mathrm{MAX}}$ (DEB mol fraction) for no product give-away. This leaves one unconstrained steady-state DOF at maximum throughput, which is related to the optimal

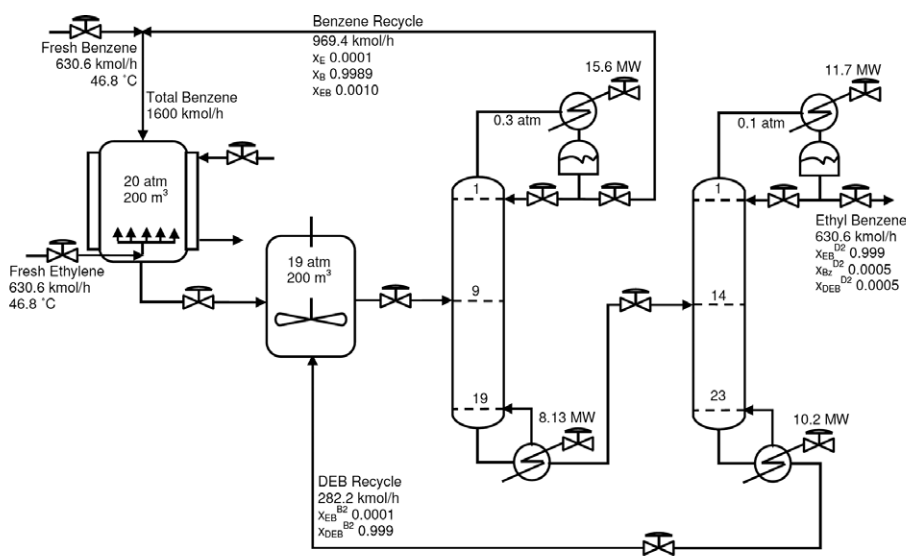

Figure 13. Schematic of EB process (Case Study III) with design and operating conditions.

Table 3. Economic Plantwide Control Structure Synthesis for EB Process (Case Study III)

Step 0: active constraint regions and economic CV's				
Region	I	II	III	Max Throughput
Additional active constraints[a]	–	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	$T_{Rxr1}^{MAX} \leftrightarrow Q_{Rxr1}$ $x_{Bz}^{D2} \leftrightarrow T_S^{Col1SP} \leftrightarrow T_S^{Col1SP} \leftrightarrow F_{Col1}^{SP}$	$V_1^{MAX} \leftrightarrow Q_{Reb1}$ $x_{DEB}^{D2} \leftrightarrow L_2/B_1^{SP} \leftrightarrow L_2^{SP}$	$V_2^{MAX} \leftrightarrow Q_{Reb2}$	$LVL_{Rxr1}^{MAX} \leftrightarrow F_{TotBz} \leftrightarrow F_{TotBz} \leftrightarrow F_{Bz}$ $LVL_{Rxr2}^{MAX} \leftrightarrow F_{Rxr2}$
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}		TPM: V_1^{SP}
		$T_S^{Col2} \leftrightarrow V_2^{SP\,b}$		$T_S^{Col2} \leftrightarrow V_2^{SP\,b}$
				$L_1/F_1 \leftrightarrow L_1{}^{b}$
Step 4: Modifications for conventional LVL_{Reb1} control loop				
$LVL_{Reb1} \leftrightarrow B_1$				
Region III		Region II		Region I
TPM: V_1^{SP}		TPM: V_1^{SP}		TPM: V_1^{SP}
$B_2 \leftrightarrow F_{TotBz}/F_{C2}^{SP}$		$T_S^{Col2} \leftrightarrow V_2^{SP\,b}$ $B_2 \leftrightarrow F_{TotBz}/F_{C2}^{SP\,b}$		$T_S^{Col2} \leftrightarrow V_2^{SP\,b}$ $B_2 \leftrightarrow F_{TotBz}/F_{C2}^{SP} \leftrightarrow F_{C2}$ $L_1/F_1 \leftrightarrow L_1{}^{b}$

[a] T_{Rxr1}^{MAX}, LVL_{Rxr1}^{MAX}, LVL_{Rxr2}^{MAX}, x_{Bz}^{D2MAX}, and x_{DEB}^{D2MAX} are always active.
[b] Unconstrained from MAX limit.

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, whereas 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.[15] The loss is deemed acceptable and is a consequence of energy being significantly cheaper than products or raw material (Douglas' doctrine[19]). At lower throughputs, over-refluxing 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 V_2 are L_1/F_{col1} and T_S^{Col2}, respectively.

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}^{D2MAX}, and x_{Bz}^{D2MAX}. 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}), whereas 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 the fresh benzene is fed as a make-up stream (Luybens' rule). Lastly, B_2 (self-optimizing CV) is flow controlled.

Step 2: Inventory (Regulatory) Control System. The remaining inventories to be controlled include the four column levels (LVL_{Cnd1}, LVL_{Cnd2}, LVL_{Bot1}, and LVL_{Bot2}) and the two column pressures (P_{Cnd1} and P_{Cnd2}). The column

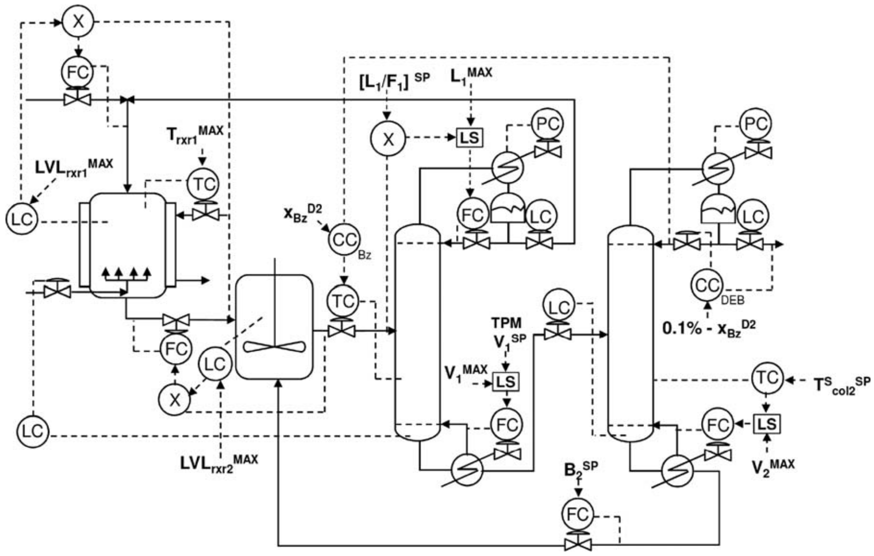

Figure 14. EB process (Case Study III) economic plantwide control structure (with long inventory loop).

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, as 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}/T^{SP}_{TotBz}. 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}/T^{SP}_{TotBz} by the level controller.

Step 3: Additional Unconstrained 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 over-refluxing in the recycle column.¶ Similarly, V_2^{SP} takes up tight control of a sensitive product column stripping tray temperature, whenever feasible at lower throughputs.

Step 4: Modifications for a More Conventional Inventory Control System. The economic plantwide control structure synthesized by the application of Steps 1–3 of our procedure is shown in Figure 14. 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 h, so that the dynamic response of LVL_{Bot2} to a change in F_{C2}/T^{SP}_{TotBz} (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}/T^{SP}_{TotBz}, which takes up "loose" control of the self-optimizing variable, B_2. The long inventory loop, $LVL_{Bot1}-F_{C2}/T^{SP}_{TotBz}$, in Figure 14 (Step 2 row in Table 3), thus, gets replaced by a long $B_2-F_{C2}/T^{SP}_{TotBz}$ loop after the re-pairing exercise to provide a conventional and robust inventory control system. The revised control system is shown in Figure 15.

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 over-refluxing 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 $L_1^{SP}<L_1^{MAX}$).

It is highlighted that in the revised pairings for more conventional inventory control (Step 4 in Table 3), B_2 must be controlled (by adjusting F_{C2}/T^{SP}_{TotBz}) 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}/T^{SP}_{TotBz} 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 that does not

¶Alternatively, $L1SP$ can take up rectifying temperature control for dual ended control.

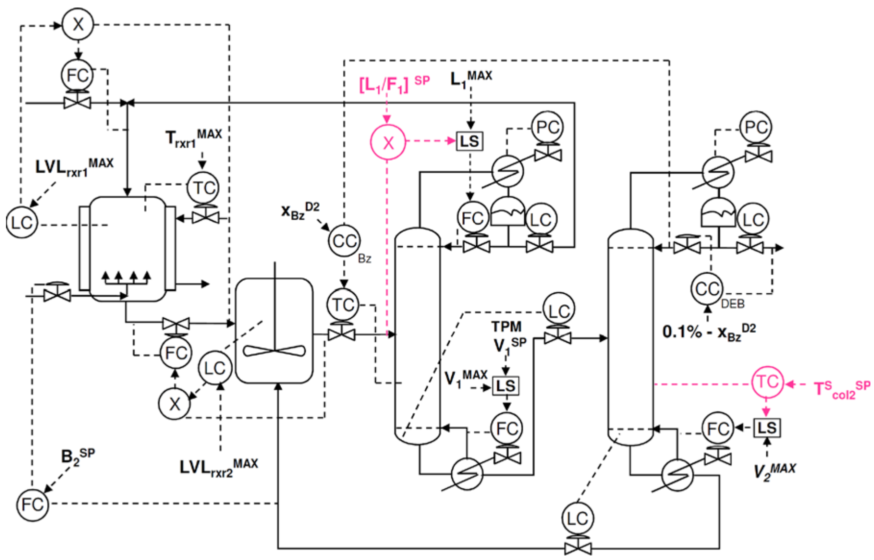

Figure 15. Modified economic plantwide control structure for EB process (Case Study III).

[Color figure can be viewed in the online issue, which is available at wileyonlinelibrary.com.]

suffer from such hidden instabilities due to slow accumulation of component inventories.

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.

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 16 (salient loop parameters in Appendix). The TPM V_1^{SP} is ramped between the nominal value (394 kmol/h) and V_1^{MAX} (977 kmol/h) at a ramp-rate of 40 kmol/h^2. The dynamic response shows that 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.

Discussion

In the proposed plantwide control system design procedure, loops are configured for tight control of all active constraints and self-optimizing CVs (if any). Tight control of the economically least significant active constraints is given up to develop a more conventional and robust inventory control system. In many situations, the last constraint to go active (bottleneck) is a direct MV (i.e., a hard constraint), is economically dominant (highest sensitivity), and its optimal variation with throughput is monotonic. The other active constraints are soft or with small/negligible economic

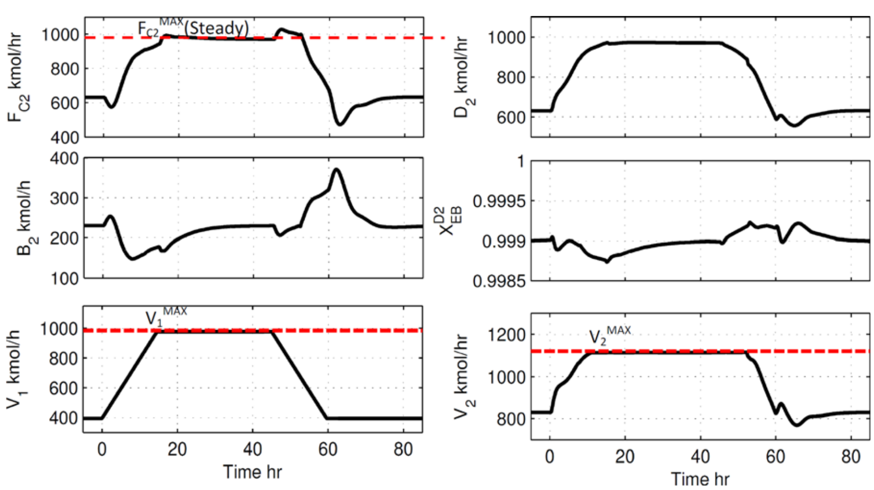

Figure 16. Low- to maximum-throughput transition of EB process (Case Study III) using modified economic plantwide control structure.

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sensitivity. The plantwide control system may then be synthesized by using the dominant hard active constraint variable as the TPM for the entire throughput range and configuring inventory controllers around it. This methodology, demonstrated in recent literature reports,[9,10,20] is then a special case of the more general approach demonstrated here.

In our procedure, process dynamics is considered only qualitatively (e.g., long loops) and not in a rigorous manner. Consequently, some of the IO pairing decisions may appear ad hoc requiring deeper investigation/analysis. In particular, the degree of robustness of the long inventory loops and the consequent need for revising the pairings may not be very clear. For such situations, candidate structures with revised pairings that eliminate some or all of these long loops can easily be obtained, the re-pairing exercise being only a thought-experiment. As the number of long inventory loops is usually very small (say, 2 or less), the number of candidate structures is then also small (say 2–5). These few candidate structures can easily be tested on a rigorous dynamic simulator (e.g., Aspen or Unisim) for a quantitative comparison of their regulatory and economic performance for various disturbance scenarios and a final recommendation on the structure that best addresses the conflicts between economic operation and robust process stabilization.

In some processes, the TPM location is not flexible and gets fixed by processing requirements such as on-demand process operation (TPM at product stream) or limited raw material feed (TPM at process feed). The methodology for control structure synthesis may still be applied to develop the best control strategy for managing the remaining control valves, excluding the TPM. In the resulting control system, the control of some of the economic CVs would usually be less tight. In particular, large back-off in an economically dominant hard constrained variable(s) due to transients can result in significant economic loss. For the three examples studied here, using a conventional control structure with TPM at one of the process feeds, the plant operating profit loss due to back-off in hard active constraints can be quite significant at a few percentage points ($\sim$7% for recycle process, $\sim$2% for isomerization process, and $\sim$2% for EB process). This only highlights that any flexibility in TPM location must be gainfully exploited for tight economic CV control.

There are practical considerations in process operation such as the impact of transients on equipment life or the inherently difficult dynamics/stabilizability of a process section, that are often quite difficult to quantitatively account for in an economic objective function. Often a particular process section is extremely expensive or is inherently difficult to stabilize, so that the plantwide control system must be designed to mitigate the transients propagated to this section. The TPM may then be chosen at the feed to this section, so that the transients are propagated away from the section via the outwardly radiating orientation of the inventory loops. The basic pairing philosophy developed here can then still be applied for prudent management of the remaining control valves.

We highlight that not using active constraint variables in the full constraint set for control results in IO pairings with critical variables being controlled by MVs that never saturate or hit a constraint. Their control is, thus, never lost. Also, consistent inventory control at maximum throughput ensures proper inventory balancing at lower throughputs too, that is, across the entire throughput range. By contrast, when the control system is designed for operation at the design steady state with only a partial active constraint set, the implemented IO pairings may result in loss of control of a critical CV such as product quality or worse, an inconsistency in the inventory balancing scheme when a constraint goes active causing snowballing in a recycle loop and consequent destabilization. Overrides must then be used to alter the control structure appropriately to ensure proper process regulation when a constraint is hit. These overrides usually significantly degrade the dynamic performance of the control system due to additional transients caused by "taking-up" and "giving-up" of control by the overrides.[14,15] Designing the control system for the full active constraint set (maximum throughput) is then the most direct way of obtaining a simple control system with the fewest (possibly none) overrides for handling constraints. Such simple control systems are likely to be dynamically superior at high-throughput operation, where the overrides get triggered and also be more acceptable to the operators.

If we look at the evolution of the plantwide control system design philosophy, the earliest systematic approach of Page Buckley[21] consisted of choosing a TPM (usually at a fresh feed) and designing robust inventory (material balance) control system pairings followed by selection of economic loop pairings, that is, a bottom-up approach. The systematic approach of Luyben et al.[3] represented the first significant departure with loops for crucial objectives such as product quality, safety, and environmental variables being implemented first followed by the inventory control loops. Before implementing these loops, the TPM is first chosen which fixes the overall inventory control system orientation. The approach is, thus, "bottom-up top-down." The approach proposed here is a natural extension to a truly top-down method that systematizes the handling of all active constraints and the best throughput manipulation strategy in light of the same.

Conclusions

In conclusion, this article has developed a systematic approach for top-down design of a plantwide control system for economically optimal operation over a wide throughput range, where the active constraint set expands with throughput. The three example case studies demonstrate the application of the methodology for synthesizing an effective control system with a smooth transition between adjacent active constraint regions. In particular, unconventional inventory control loops that are nonlocal and long can provide acceptable inventory control (stabilization) with tighter control of the economic CVs. Further, by appropriate shifting of the TPM in the different operating regions, a simple plantwide control system with the tightest possible control of the economic CVs over the entire throughput range can be synthesized.

Acknowledgments

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Appendix

Rigorous dynamic simulations are performed to test the synthesized control structures in the example case studies. Hysys is used as the simulation platform for Case Study I and Case Study II, whereas Case Study III is performed in Aspen Plus. The tuning procedure used to tune to the different loops in the case studies along with the salient tuning parameters so obtained are summarized in this Appendix.

Unless specified otherwise, all flow/pressure PI controllers are tuned tight for a fast and snappy servo response. The long B_2 loop in Case Study III is tuned by hit-and-trial for a smooth overall plantwide response. In all case studies, the nonreactive level controllers are P-only with a gain of 2. The only exception is the Case Study I unconventional stripper sump level controller with overrides. For the three different pairings in the three operating regions (see Figure 5), distinct conservative (nonaggressive) tunings are used to dampen flow variability. All reactive levels (i.e., CSTR levels) are controlled using a PI controller for offset free level tracking. In Case Study I, 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. In Case Study III, the Aspen dynamics 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 min 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 min each is applied to all composition measurements. The tuning parameters of salient loops for each of the case studies are reported in Tables A1.

Table A1. Salient Controller Tuning Parameters for Case Study I

CV	Sensor Span	MV	MV Span	K_C	τ_i (min)
x_B^{Rxr}	0–1	$(F_B/F_{Tot}^{Rxr})^{SP}$	0–5	0.8	400
T_{Rxr}[a]	60–130°C	Q_{Rxr}	−50–630 kW	1	10
LVL_{Rxr}	0–100%	F_{Tot}^{Rxr}	0–800 kmol/h	0.5	25
T_{Stp}	100–160°C	F_{Stp}^{SP}	0–800 kmol/h	0.5	15
T_S^{Col}	140–180°C	B_1	0–250 kmol/h	0.6	25
x_B^{ColD}	0–0.02	T_{Stp}^{SP}	100–160°C	0.1	40
x_D^{ColD}	0–0.0004	L_2/B_1^{SP}	0–5	0.1	30
Tuning for LVL_{Reb1} override control					
LVL_{Reb1}^1	0–100%	V_1	0–650 kmol/h	0.8	200
LVL_{Reb1}^2	0–100%	T_{Rxr}	60–130°C	0.6	250
LVL_{Reb1}^3	0–100%	x_B^{Rxr}	0–1	0.5	400

[a]Derivative action used with $\tau_D = 2$ min.

Table A2. Salient Controller Tuning Parameter for Case Study II

CV	Sensor Span	MV	MV Span	K_C	τ_i (min)	τ_d (min)
T_{Rxr}	160–240°C	D_2	0–400 kmol/h	2	1	0.1
T_{Cnd}	40–60°C	Q_{Cnd}	0–3333 kW	0.3	20	2
T_{Col2}	40–80°C	B_1	0–600 kmol/h	0.2	15	1.5
x_{nC4}^{D1}	0–0.04	$[D_1/L_1]^{SP}$	0–20	0.2	120	–
x_{nC4}^{B2}	0–0.03	T_{Col2}^{SP}	40–80°C	0.08	300	–
ΔT_{DIB}	0–0.03	V_1	0–2600 kmol/h	0.5	40	–

Table A3. Salient Controller Tuning Parameter for Case Study III

CV	Sensor Span	MV	MV Span	K_C	τ_i (min)
LVL_{Rxr1}	0–100%	F_{Rxr2}/F_{TotBz}	0–2	5	30
LVL_{Rxr2}	0–100%	F_{col1}/F_{Rxr2}	0–1.5	5	30
T_{Rxr1}	0–400°C	Q_{Rxr1}	0–22 MW	4.0	25
T_{Col1}	77–157°C	F_{Col1}	0–2500 kmol/h	0.8	18.5
T_{Col2}	0.0–244.7°C	V_2	0–1113.3 kmol/h	4.4	11
x_{Bz}^{D2}	0–0.0016	T_{Col1}^{SP}	77–157°C	0.08	100
x_{DEB}^{D2}	0 0.002	L_2	0–83251 kg/h	0.32	88.5
B_2	0–500 kmol/h	F_{C2}	0–1500 kmol/h	0.4	400

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